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Non-energy benefits and profitability of energy efficiency measures

An evaluation of retrofit proposals of heat exchanger networks in a refinery

Master's thesis in Sustainable Energy Systems

LINNÉA SVENSSON & TOVE UDD

Department of Space, Earth and Environment
CHALMERS UNIVERSITY OF TECHNOLOGY
Gothenburg, Sweden 2018

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Master's Thesis 2018
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Cover: Panoramic photo of Preemraff Lysekil, by Preem AB

Typeset in LaTeX
Printed by Chalmers Reproservice
Gothenburg, Sweden 2018

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Abstract

The aim of this thesis was to evaluate different retrofit proposals for heat exchanger networks at Preem's refinery in Lysekil. The proposals were designed in previous work with the goal to bring both higher energy efficiency and other implications to the refinery processes. Four different proposals were selected and evaluated based on their profitability and other positive effects they had for the process - so called non-energy benefits (NEBs).

The methodology was to first estimate investment costs and utility cost savings of the retrofit proposals. Then the NEBs of the proposals were identified, and thereafter the economic value of selected NEBs (reduced CO_2 emissions and increased production) were estimated. Finally the profitability of the retrofits, were calculated and compared, with and without the economic value of NEBs.

The results show that in three out of the four evaluated retrofit proposals, the estimated monetary value of the NEBs is larger than the value of the energy savings. These proposals all included increased productivity as a non-energy benefit, and inclusion of NEBs, decreased the payback period by more than 50% for those three proposals. Several other NEBs than the quantified ones were also identified for each retrofit proposal, and even though these have not been assigned a monetary value, they can still increase the strategic value of the proposed energy efficiency measures.

Non-energy benefits are generally not included in profitability calculations when evaluating energy efficiency investments. The results of this thesis show that including NEBs could make a great impact on the evaluation, as they increase both the profitability and the strategic value of the investments.

Keywords: Energy efficiency, non-energy benefits, process integration, oil refinery, retrofit

Acknowledgements

First, we would like to thank our supervisors Sofie Marton and Elin Svensson. Sofie mostly for her guidance and feedback during this thesis, but also for the nice plants, and Elin for her valuable feedback and support. We would also like to thank Preemraff Lysekil, especially Jörgen Bengtsson, for the helpful input regarding our project. Last we would like to thank the Division of Energy Technology at Chalmers for giving us the possibility to write our thesis at their division.

Linnéa Svensson & Tove Udd, Gothenburg, June 2018

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List of Abbreviations

ΔT_{lm}	Logarithmic mean temperature difference [K]
Δp	Pressure drop [bar or kPa]
$\dot{q}$	Volumetric flow rate [m^3/s]
ϵ	Pump efficiency
η	Efficiency
w_s	Shaft power [kW]
A	Area [m^2]
C	Cost [€]
c_{fuel}	Specific CO_2 emissions from fuel combustion [kg/MWh]
CCR	Continuous catalyst regeneration
CDU	Crude distillation unit
CEPCI	Chemical engineering plant cost index
CO_2-eq	Carbon dioxide equivalents
CRU	Catalytic reforming unit
CS	Carbon steel
E	Electricity demand [MWh/yr]
e_{CO_2}	Emissions of CO_2 [kg/yr]
EBITDA	Earnings before interest, taxation, depreciation, and amortisation
f	Cost factors
GHG	Green house gases
HAGO	Heavy atmospheric gas oil
HEN	Heat exchanger network
HFO	Heavy fuel oil
HHAGO	Heavy heavy atmospheric gas oil
HP	High pressure
HX	Heat exchanger
ICR	Iso cracker unit
LNG	Liquefied natural gas
LP	Low pressure
LPG	Liquid petroleum gas
MHC	Mild hydro cracker unit
MP	Medium pressure
NEB	Non-energy benefit
NHTU	Naphtha hydro treatment unit
$P_{electricity}$	Price of electricity [€/MWh]
P_{fuel}	Price of fuel [€/MWh]
PBP	Payback period [yr]
Q	Heat load [W]

Q_{fuel}	Fuel demand [MWh/yr][1]
Q_{steam}	Steam demand [MWh/yr]
RP	Retrofit proposal
SS	Stainless steel
T	Temperature [K]
t	Annual operating time [h/yr]
U	Overall heat transfer coefficient [W/m^{2}K]
UCO	Unconverted oil
VBU	Visbreaker unit
VDU	Vacuum distillation unit
VGO	Vacuum gas oil
VHP	Very high pressure

[1] All energy values are based on LHV of fuel

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1. Introduction

In 2015, the energy use in Swedish industry was 140 TWh, making industry one of the three sectors with the largest energy use in Sweden (Energimyndigheten, 2017a). The industry sector emits 16.9 Mt CO_2-eq which is 32% (2016) of Sweden's total amount of fossil greenhouse gas emissions (Naturvårdsverket, 2017a). In the industry sector, refineries stands for one fifth of that amount, corresponding to 2.7 Mt CO_2-eq per year.

Preem's refinery in Lysekil, Preemraff Lysekil, is the largest refinery in Scandinavia with a capacity to refine 11.4 Mt crude oil per year (Preem AB, 2018). It is also one of the largest individual energy users and point sources of emissions of fossil CO_2 in Sweden (Naturvårdsverket, 2017b). Preemraff Lysekil is one of the more energy efficient refineries in Europe, both due to a continuous work with energy efficiency measures, and to their possibility to supply Lysekil's municipality with 50 GWh of district heating, by excess heat from the process (Preem AB, 2018). There are still possibilities to improve the energy efficiency even more, which makes Preem's refinery in Lysekil an interesting object to investigate in further detail. By including impacts from possible non-energy benefits (further described in Section 1.1), investment can be easier to motivate and the energy efficiency can be pushed even further.

1.1 Driving forces for and effects of energy efficiency

The main reasons to increase energy efficiency in a process are to save money and fuel and to decrease emissions. Some of the driving forces for change towards new or more efficient processes in refineries may be: rising fuel costs, lack of feedstock supply and environmental regulations (Chew et al., 2013). The threat of global warming, caused by emissions of greenhouse gases (GHG), is causing policy-makers and consumers to demand more sustainable products and production processes. An improved public image is an important matter for companies, and this is also a motivation to using energy more efficiently.

One way to make processes more energy efficient is to recover more process heat and increase the internal heat integration in the process, by retrofitting the heat exchanger network (HEN) (Marton et al., 2016). However, heat integration comes with some implications to the process, such as potentially high capital costs, disruptions in the process operations and negative (or positive) effects on operability.

When process integration is increased, different parts of the process can become more interconnected and thus more dependent on each other (Marton et al., 2016). Consequently, this can affect the operability of the process. There are several different definitions of operability, and in this thesis the definition adopted is the following:

"Operability is the ability to operate equipment, process units and total sites at different external conditions and operating conditions, without negatively affecting safety or product quality and quantity. This includes both steady-state and dynamic aspects of operation."

This is the definition proposed by Marton et al. (2016) after a literature study about the concept. Operability can be divided into the subcategories: flexibility, controllability, startup/shutdown, reliability/availability and practical considerations.

Since heat integration in a process affects several parts of the system, it can be hard to take all effects into account. When energy is saved in one part of the system, it might not lead to a decrease in the total energy demand (Marton et al., 2017). For example there can be a surplus of steam produced in the process which means that even if less steam is needed in one part of the process, the total energy demand will be the same.

Today it can be hard to motivate energy efficiency measures, as they often are of low priority by companies (Pye & McKane, 2000). If there are other investments that can be made that lead to higher profit, are necessary to fulfill new legal requirements or have a higher and more clear strategic value, energy efficiency is often not prioritised. One conventional way to estimate the feasibility of energy efficiency investments is to calculate the payback period (PBP) of the investment (i.e. how long time it takes before the savings of energy/fuel is equal to the invested money). Industries often demand a quite short PBP, below around 3 to 5 years, on investments that improve energy efficiency, which can sometimes be difficult to achieve. To give shorter PBPs, increase the profitability and add more strategic value to energy efficiency measures, NEBs could be included in the proposals to make the measures more interesting.

Non-energy benefits, also called co-benefits, are positive effects from energy efficiency improvements other than savings of energy and energy costs (Nehler, 2016). Non-energy benefits can exist on different levels (individual, national and international) (Nehler, 2016) and according to IEA (2014) the possible benefits from increased energy efficiency are many. Energy efficiency measures could affect both macroeconomics, the local and global environment, public health, the energy system and industrial productivity. However, in this thesis the focus of non-energy-benefits are only on the ones that concern industrial productivity.

Non-energy benefits within the industrial productivity category can be divided into different types of benefits, concerning for example production, operability, maintenance, work environment, waste and emissions (Nehler, 2016). Some examples of non-energy benefits at Preemraff Lysekil could be: reduced need for maintenance, reduced emissions and increased yield or product quality, by for example removing a bottleneck in the process (Marton & Svensson, 2016). A bottleneck is something that restrains the production, for example a limited capacity in a furnace.

1.2 Previous work

Chalmers University of Technology has had a research collaboration with Preem since 2008, where some of the subjects are: renewable raw materials, energy efficiency in the refineries and process development. The background to this thesis is the collaboration about energy efficiency, where pinch analysis has been performed with data collected from Preemraff in Lysekil. Pinch analysis is a tool to establish the minimum heat demand in a HEN, and this can be used to establish targets for heat recovery. The results from these analyses are presented in this section.

From the energy inventory and pinch analysis of the refinery, potential energy savings were calculated (Andersson et al., 2013). All process data was collected in 2010, and the analysis was made in 2012. The total net heat demand for the whole process at that time was 409 MW. The theoretical minimum heat demand was estimated to 199 MW, which would mean a potential saving of 210 MW. This is only a theoretical number, assuming that any process streams in the entire refinery can be heat exchanged, independent of process conditions and geographical location. The full potential energy saving is thus not possible to achieve in practice. The hinders are for example practical reasons such as too long distances between units or streams in the process, safety issues and other process constraints. Despite these practical constraints, the large theoretical potential indicates that there is a lot of energy to be saved.

The minimum heat demand for each of the refinery's 18 process units was also calculated, and after taking into consideration process constraints, the potential energy savings for each process unit were calculated. The result showed that the five most heat demanding process units at the refinery were: CDU + VDU (crude distillation and vacuum distillation), NHTU (naphtha hydrotreating), CRU (catalytic reforming), MHC (mild hydrocracker) and ICR (iso cracker) (Åsblad et al., 2014). Further descriptions for these units are presented in Section 2.1 and a schematic overview of the refinery and all the process units can be seen in Figure 2.1. According to Åsblad et al. (2014) the aforementioned process units account for approximately 90% of the added external heat at the refinery. Since CDU was rebuilt after the analysis was made, CRU, NHTU, MHC and ICR are the four process units that are considered in this thesis. The results for these four process units are presented in Table 1.1.

Table 1.1: Results from pinch analysis on four process units in the refinery (Åsblad et al., 2014).

	Present heat demand [MW]	Minimum heat demand [MW]	Energy savings potential [MW]
NHTU	45.4	40.4	5.0
CRU	78.4	61.5	16.9
MHC	36.7	11.4	25.3
ICR	45.6	5.8	39.8

The four process units mentioned in Table 1.1 (NHTU, CRU, MHC and ICR) were evaluated further in an interview study with plant engineers at Preemraff (Marton & Svensson, 2016). Different retrofit proposals considering these units were designed to see how increased heat integration affects the operability of the process. During the interview study, potential non-energy benefits of the retrofit proposals were also discussed.

Some of the retrofit proposals that seemed to involve possible non-energy benefits are now to be evaluated further. Non-energy benefits can, as mentioned earlier, be of use when trying to motivate energy efficiency investments, to give the investments higher values than when only taking energy savings into account.

So far in the work with the retrofit proposals, no estimations of monetary values have been made. As some of the savings in energy affects a complex steam network, the exact effects of some retrofit proposals are hard to calculate, therefore a model of the steam network was developed by Marton et al. (2017). This model can be used in an economic evaluation of the retrofit proposals that affect the steam production. By comparing the value of the energy savings to the value of the non-energy benefits, the analysis can show how big impact the inclusion of non-energy benefits has on making the energy efficiency measures cost efficient.

1.3 Aim and objectives

The aim of the thesis was to evaluate different retrofit proposals for heat exchanger networks (HENs) in chosen process units at Preem's refinery in Lysekil. The following objectives were included:

- Select appropriate HEN retrofit proposals to investigate in the thesis
- Identify the need for new equipment, changes in the refinery's energy balances and possible non-energy benefits, for the selected HEN retrofit proposals
- Assess the impact of implementation of the retrofit proposal with respect to:
  - Investment costs and changes in operating cost
  - Changes in the refinery's CO_2 emissions
  - Economic value of identified non-energy benefits

1.4 Limitations

In this thesis, only the earlier studied process units (NHTU, CRU, MHC, ICR) in Preem's refinery Preemraff Lysekil are considered. The retrofit proposals that are evaluated were selected from earlier studies by Marton and Svensson (2016). The analysis is based on stream data from previous studies, no other operating points are taken into account. The evaluation only considers increases in energy efficiency, savings of energy costs, investment costs for new/enlarged equipment, changes in CO_2 emissions and other non-energy benefits concerning industrial productivity.

2. Process description of Preemraff Lysekil

Preem's refinery in Lysekil, Preemraff Lysekil, is the biggest refinery facility in Scandinavia and according to Preem's webpage one of the most energy and environmentally efficient and modern refineries in Europe (Preem AB, 2018). Preemraff Lysekil is a large energy user, due to the energy intensive processes in the refinery. In 2015 they consumed 6.9 TWh of energy (Preem AB, 2016a), corresponding to 1.9% of Sweden's total energy use (370 TWh) (Energimyndigheten, 2017a). The energy used in the refinery comes primarily from burning fuel but also from electricity (Preem AB, 2016a). The fuel used is primarily fuel gas (non-condensable gases from the processes in the refinery), but also coke from cracking and imported liquefied natural gas (LNG). These energy sources combined with the large energy consumption makes Preemraff Lysekil a major emitter of fossil CO_2. In 2015 the refinery emitted 1.7 Mt CO_2 making Preemraff Lysekil one of the top five emitters of fossil CO_2 in Sweden (Preem AB, 2016a; Naturvårdsverket, 2017b). A summary of some of this data from Preemraff Lysekil is presented in Table 2.1.

Table 2.1: A summary of Preemraff Lysekil's yearly energy use, CO_2 emissions and capacity to refine crude oil (Preem AB, 2016a). Data for year 2015.

Capacity to refine crude oil	11.4 Mt
Refine crude oil	11 Mt
CO_2 emissions	1.7 Mt
Energy use	6.9 TWh
Fuel	6.4 TWh
Fuel gas	5.3 TWh
Coke	0.7 TWh
LNG	0.4 TWh
Electricity	0.5 TWh

The refinery is a so called complex refinery and this type of refinery often has higher yield compared to simple refineries. The difference between a simple and a complex refinery is that the complex refinery has an additional catalytic cracker while a simple refinery only has a hydro cracker (Leffler, 2008). There is also additional gas processing with a vacuum distillation unit (with a vacuum flasher and a visbreaker).

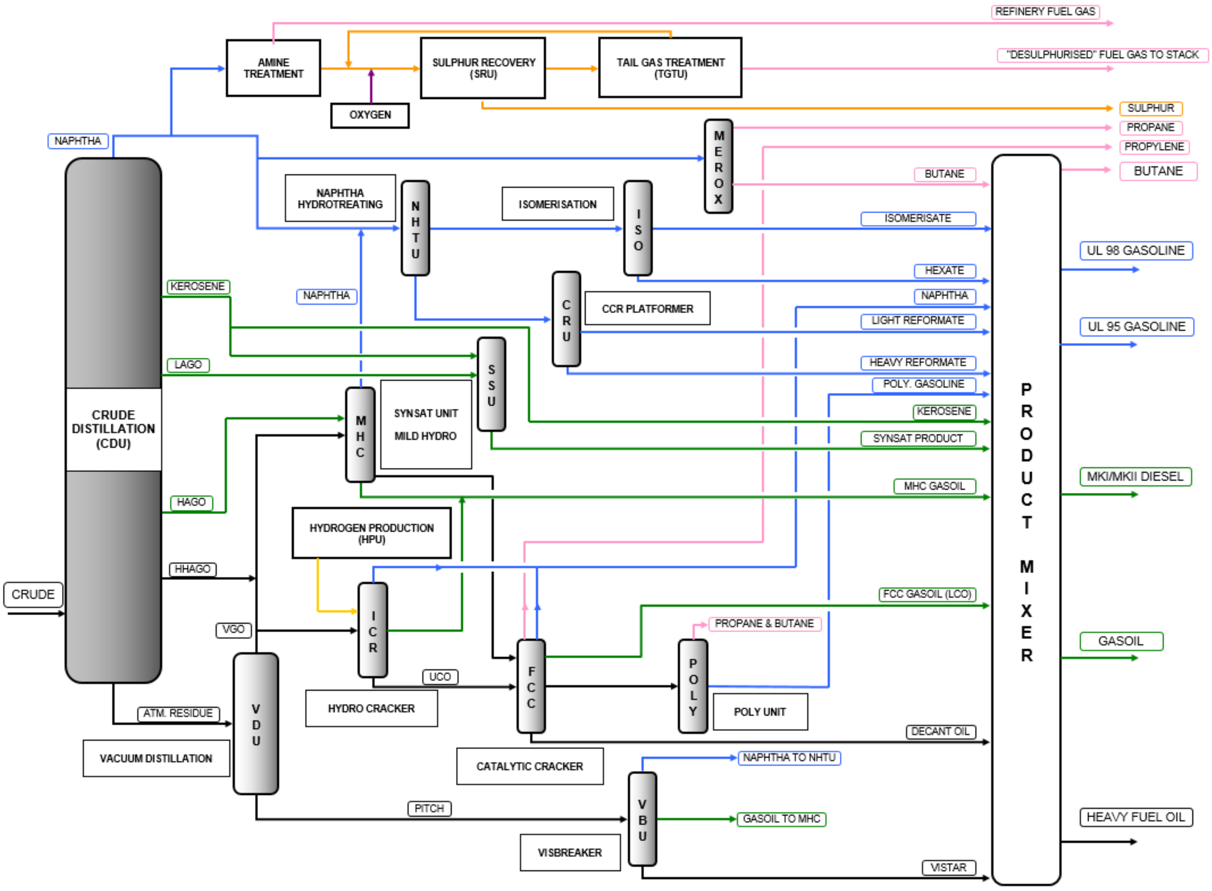

Figure 2.1: Overview of the process units and streams at Preem's refinery in Lysekil.

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In the refinery, crude oil is first distilled in the crude distillation (CDU), where the lighter products goes out in the top and the heavier products in the bottom (Leffler, 2008). The streams are then further processed in several process units (the process units that are concerned in this thesis are further described in the next section), as can be seen in Figure 2.1, before the final products are blended in the product mixer. The main products out from the refinery are diesel, gasoline, gasoil, propylene and HFO (heavy fuel oil).

2.1 Process units involved in this thesis

In this thesis, retrofit proposals (RPs) for the units NHTU, CRU, MHC and ICR are evaluated. These process units are described further in the sections below.

2.1.1 NHTU

The task of the naphtha hydrotreating unit (NHTU) is to desulphurize naphtha. The feed consists of naphtha from the crude distillation unit (CDU), the mild hydrocracker unit (MHC) and the visbreaker unit (VBU) and hydrogen from the steam reformer (Preem AB, 2016b). After the feed is preheated in the furnace, the feed is desulphurized in two catalytic reactors. This part of the process is presented in Figure 2.2.

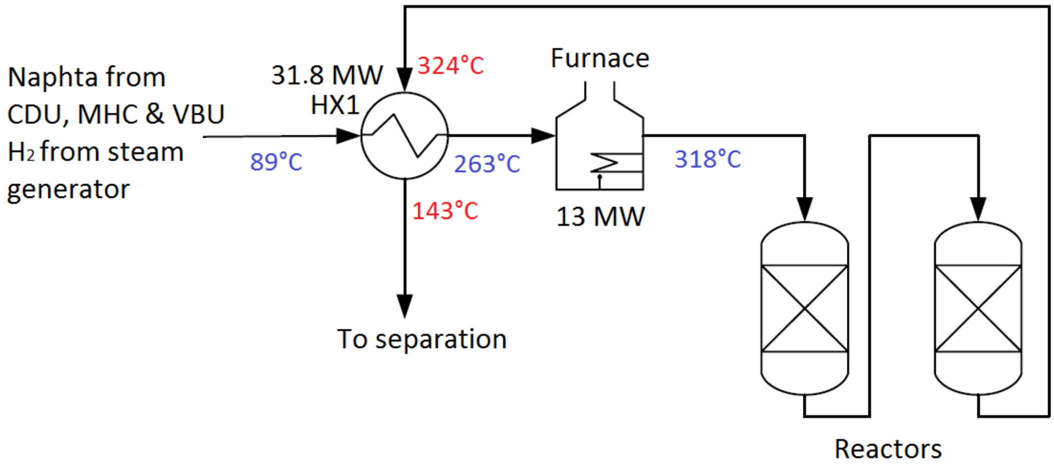


Figure 2.2: Current process scheme for part of the NHTU.

After desulphurization, the naphtha is stabilised in a distillation tower and then divided into a lighter and a heavier fraction. The heavy fraction is sent to the catalytic reforming unit (CRU), which is described in Section 2.1.2.

2.1.2 CRU

The purpose of the catalytic reforming unit (CRU), also called continuous catalyst regeneration (CCR), is to convert low octane number naphtha to high-octane components of gasoline (Preem AB, 2016b). The CRU is fed with hydrotreated naphtha from the NHTU. The feed is preheated before entering the platforming reactor. The reaction requires high temperatures (about 500°C), and downstream of the reactor the stream needs to be cooled in both a heat exchanger (HX) and an air cooler before separation. In the separation steps, first hydrogen and then hydrocarbons are removed. The final product from the CRU is high octane reformate, which is sent to the product mixer. A simplified process scheme of the unit including preheating, platforming reactor and cooling after reaction is shown in Figure 2.3.

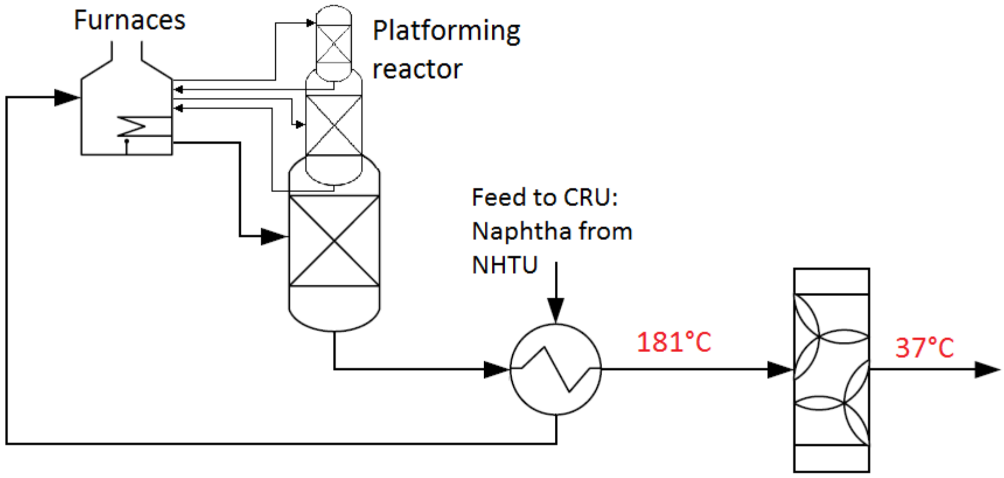


Figure 2.3: Current process scheme for part of the CRU.

The CRU and the NHTU are located close to each other at the refinery. Since the feed to the CRU is one of the streams from the NHTU, they are most often operated at the same time (Marton & Svensson, 2016). Therefore, these two process units can be possible to combine in a retrofit of their HENs.

2.1.3 MHC

In the mild hydro cracker unit (MHC), gas oil is desulphurized and heavy components in the gas oil are decomposed to lighter components through catalytic cracking (Preem AB, 2016b). The incoming gas oil feeds to the MHC unit are mainly heavy heavy atmospheric gas oil (HHAGO) and heavy atmospheric gas oil (HAGO) from the CDU and vacuum gas oil (VGO) from the vacuum distillation unit (VDU). In addition to the gas oil, hydrogen from the CRU is fed into the MHC unit.

The gas oil feed and the hydrogen feed are heated separately by the effluent flows from the reactors, in two HXs in series (Marton & Svensson, 2016), as can be seen in Figure 2.4. The two feeds are blended and heated in a furnace. After the furnace, the feed with gas oil and hydrogen has a temperature of around 400°C (Preem AB, 2016b). The feed then passes through two reactors in series where the catalytic cracking and desulphurization take place.

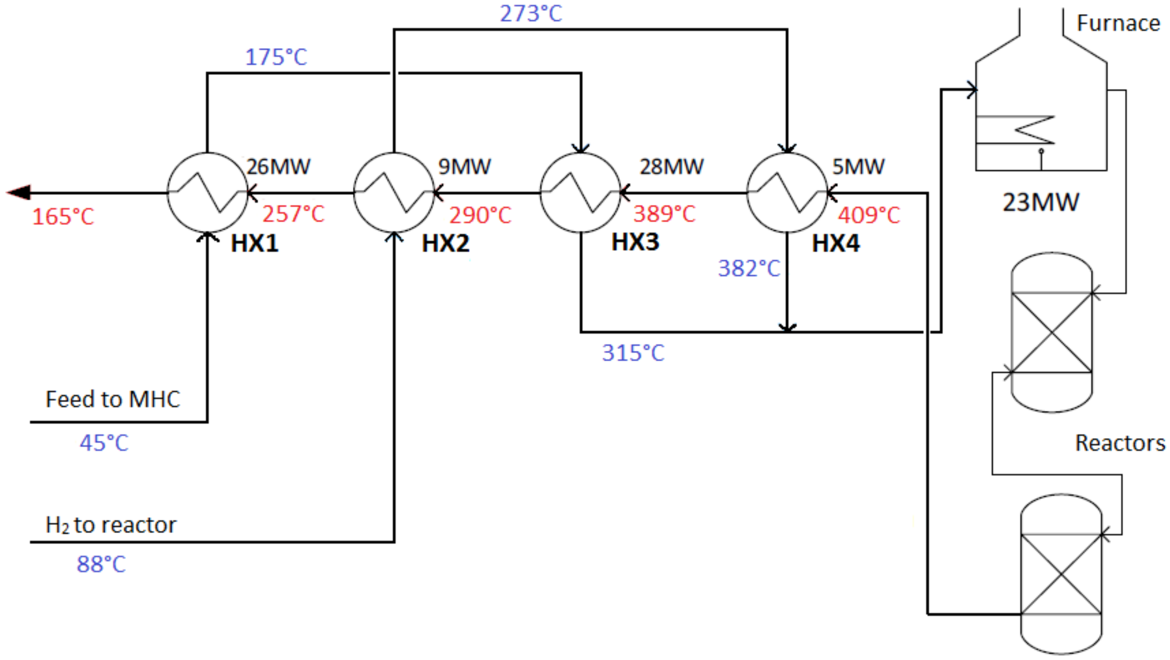


Figure 2.4: Current process scheme for part of the MHC.

The outgoing products from the MHC unit are naphtha, light gas oil and desulphurized vacuum gas oil (Preem AB, 2016b). Naphtha is further used in the NHTU unit, desulphurized vacuum gas oil is send to the catalytic cracker unit and the light gas oil goes to the product mixer.

2.1.4 ICR

The iso cracking unit (ICR) is a type of hydro cracker, and in this unit VGO is desulphurized and cracked in a reactor. VGO is fed into the unit and blended with hydrogen, and then it goes into the catalytic reactor. After the reactor, the stream is cooled in a HP (high pressure) steam generator and then the stream is separated through a number of gas/liquid separators. In the end of the unit, the stream is preheated before the furnace and then finally separated into naphtha, kerosene, gas oil and UCO (unconverted oil). The end of the unit together with the HP steam generator, including all temperatures, can be seen in Figure 2.5.

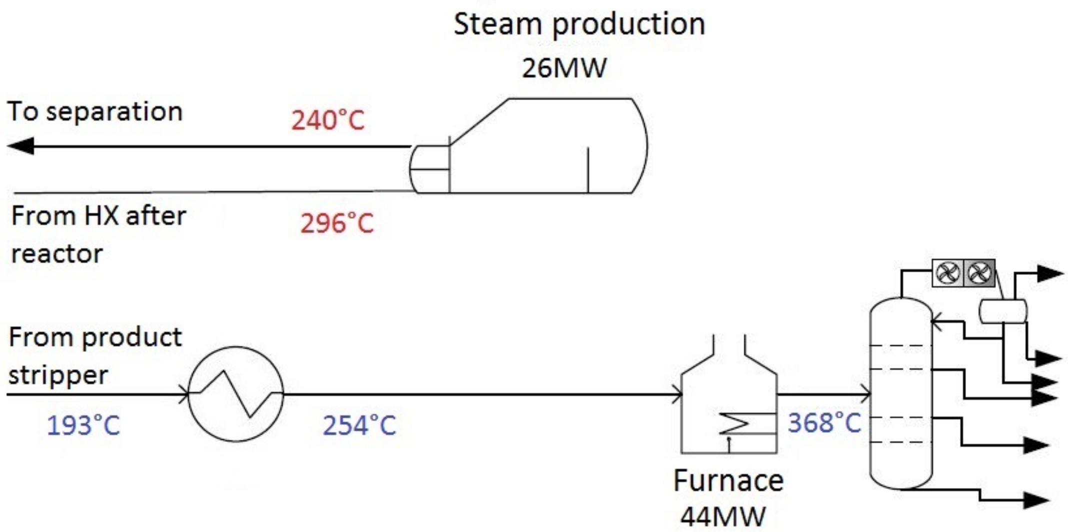


Figure 2.5: Current process scheme for part of the ICR.

2.2 Steam utility network

In Preemraff Lysekil's steam network, steam is produced at four main pressure levels. These levels are very high pressure (VHP), high pressure (HP), medium pressure (MP) and low pressure (LP) (Marton et al., 2017). The pressure of the different pressure levels are presented in Table 2.2 and the steam utility network is presented in Figure 2.6.

Table 2.2: The pressure for the different pressure levels (Subiaco, 2016).

Pressure level	Pressure [bar]
VHP	40
HP	22.4
MP	11.6
LP	4.6

VHP steam is produced in gas fired steam boilers and in flue gas heat recovery boilers (Preem AB, 2016b). Steam at the lower levels is mainly produced by steam generators (also called process coolers), but it is also obtained from reducing steam from a higher pressure level. These different production ways for obtaining steam can be seen in Figure 2.6. In the gas fired steam boilers, fuel gas is mainly used as fuel, but LNG is also used when there is a shortage of fuel gas. During some periods there is an excess of fuel gas produced in the refinery. To get rid of this fuel gas, additional steam is produced, which sometimes leads to an excess in steam. The excess steam is vented if it cannot be used in the process, which means that this energy will then be lost.

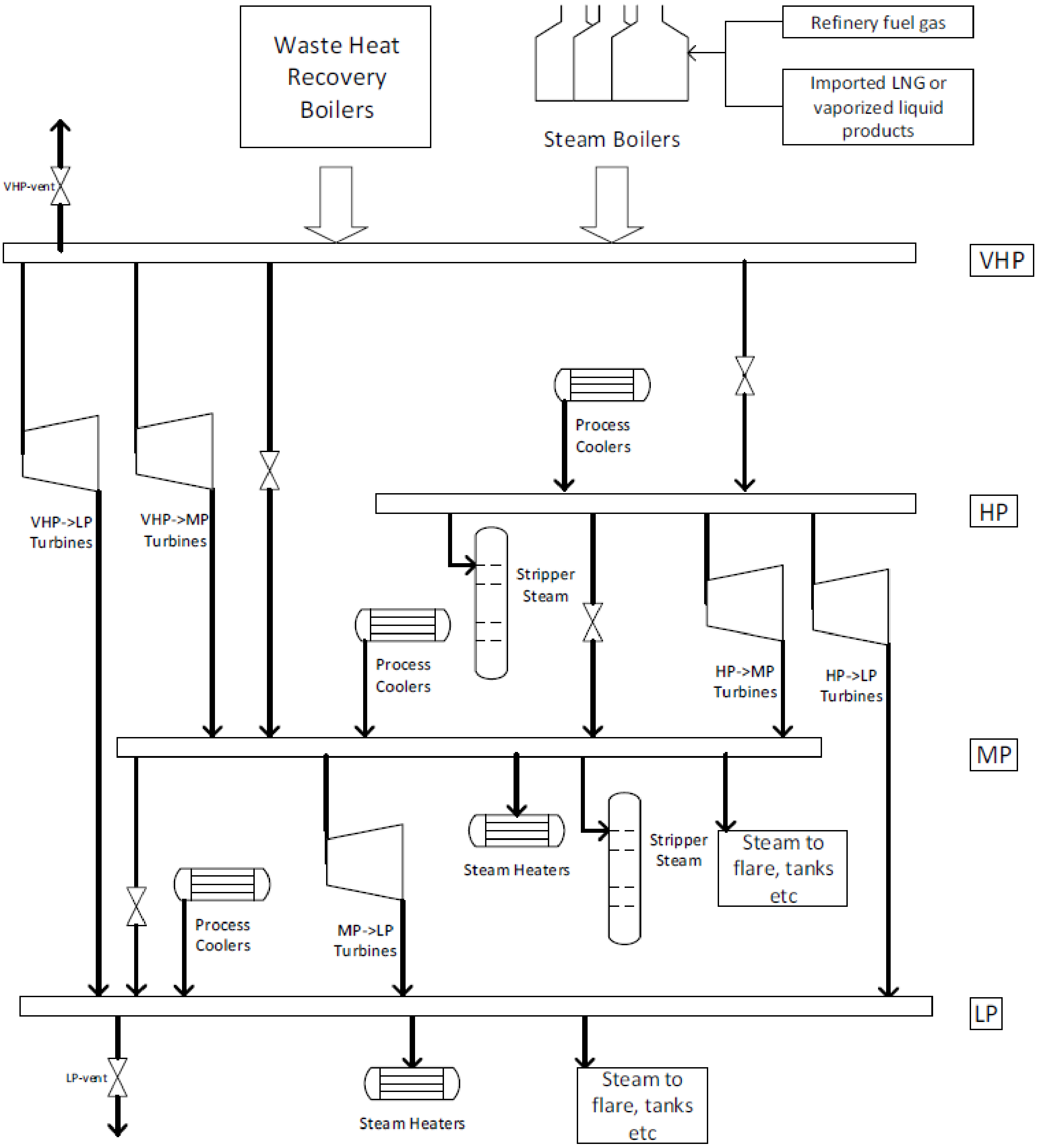


Figure 2.6: Steam utility network at Preemraff Lysekil. Source: (Marton et al., 2017).

To see how the steam network is affected by changes in the process, a steam network model was developed by Subiaco, (2016). This model uses scenarios from different operating points to see what impact changes would make depending on how the process works at that time. The different operating scenarios used in the model is one main scenario from September, a second from April and a third from July. Scenarios 1 and 2 are both considered normal operation points whereas scenario 3 is representative of maintenance period for some units of the refinery.

The steam network model was used for evaluation of the RPs by Marton (2017), to see how the steam, fuel and electricity balances are affected by the RPs that involved parts of the steam network. The results for the RPs investigated in this thesis can be found in Section 3.3 and 3.4.

2.3 Process equipment

The equipment involved in the RPs include air coolers, furnaces, pumps and HXs. The last two are of most interest in this thesis, due to the changes in the heat exchanger networks (HENs). Some theory about pumps and different types of HXs are presented in this section.

2.3.1 Heat exchangers

There are many different kinds of HXs, the two main types are tube-and-shell and plate-and-frame. Most of the HXs that are used today at Preemraff Lysekil are tube-and-shell HXs, thus this type is considered when adding new, or expanding old, HX units (Marton & Svensson, 2016) in RP2, 3 and 4. In RP1, welded plate HXs are proposed to be used. Some characteristics about these different types are described in this section.

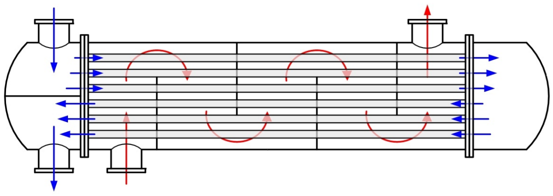


Figure 2.7: Example of tube-and-shell heat exchanger. ("BEM Type Shell and Tube Heat Exchanger (with Two Tube Passes)" (Oschal, 2014) (CC BY SA) Modified from original).

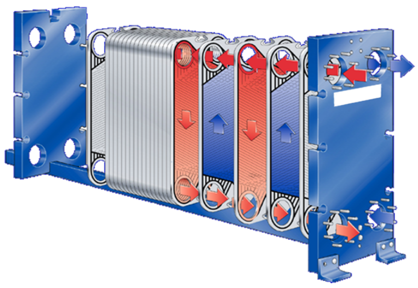


Figure 2.8: Example of plate-and-frame heat exchanger. ("Plate heat exchanger" (Rattanamaung, 2015) (CC BY SA) Modified from original).

Tube-and-shell HXs are found in most industrial fluid-processing plants and is the type that occurs most in calculations and examples in the literature (Ulrich & Vasudevan, 2004). The main concept of the tube-and-shell HX is that one fluid flows inside the tubes and one on the outside, the shell side. Some examples of tube-and-shell HXs are u-tube, floating head and fixed head, the latter can be seen in Figure 2.7. They can be designed for high pressures and temperatures and are adaptable for different combinations of fluids as to their state, pressure and flow rate (Coulson et al., 1999). Fouling inside HXs cause pressure drops and a decreased heat transfer coefficient and therefore cleaning of the HXs is important. The tube side is often possible to clean during operation, but to clean the shell side the HX needs to be taken off-line, which interrupts the process.

Plate-and-frame HXs are constructed by plates that are put side by side inside a gasket/frame, with passes in between them where the different fluids go (Hewitt et al., 1994). An example is shown in Figure 2.8. Plate-and-frame HXs are most suitable for liquid/liquid use. The advantages of plate HXs are that they are cheaper for a given load, mainly because they have around 2-4 times higher heat transfer coefficient than tube-and-shell HXs. This means they require less heat transfer area than tube-and-shell for the same heat load. Fouling is more problematic in plate HXs and they cannot be cleaned during operation, so if stops should be avoided, there need to be two parallel plate HXs so that the flow can be redirected when one has to be cleaned. Generally, plate HXs have a limited range of temperature and pressure they can handle, often up to around 30 bar in pressure. Welded plate HXs are more compact and robust than other plate-and-frame types and can thereby handle higher pressures. Some welded plate HXs can have a design pressure of up to 42 bar (Alfa Laval, 2016).

2.3.2 Pumps

Addition of new HXs, additional heat transfer area and associated piping, leads to increased pressure drops in the process units. Pumps are needed to drive the streams forward, and with increased pressure drops, more pump work is needed. This might lead to an increased load on existing pumps or need for new ones. The most common pump type in process industry is the centrifugal radial pump (Ulrich & Vasudevan, 2004). Thus, this is the type that is considered in this thesis. An example of a centrifugal radial pump is shown in Figure 2.9.

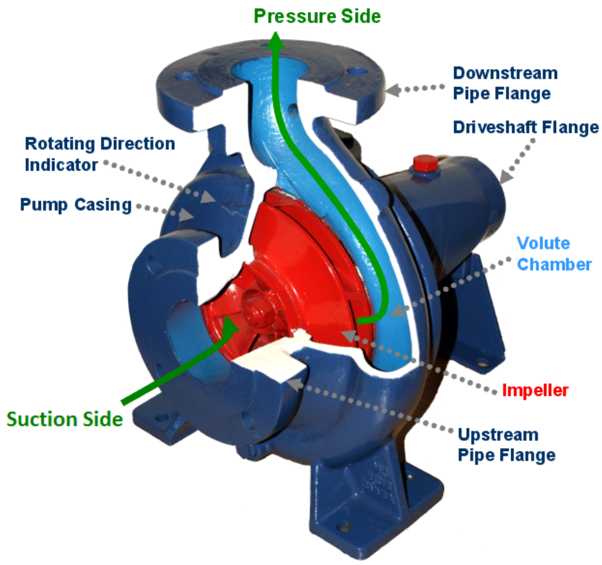


Figure 2.9: Example of centrifugal radial pump. ("Centrifugal Pump" (Fantagu, 2008) (CC BY SA) Desaturated from original).

3. Retrofit proposals

The selected retrofit proposals (RPs) that are evaluated in this thesis are presented in this chapter. The results from the previous interview study for each proposal are also described. The previous work of Marton & Svensson (2016) is the main source for data presented in this chapter, if nothing else is stated. The implications from the RPs, found by Marton & Svensson (2016) are presented in Table 3.1.

Table 3.1: Implications of retrofit measures.

Implications of retrofit measures	RP 1 (NHTU/CRU)	RP 2 (MHC)	RP3 (MHC)	RP4 (ICR)
De-bottlenecking		x	x	x
Network complexity		x		
Reduced load on a furnace	x	x	x	x
Reduced load on an air cooler	x	x		
Change in pressure drop	x	x	x	
Change in steam balance			x	x
Heat exchange between process units	x			
New equipment installation	x	x	x	x
Rebuilding existing equipment	x	x	x	
Pressure differences between streams or high pressures				x

3.1 Retrofit proposal 1 (NHTU/CRU)

The part of the NHTU affected by retrofit proposal 1 (RP1) is the streams into and out from the reactors, before the separation. The current layouts of the process, including temperatures and heat loads, are presented in Figure 2.2 and 2.3.

This RP involves preheating of the inlet stream to the NHTU with a stream from the CRU. This could decrease the load on the furnace in the NHTU by 7 MW, if the load on the existing HX (HX1) is kept the same by increasing its area. The stream that is used from the CRU is the one going out from the platforming reactor. This stream is currently cooled from 181°C to 37°C by an air cooler, which means this energy is lost. The proposed layout of the retrofit, including temperatures and heat loads, is presented in Figure 3.1.

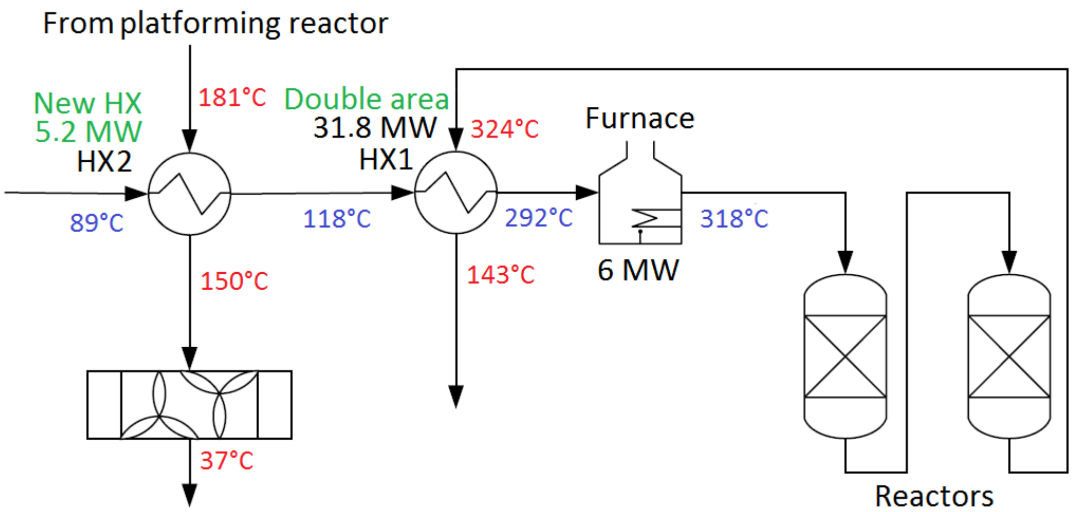


Figure 3.1: Retrofit proposal 1 - NHTU/CRU.

In the current process, HX1 is a large HX with four tube bundles. To keep the same load on HX1 (31.8 MW), the area needs to be twice as large due to the decreased temperature difference (if HX with the same U-value is used). Since there are spatial limitations, an addition of four new tube bundles would probably not be possible. Additionally, an increased area would lead to increased pressure drop. The pressure drop over this HX already limits the feed to this process unit. One solution for handling higher pressure drops could be to install a new pump.

Today there is a large impact from fouling in HX1, leading to big pressure drops, which means that there are interruptions in the production when the shell side of the HXs needs cleaning. According to the interview study one possible solution is to replace the current HX with parallel plate HXs. Then it would be possible to clean the equipment without interrupting the production, which would decrease the fouling and could lead to an increase in total production. Description about the different types of HXs is described in Section 2.3.1. The pressures in the streams vary from 8 to 35 bar, which is low compared to other process units but at the high end of the range where plate HXs can be used.

The new HX, HX2, handles one stream from CRU and one from NHTU. This could be a problem if the process units are not operated at the same time, as it could cause material problems in the HX. However, the units are most often operated at the same time given that the CRU handles a stream from the NHTU. This means that if NHTU runs at lower load, the feed to CRU will be smaller and thus the streams of the different units should be proportional. There might be a few cases when the CRU runs with another feed, a solution then could be to have a bypass on the hot side of the HX, to avoid damaging the equipment.

A way to mitigate the problem with the spatial limitations could be to add smaller new HX instead of the proposed sizes. If HX2 was smaller, less additional area would be needed in HX1 to keep the same load as before the retrofit. This would mean that fuel is still saved, but not as much, and that the investment cost would go down. The size of the new HXs could be cost optimised based on different minimum temperature difference (which in this RP is set to 25°C).

The energy that could be saved in RP1 is 7 MW of fuel in the furnace. There would also be a lowered demand for electricity to drive the air coolers for the stream from the platforming reactor.

3.2 Retrofit proposal 2 (MHC)

The main streams concerned in retrofit proposal 2 (RP2) are the in- and outgoing streams from the reactors in the MHC unit. In Figure 2.4 the current process scheme, including temperatures and heat loads, is presented and in Figure 3.2 the RP can be seen.

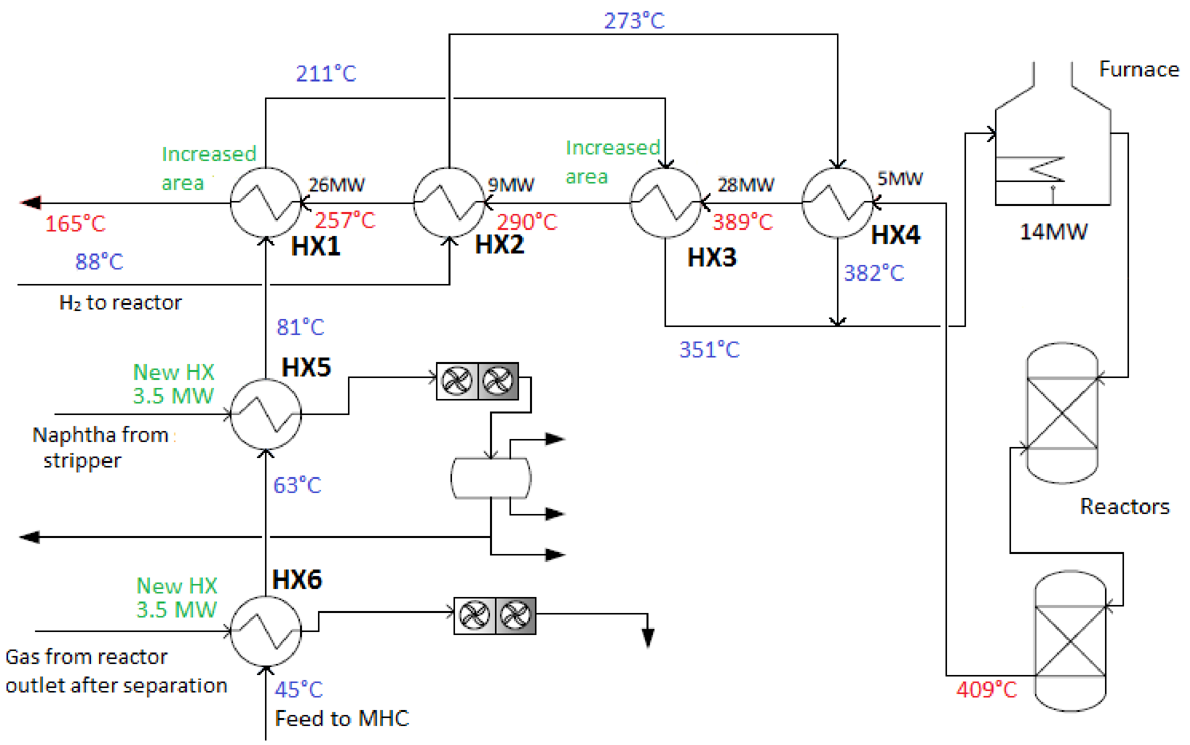


Figure 3.2: Retrofit proposal 2 - MHC.

In RP2 the feed into the reactors is preheated in two new HXs (HX5 and HX6), as shown in Figure 3.2, each with a load of 3.5 MW, in addition to the existing HXs (HX1 and HX3). Both HX1 and HX3 will need increased areas to be able to have the same load as in the current configuration. Since HX1 and HX3, in the existing design, consist of large tube-and-shell HXs, the heat transfer area cannot be increased without adding new tube bundles. There are however some spatial limitations in this area of the refinery.

The outflow from the reactors is cooled in HX4-HX1 before it is separated in a gas/liquid separator. The gas flow out from this separation is further used as a hot stream in HX6. In HX6 the feed to MHC is heated from 45°C to 63°C by the gas from the reactor outlet. The feed to MHC is then heated from 63°C to 81°C in HX5 by a top stream of naphtha from a stripper column. After the feed has been preheated in HX6 and HX5, it continues the same way as in the current design but with the additional heat transfer area in HX1 and HX3. The additional heat transfer areas in HX1 and HX3 is added to be able to keep the load on the HXs and to not affect the target temperature on the hot side. In the RP, the temperatures of the feed out from HX1 and HX3 are 211°C and 351°C, respectively. The feed in to and out from the reactor have high pressure (approximately 69 bar). The hot stream in HX6 has almost the same pressure as the flow out from the reactor, but the hot stream in HX5 has a pressure of around 1.5 bar.

Since the feed to the reactors is preheated, the temperature of the flow into the furnace is higher and therefore the fuel consumption in the furnace can be reduced. In this proposal, the fuel consumption could be decreased by 9 MW resulting in a total furnace load of 14 MW.

In the current configuration, there is a limitation in the flue gas channels connected to the furnace. At the end of each catalyst cycle, the catalyst is less efficient. Thus, to avoid a large reduction of the reactions in the reactor, the temperature of the feed into the reactor needs to be higher. To get a higher temperature on the feed it needs to be heated more in the furnace, which means that more fuel is burned and more flue gases are formed. In the current configuration, the flue gas channel is too small to be able to handle the extra flue gases that comes from the additional combustion. This makes the flue gas channel a bottleneck for the production, and the production is decreased to deal with this problem. The reduced furnace load could however mitigate this issue and make it possible to keep the production on the same level instead of decreasing it. This would however mean that less fuel is saved during the time when the production is increased.

Depending on the type of raw material that is fed into the process unit, and on how the process unit runs, the properties of the hot streams in HX5 and HX6 can vary. These streams also depend to a large extent on other streams in the process unit. This causes uncertainties regarding how the temperatures and flow rates in the process unit will vary. Variation of the stream properties in addition to these uncertainties can also cause problems with flexibility and how the processes are run.

3.3 Retrofit proposal 3 (MHC)

Retrofit proposal 3 (RP3) is presented with a process scheme, including temperatures and heat loads, in Figure 3.3. RP3 is similar to RP2 but instead of letting internal process streams preheat the feed, LP steam is used. The LP steam (4.6 bar) is used to preheat the feed from 45 °C to 94 °C in HX5, which corresponds to about 10 MW. Currently in the refinery, there is a surplus of LP steam during most of the year, that can be used to preheat the feed.

The preheat of 10 MW means that the existing HXs (HX1 and HX3) will need increased areas to be able to have the same load as today, due to decreased temperature differences. As mentioned earlier, the existing HXs are tube-and-shell HXs and therefore the heat transfer areas are difficult to increase, and instead new tube bundles will be added to the current ones. The preheating of the feed also means that the temperature of the feed out from HX1 and HX3 will be 224°C and 364°C, respectively. This leads to a possible decrease of the fuel load on the furnace with almost 13 MW, from 23 down to 10 MW, resulting in large savings of fuel.

As mentioned in the previous section, the flue gas channel is a bottleneck for the production capacity of this process unit in the end of each catalytic cycle, due to the need for increased temperatures for the feed into the reactor. This means that this RP also can make it possible to keep the production level in the unit since the load of the furnace is decreased.

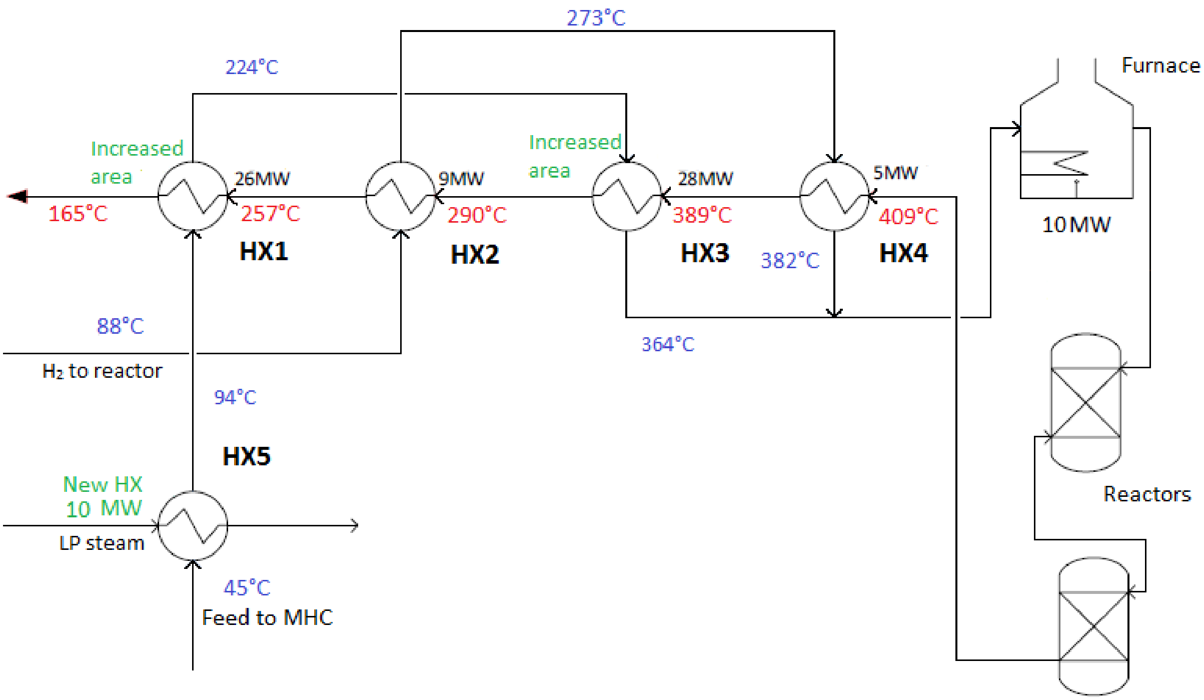


Figure 3.3: Retrofit proposal 3 - MHC.

The largest disadvantage with this RP is the problem with limitations in spatial area. This means that it is difficult to increase and add new HXs in the area, similar to RP2.

The effect of this RP on the steam network was modelled in previous work (Marton et al., 2017). Different operating scenarios were investigated using the model to see how the increased consumption of LP steam affects the steam, fuel and electricity balances in different ways. For the main scenario, which is considered in this thesis, the increased need for fuel in the steam boilers is 0.7 MW. More information about the steam network model, and its different scenarios, can be found in Section 2.2.

3.4 Retrofit proposal 4 (ICR)

The point of retrofit proposal 4 (RP4) is to increase the preheating of the product stream before the furnace, that lies upstream of the product fractionator, to be able to decrease the load on the furnace. This is done by adding an additional HX before the furnace. The current configuration of this part of the unit is shown in Figure 2.5 and the RP is shown in Figure 3.4.

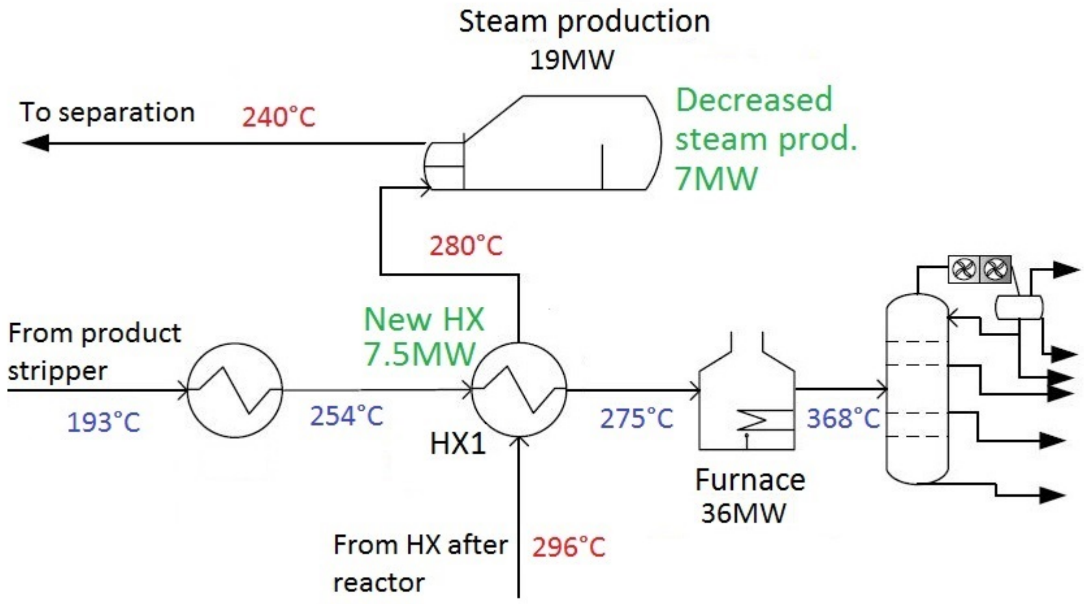


Figure 3.4: Retrofit proposal 4 - ICR.

As can be seen in Figure 3.4, the stream that is used as hot stream in the new HX, HX1, is usually used to produce steam. The produced steam is HP-steam, with a pressure of 22.4 bar, see Section 2.2. As can be seen in Figure 2.6, HP steam is either used as stripper steam, in turbines to create work or throttled down to MP-steam.

The effect from this RP on the steam network was investigated in previous work (Marton et al., 2017), just as RP3, to see how the reduced production of HP steam affected the steam, fuel and electricity balances of the process. For the main scenario, which is considered in this thesis, there is no change in either fuel gas consumption in the steam boilers or in the sites electricity consumption due to this RP.

The possible fuel savings from the RP are around 8 MW. The furnace is currently a bottleneck in the process, since its limited capacity hinders the stream from reaching the desired temperature before separation. This gives a lower yield than what would otherwise be possible. The increased available capacity could therefore be used to increase the temperature further, instead of just saving energy.

Since the working pressure in this unit is very high, around 140 bar, it needs special equipment that can handle high pressures. The hot and the cold streams that are involved in the RP have the pressures 132 bar and around 1 bar, respectively. The new HX, and any necessary surrounding equipment, will therefore be expensive.

One of the main problems with this RP, besides high costs, is that there are spatial limitations in the unit. Also the temperatures of the streams vary a lot, as they are highly dependent on where in the catalyst cycle the reactor is working. This means that a new possible temperature into the product fractionator will be hard to estimate.

4. Non-energy benefits

As stated in Section 1.1, non-energy benefits (NEBs) are positive effects from energy efficiency improvements, other than savings of energy and utility costs. It has been shown in earlier studies, for example by Pye (1998), that the value of NEBs is often larger than the value of energy savings. Including the economic value and other strategic values of NEBs can provide additional incentives to implement energy efficiency projects (Pye & McKane, 2000). Today, energy efficiency is often not the highest priority in business, as there often are other investments that are considered more important and/or will bring a higher profit. If NEBs of energy efficiency measures are quantified, the PBP of an investment might be shortened and the net revenue higher. Including motivations like higher productivity or better overall process efficiency, can in this way make energy efficiency measures more compelling for companies.

According to Cooremans (2012), the strategic value of an investment is often more important than its profitability. The strategic value of an investment is a subjective value which depends on which other strategies the company have. There are different categories of investments and energy efficiency is often not considered as strategically important as other types of investments. By pointing out all values of an energy efficiency measure, including NEBs, they can be able to move up to a higher prioritised category. However, the profitability of an investment is still always evaluated, even though it is not the only important aspect of the decision making. This makes it important to estimate the monetary values of the NEBs.

Industrial productivity is one of the categories of NEBs, and within this category there are many different types. Examples of NEBs concerning for example production are: operability, maintenance, work environment, waste and emissions (Nehler, 2016). According to Rasmussen (2014), NEBs are quantifiable to varying levels. Some of the NEBs that could be included in this thesis, under the category industrial productivity, are listed in Table 4.1 in order of how high quantifiability they have.

Table 4.1: Quantifiability of different non-energy benefits according to Rasmussen (2014).

High quantifiability
Improved equipment performance
Increased production
Reduced amount of raw material
Reduced labour cost
Reduced maintenance cost
Reduced operational cost
Reduced wear and tear of equipment
Medium quantifiability
Improved efficiency
Improved process control
Improved product quality
Increased capacity
Lowered cooling demand
Productivity gains
Reduced emissions
Low quantifiability
Additional space
Competitive advantage
Decreased noise
Improved air quality
Reduced risk (legal, energy price & energy supply)

As NEBs are very different from case to case, this ranking might not be applicable for all processes. However, it can be a good starting point when trying to decide which NEBs to include in an economic evaluation.

Other than the NEBs listed above, the RPs also affect the operability of the process, defined in Section 1.1. This includes for example flexibility, controllability and reliability of the process.

5. Methodology

In this section, the methodology that was used in this thesis is described. Later in the chapter, more detailed methods for calculations and further evaluations are presented. The general workflow for this thesis can be seen in Figure 5.1.

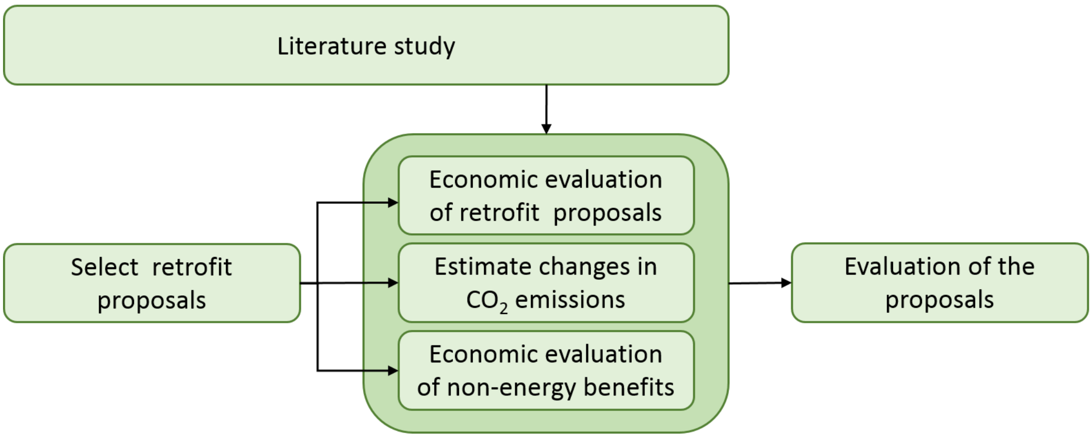


Figure 5.1: The workflow for this thesis.

First a literature study was carried out. It was focused on the processes in a refinery, previous work in the project between Chalmers and Preem, methods for economic evaluation and non-energy benefits (NEBs) of energy efficiency. The literature study was continued in parallel during most of the work.

Based on previous work, four of nine retrofit proposals (RPs) were chosen. The choice was made based on the results from the interview study, where the four chosen ones were considered most interesting for the aim of this thesis since they included interesting possible NEBs. Every chosen proposal was evaluated by its increase in energy efficiency, change in costs, change in CO_2 emissions and other NEBs.

An economic evaluation of the chosen RPs was made, where investment cost and decrease in utility costs were considered. The method for this is presented in Section 5.1 and 5.2. The value of the reduced CO_2 emissions due to decreased fuel consumption was also calculated. Additional literature review was made about NEBs and then evaluation of some of the other found NEBs was made. This is further described in Section 5.3. Finally the value of the energy efficiency measures was compared, with and without including the evaluated NEBs. All calculations were made in Excel or MATLAB.

5.1 Investment cost

The investment cost, or capital cost, is the fixed cost of an investment in new equipment. Investment costs of process equipment are often estimated by cost functions depending on the capacity of the equipment (for example by area, volumetric flow rate or power). For heat exchangers (HXs), the capacity that is most commonly used is the heat transfer area. A simple type of cost function is shown in Equation 5.1, where A is the heat transfer area and a, b and c are coefficients that are specified for different types of HXs. To make the result more accurate, more complex functions can be used.

$$\text{Installed Capital Cost} = a + bA^c \tag{5.1}$$

The heat transfer areas of the new HXs were calculated from Equation 5.2, where Q is the load of the HX, U is the overall heat transfer coefficient, A is the heat transfer area and ΔT_{lm} is the logarithmic temperature difference. The U-values for the HXs were estimated from data in the literature (Ulrich & Vasudevan, 2004). The estimated U-value for all HXs can be found in Appendix A. ΔT_{lm} was calculated from Equation 5.3 with temperatures found in the process data gathered during previous work. Q was also found in data from previous work. The areas and U-values were compared with specifications from existing equipment at Preem, to see if they where reasonable.

$$Q = UA\Delta T_{lm} \tag{5.2}$$

$$\Delta T_{lm} = \frac{(T_{hot,in} - T_{cold,out}) - (T_{hot,out} - T_{cold,in})}{ln\frac{(T_{hot,in} - T_{cold,out})}{(T_{hot,out} - T_{cold,in})}} \tag{5.3}$$

There are many cost functions available in the literature, and different methods for investment cost estimations result in different investment costs. Therefore, it is important to use the same method when evaluating alternatives for plant design (Feng & Rangaiah, 2011). Well known literature about economic analysis of process design was searched for cost functions. Since not all of these sources included cost functions for plate HXs, a selection was made among the sources that did. The same source was then used for cost calculations for tube-and-shell HX and pumps. Cost functions for plate HXs were found in these four different books by: Sinnott & Towler (2009), Peters, Timmerhaus & West (2003), Turton et al. (2012) and Ulrich & Vasudevan (2004). The functions from these sources are presented in Figure 5.2, where the purchased cost for plate HXs is shown as a function of heat transfer area. The functions were plotted from their lower until their upper capacity bound, i.e. to the highest area where the functions are valid.

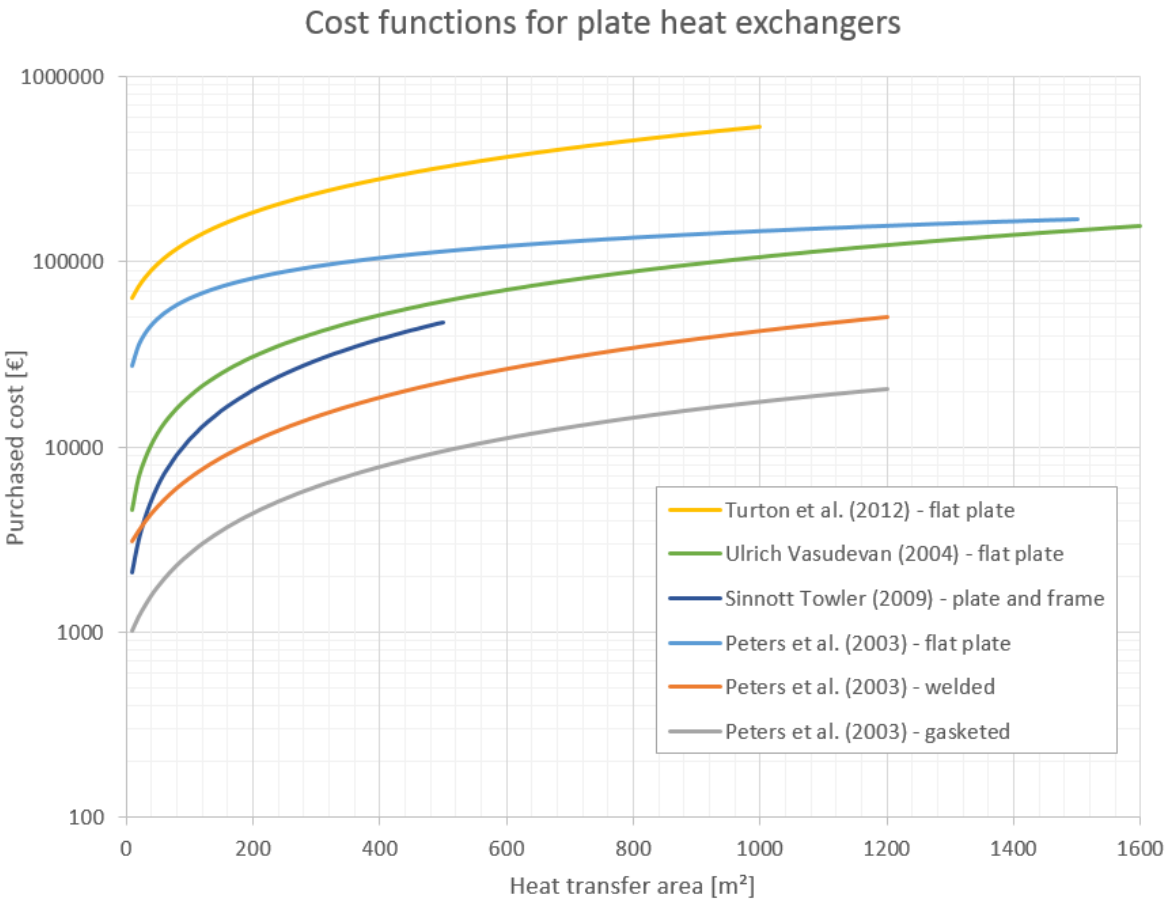


Figure 5.2: Purchased cost for plate heat exchangers as a function of heat transfer area. Comparison between different cost functions. Plate material is carbon steel for all functions and they are valid for normal pressure and temperature. All functions are converted to 2016 values. Note the logarithmic scale of the cost axis.

The cost functions are made through collected cost data from equipment manufacturers from different years. Since the functions are based on data from 2001 to 2007, the result was converted to 2016 exchange rate and CEPCI, the chemical engineering plant cost index, to represent 2016 cost levels. CEPCI is the most common cost index used to adjust process plant construction costs and process equipment cost from one period to another due to inflation (Turton, 2012). Equation 5.4 shows how to use the CEPCI values to convert a known cost (C_1) to a new (C_2). The value that was used was the annual CEPCI for Equipment: HXs and tanks, which was 557 in 2016 (“Economic Indicators”, 2017).

$$C_2 = C_1 \cdot \frac{CEPCI_2}{CEPCI_1} \tag{5.4}$$

As can be seen in Figure 5.2, the results from the different cost function differs a lot. Therefore, a sensitivity analysis was made for RP1, to show how the results can change with use of different cost functions. The results from the sensitivity analysis are presented in Figure 6.2 in Section 6.1.

The source of cost functions that was used for all other equipment in this thesis was Ulrich & Vasudevan (2004). Ulrich & Vasudevan provide cost functions for a large variety of equipment and their cost functions also have the capacity to estimate values for large areas, which makes it easy to use. Furthermore, they use specific installation factors (presented in Table 5.2) for different types of equipment, instead of the Lang-factor which is the same for all parts of a plant (Ulrich & Vasudevan, 2004).

Since all costs in Figure 5.2 are for base cases with carbon steel as material and normal pressure, the actual purchased cost in the RPs studied in this thesis will be higher. The desired material in all the new HXs was assumed to be stainless steel, and there are high pressures in many of the new HXs. Corresponding cost correlation were made using Equation 5.5 where f_p and f_m are cost factors for pressure and material and C_B is the cost for the base case. All factors that were used for each new HX are showed in Table 5.2.

$$C_P = f_p f_m C_B \tag{5.5}$$

For tube-and-shell HXs, the pressure factor was estimated from a pressure function and varies between f_p=1.1 and 1.2 for the concerned pressures (Ulrich & Vasudevan, 2004, p. 384, Fig 5.37). No functions for pressure factors for plate HXs were found in the literature. However, Ulrich & Vasudevan (2004) claims that f_p=1.0 for flat plate HXs, although this is said to only be valid for pressures up to 27 bar. To take the high pressures into account for plate HX in RP1, a value of f_p=1.1 was chosen, which is the value for a tube-and-shell HX with the same pressure (35 bar).

The material cost factors for plate HXs are shown in Table 5.1. Stainless steel is the most common material in plate HXs and also the material of the old HXs in RP1, therefore f_m=2.3 was chosen for those. The material in all new tube-and-shell HX that are installed was also assumed to be stainless steel, which gives f_m=3.0 (Ulrich & Vasudevan, 2004).

Table 5.1: Material adjustment factors for plate heat exchangers.
(Ulrich & Vasudevan, 2004)

Plate material	**Factor**
Carbon steel	1.0
Copper	1.2
Stainless steel	2.3
Nickel alloy	2.8
Titanium	7.2

Finally, to get the investment cost for the equipment, the purchased cost needs to be multiplied with an installation factor, f_i, as shown in Equation 5.6.

$$C_{cap} = f_i C_P \tag{5.6}$$

The installation factor represents the costs for installation labour and material as a share of the total purchased cost of the equipment. Ulrich & Vasudevan (2004) use specific factors for different kinds of equipment, and for HXs, a value of 3.76 is suggested. This includes both installation material (piping, instruments etc.), cost of labour for installation and indirect costs, such as for example contractor engineering expenses (Ulrich & Vasudevan, 2004, p.341). This factor was used for all types of HXs.

Table 5.2: Summary of factors for estimation of investment cost of heat exchangers.

	Type	Pressure [bar]	Pressure factor, f_p	Material	Material factor, f_m	Installation factor, f_i
RP1		hot/cold				
HX 1	Plate	28/35	1.1	SS	2.3	3.76
HX 2	Plate	8/35	1.1	SS	2.3	3.76
RP2		tube/shell				
HX 1	U-tube	67/70	1.2	SS	3	3.76
HX 3	U-tube	67/70	1.2	SS	3	3.76
HX 5	U-tube	70/1	1.1	SS	3	3.76
HX 6	U-tube	64/70	1.2	SS	3	3.76
RP3		tube/shell				
HX 1	U-tube	67/70	1.2	SS	3	3.76
HX 3	U-tube	67/70	1.2	SS	3	3.76
HX 5	U-tube	70/5	1.1	SS	3	3.76
RP4		tube/shell				
HX 1	U-tube	134/1	1.12	SS	3	3.76

5.1.1 Pumps

Since all the RPs include increased heat transfer area and/or new HXs, they will all lead to increased pressure drops. Every proposal was evaluated with respect to the need for additional pump work. As an estimation for the additional pressure drops, and thereby pump work, specifications of the old HXs were used. The new HXs were assumed to be of the same type and have similar dimensions, and thereby the pressure drop was assumed to increase in proportion to the increased heat transfer area.

The most common pump type in process industry is centrifugal radial pump (Ulrich & Vasudevan, 2004), therefore this was the type that was considered when making calculations. Cost function from Ulrich & Vasudevan was used, just as when estimating cost of HX.

Ulrich & Vasudevan (2004) uses shaft power in kW as capacity unit in their cost function, which was calculated through Equation 5.7. w_s is the shaft power in kW, $\dot{q}$ is the volumetric flowrate in m^3/s, Δp is the additional pressure drop in kPa and ϵ is the efficiency of the pump (0.85 for centrifugal radial pump according to Ulrich & Vasudevan (2004)).

$$w_s = \frac{\dot{q}\Delta p_{add}}{\epsilon} \tag{5.7}$$

The cost from the cost function was converted into 2016 value in the same way as for the HXs. CEPCI for pumps was used, which was 970 in 2016 ("Economic Indicators", 2017). Then factors for material, pressures and installation were also added. All factors are presented in Table 5.3. The purchased cost is for 11 bar, so above that pressure factors were found in a pressure diagram (Ulrich & Vasudevan, 2004, p. 391, Fig 5.50).

Table 5.3: Summary of factors for estimation of investment cost of pumps.

	Max pressure [bar]	Pressure factor, f_p	Material	Material factor, f_m	Installation factor, f_i
RP2	70	2.4	SS	1.9	3.61
RP3	70	2.4	SS	1.9	3.61
RP4	134	4.5	SS	1.9	3.61

5.2 Utility costs

The change in utility cost comes primarily from savings of fuel in the furnaces. The fuel that is saved in the different RPs, was assumed to be liquefied natural gas (LNG). This was assumed since the refinery, besides using their own fuel gas from the processes, imports LNG to cover their heating demand in the processes. LNG is only used as fuel when there is a deficit of fuel gas, which is the case approximately 75% of the year (Marton et al., 2017). The refinery could also use both butane and propane instead of LNG, however LNG was assumed to be the marginal fuel in this thesis.

The decreased fuel demand was calculated by Equation 5.8 where ΔQ_{fuel} is the change in demand of fuel, Q_{heat} is the demand of heat from the furnace before and after the retrofits and $\eta_{furnace}$ is the efficiency in the furnace. $\eta_{furnace}$ is 0.78 for the furnace in RP1, 2 and 3 and 0.93 for the furnace in RP4 in the calculations. The annual savings in utility cost, from the decrease of fuel, was calculated by Equation 5.9 where ΔC_{fuel} is the change in fuel cost, P_{fuel} is the price of the fuel and t is the operating time in hours per year. The price of LNG was set to 0.97 €cent/MJ, corresponding to 35 €/MWh, which was the average price for a large industry in Sweden in 2016 (eurostat, 2018).

$$\Delta Q_{fuel} = \frac{(Q_{heat,old} - Q_{heat,new})}{\eta_{furnace}} \tag{5.8}$$

$$\Delta C_{fuel} = \Delta Q_{fuel} P_{fuel} t \tag{5.9}$$

There is also energy saved in form of electricity that drives the air coolers due to decreased cooling demand. However, the change in electricity use in this area is a relatively small amount and was therefore neglected in the calculations of utility cost savings (Marton & Svensson, 2016). The additional electricity that would be needed for the pumps due to increased pressure drops was however taken into account.

The cost of the additional electricity due to increased pump work was calculated by using Equation 5.7, to find the additional shaft work needed, and then multiplying this with the cost of electricity and the operation time per year, see Equation 5.10. The price of electricity was assumed to be 2.88 €cent/kWh, which is an average value for 2016s spot price on NORD POOL (2018).

$$\Delta C_{el} = \Delta w_s P_{el} t \tag{5.10}$$

In RP3 and RP4 the steam network is affected, see Section 3.4, which could cause a change in the fuel or electricity balances. The results from the main scenario in the steam network model was used for both RPs (Marton et al., 2017). For RP3 the fuel demand in the steam boilers increased with 0.7 MW due to the use of additional LP steam, which was included in the calculation of utility costs for that RP. There was no increase in either fuel or electricity demand for RP4. The main scenario is from one operating point where the production is considered stable and normal. The other scenarios for the steam model showed different results, but these were not evaluated in this thesis.

5.3 Evaluation of non-energy benefits

First all the NEBs (concerning industrial productivity) for each RP were listed to bring an overview of all effects. The result of this evaluation is presented in Section 6.3. The NEBs which were considered most straightforward to quantify, increased production and CO_2 emissions, were then evaluated further.

5.3.1 Increased production

Increased production can be a result of several implications of the retrofit measures that are evaluated in this thesis. From previously mentioned process implications (see Table 3.1), the following can result in increased production: De-bottlenecking by reducing load on a furnace or an air cooler or rebuilding of existing equipment. The NEBs, and the values of the ones that were evaluated, of each RP are presented in Section 6.3.

As can be seen in Figure 2.1, the refinery's many units are connected and it is difficult to see which units will affect which products. To calculate the value of increased production, several assumptions have to be made and some questions regarding the processes need to be answered. The questions are:

- How much of the operating time is the process limited by the concerned part of the process?
- Which products would be affected by the increased production?
- What is the value of these products?

All numbers that were needed, and assumptions that were made, to be able to calculate the value of the increased production are listed in Table 5.4. The numbers and assumptions needed to evaluate the increased production were first estimated from literature and then discussed together with process engineers at Preemraff Lysekil, to validate their reasonability.

Table 5.4: Data needed and assumptions made for estimation of values of increased production. The numbers without reference were assumed after discussion with engineers at Preemraff Lysekil.

For RP1, 2 and 3:
• 11 Mt crude oil is refined/yr (Preem AB, 2016a)
• 30 l gasoline produced from 1 barrel crude oil
• The gross margins for gasoline and diesel are 1.42 SEK/l and 1.07 SEK/l, respectively (SPBI, 2018)
For RP1:
• The increased production only affects the production of gasoline
• The production can be increased by an average of 3% during 50% of the operating time, which gives a production increase of 1.5%
For RP2 and 3:
• The increased production affects the production of gasoline and diesel.
• The increased amount of product is half gasoline and half diesel and they are assumed to need the same amount of crude oil per l
• The production can be increased with an average of 5% during 20% of the operating time, which gives a production increase of 1%

From these assumptions, the value of the increased production was calculated. The RPs affects the productivity in different ways. For RP1, 2 and 3 the production can be increased but for RP4 the quality of the products can potentially be increased instead. No values for increased quality of products were calculated but the effects are discussed in Section 7.4.

Since less fuel will be saved if the production increases, a new value of the fuel demand was calculated. The fuel demand was assumed to increase with the same percentage as the production increases. Meaning the fuel demand will be 1.5% higher when the production is increased with 1.5%, giving a smaller saving in fuel demand.

Since there are uncertainties regarding the increased production for RP2 and 3, a sensitivity analysis was made to see how the annual value of increased production varies with different assumptions. In the sensitivity analysis the increased production was assumed to increase by 1-10% during 10-30% of the time.

5.3.2 CO_2 emissions

To estimate the monetary value of reduced emissions of greenhouse gases (GHG) the amount of emitted CO_2 equivalents, CO_2-eq, was first estimated. Then the cost savings were calculated from the decreased need for CO_2 permits from EU ETS (Emission Trading System), assuming that they could sell emission permits equal to their reduced emissions of CO_2-eq. Only local emissions were considered. Thus, emissions from increased electricity use were not taken into account.

The decrease of CO_2 emissions comes from the decrease of fuel used in the furnaces in all the RPs. Fuels emit different amounts of CO_2 due to their carbon intensity. The fuel was assumed to be LNG, as in the calculation of utility costs, and LNG emits 71.8 g CO_2-eq/MJ (258 kg CO_2-eq/MWh) when combusted (Naturvårdsverket, 2017c; Gode et al., 2011). These numbers together with Equation 5.11 were used to calculate the annual decrease of CO_2 emissions. Δe_{CO_2}

is the change in emissions of CO_2-eq, ΔQ_{fuel} is the change in fuel demand, c_{fuel} is the specific emissions of the fuel and t is the operating time per year.

$$\Delta e_{CO_2} = \Delta Q_{fuel} c_{fuel} t \tag{5.11}$$

The cost of CO_2 permits was set to 5.5 €/ton, which is an average value for the emission permits in EU from 2016 (Energimyndigheten, 2017b). The annual cost savings from decreased CO_2 emissions was calculated from Equation 5.12 where ΔC_e is the decrease in cost of emissions on a yearly basis and $C_{permits}$ is the cost of emission permits per ton CO_2-eq.

$$\Delta C_e = \Delta e_{CO_2} C_{permits} \tag{5.12}$$

5.4 Profitability calculations

Payback period (PBP) is an indicator used to quantify the economic risk of an investment. It shows how long time it takes for the total return of an investment to equal the fixed-capital investment (Peters et al., 2003). It can be calculated through Equation 5.13, where the total investment is the investment cost of all new equipment in each RP, and the annual net revenue is the annual earnings (from reduced cost of fuel, additional profit from increased production etc.) minus possible additional utility costs.

$$PBP = \frac{\text{Total investment}}{\text{Annual net revenue}} \tag{5.13}$$

The PBPs for each RP were calculated in two steps, first only taking energy savings into account and then including the value of the NEBs. These values were then compared to see how large effect NEBs have on the profitability of the selected energy efficiency measures.

To be able to compare the benefits of the measures after the investments have been payed back, the EBITDA of each proposal was also calculated. EBITDA denotes "earnings before interest, taxation, depreciation and amortisation", i.e. the net revenue of an investment (Law, 2016). This was made both with and without evaluated NEBs included for each RP, according to Equation 5.14. The reduced cost are here the ones for utilities and CO_2 permits and the earnings are from increased production.

$$\text{EBITDA} = \text{Reduced costs} + \text{Earnings} \tag{5.14}$$

5.4.1 Cost optimisation of RP1

As mentioned in Section 3.1, there are spatial limitations in the NHTU, which could make it better to add smaller new HXs. If HX2 was smaller, less additional area would be needed in HX1 to keep the same load as before the retrofit and the investment cost would decrease. This would however mean that less fuel also would be saved. A cost optimisation of RP1 was therefore conducted.

After calculating the different investment costs and fuel costs, an analysis of how the PBP differs based on different ΔT_{min} in the HXs was made. This was made to see if it would be better to invest in smaller new HXs and get a smaller decrease in fuel demand, from an economic point of view.

The analysis was made both with and without including the value of evaluated NEBs. Since the possible production increase is a result from replacing existing HX with new parallel plate HX, the increased production would be the same independently of the size of the new HXs. The results are shown in Section 6.4.1.

6. Results

In this chapter, the results from the thesis are presented. The results are divided into investment cost estimation, utility costs, NEBs and finally a summary and profitability analysis. Some additional results from the calculations are also presented in Appendix A and B.

6.1 Investment cost

In all retrofits there is need for new equipment. New HXs or increased heat transfer area in existing HXs (which also means new HX, since the increased heat transfer area is large enough to demand additional tube bundles) are required in all RPs. In some RPs there is also need for new pumps. The required new equipment and their estimated size can be seen in Table 6.1.

For RP1, the old HX, HX1, is suggested to be replaced by a welded plate HX, and an additional new HX, HX2, is also added. As there is an additional HX installed (including additional piping) and the old tube-and-shell exchanger is replaced by a plate HX, the pressure drop would increase. But, since there would be parallel HX installed to be able to clean during operation, there would be less impact from fouling. No additional pump work was therefore assumed to be needed.

For RP2 and 3, heat transfer area is added to the old HXs, HX1 and HX3, and completely new ones are also added. The new HXs are almost as large as the existing ones. The demand for additional pump work due to the pressure loss in the HXs was therefore assumed to be twice as large as the demand for the old ones. The pump that today drives the feed in the MHC operates close to full capacity, thus a new pump was assumed to be needed.

For RP4, one new HX is added. This new HX would have almost twice as large area as the existing one. Thus, the additional pressure drop was assumed to be twice as large as for the existing HX. Additional pump work and one new pump were therefore assumed to be needed in this RP as well. The new pump might not be necessary, as the existing one is not currently working at its capacity limit, but since the cost of a new pump would make a big impact, an estimation of the additional cost was still made.

Table 6.1: New equipment, and estimated size (heat transfer area or shaft power), for all the retrofit proposals.

RP1	Plate HX: 3x701 m^2 + 2x107 m^2
RP2	Tube HX: 419 m^2 + 524 m^2 + 304 m^2 + 385 m^2
	Pump: 17.5 kW
RP3	Tube HX: 614 m^2 + 802 m^2 + 210 m^2
	Pump: 17.5 kW
RP4	Tube HX: 2x460 m^2
	Pump: 92 kW

In Table 6.2 the investment costs for HXs, pumps and the total investment cost for all the RPs are presented. Figure 6.1 shows the part of the total investment cost that comes from HXs and pumps. The figure shows that the pump cost is a small, but significant, cost compared to the cost for the HXs in RP2 and 3. The pump cost in RP4 is however much larger and almost half of the total investment cost. Noticeable is that the cost of HXs are much more carefully calculated then the pump costs, which contains a lot of assumptions. More detailed numbers for the investment cost for each RP are presented in Appendix A.

Table 6.2: Investment cost for all retrofit proposals with cost functions from Ulrich & Vasudevan (2004).

	RP1	**RP2**	**RP3**	**RP4**
Investment cost, HX [M€]	2.66	2.14	1.96	1.11
Investment cost, Pump [M€]	-	0.22	0.22	0.85
Total investment cost [M€]	**2.66**	**2.36**	**2.18**	**1.96**

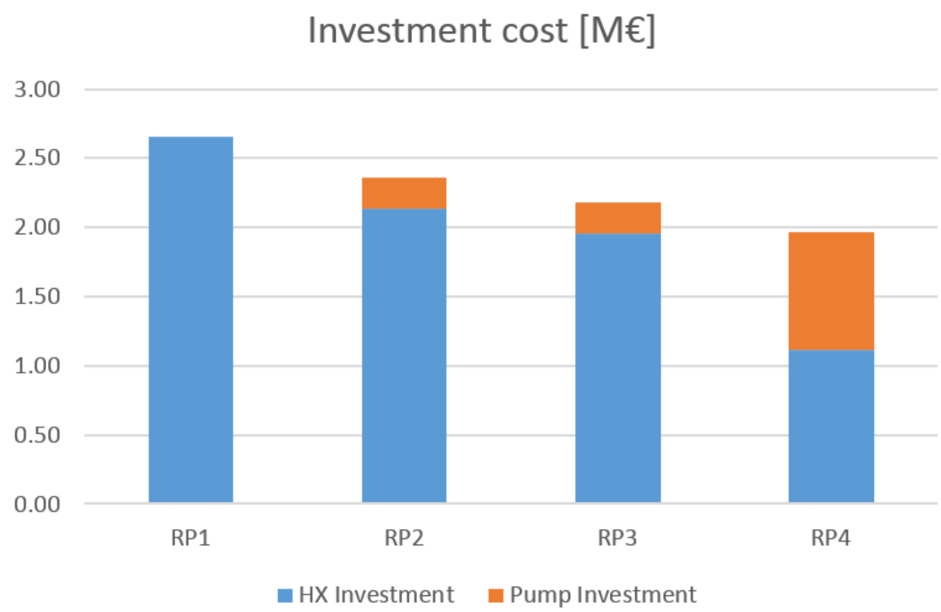


Figure 6.1: Investment cost for all retrofit proposals with cost functions from Ulrich & Vasudevan (2004).

The sensitivity analysis of investment cost for RP1, that was made for the use of different cost functions, is presented in Figure 6.2. The figure shows the investment cost for the RP, obtained using different cost functions. All results were converted into 2016 values with CEPCI, and includes factors for material, pressure and installation.

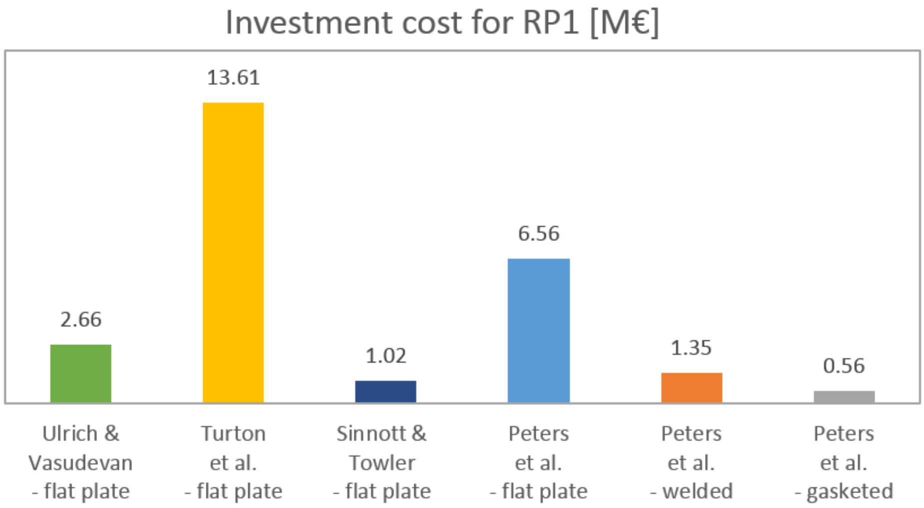


Figure 6.2: Sensitivity analysis of investment costs for RP1, based on six different cost functions from Turton et al. (2012), Ulrich &Vasudevan (2004), Sinnott & Towler (2009) and Peter et al. (2003). The values from the cost functions are converted into 2016s value and includes factors for material, pressure and installation.

6.2 Utility costs

The new configurations in the RPs would lead to lower utility costs in the process units, due to the decreased demand of fuel in the process unit. The cost savings were calculated by the decreased fuel demand over the year, times the cost of LNG. The resulting savings in fuel cost for each proposal can be found in Table 6.3.

In RP1 and 2, load on air coolers is reduced. Consequently, the electricity consumption in the air coolers will decrease. However, since the decrease in electricity demand in the air coolers likely would be small, this was neglected in the economic calculations.

The new pumps that are added was assumed to be powered by electric motors. The cost of the additional electricity demand is presented in Table 6.3. The saved fuel was assumed to be LNG 75% of the year and the rest of the time internal fuel gas (free of charge). The price of LNG was assumed to be 0.97 €cent/MJ, corresponding to 35 €/MWh, which is an average value for 2016 (excluding VAT and other recoverable taxes and levies) (eurostat, 2018). The price of electricity was assumed to be 2.88 €cent/kWh, which is an average value for 2016s spot price on NORD POOL (2018).

Table 6.3: Savings in utility cost for the retrofit proposals.

	RP1	**RP2**	**RP3**	**RP4**
ΔQ, fuel saved [MW]	6.73	9.01	11.70	8.06
Yearly savings in fuel cost [M€]	1.54	2.07	2.69	1.85
ΔE, el. demand [MW]	-	0.02	0.02	0.09
Additional annual el. cost [M€]	-	0.01	0.01	0.03
Total savings in utility cost [M€]	**1.54**	**2.06**	**2.68**	**1.82**

6.3 Evaluation of non-energy benefits

The NEBs of the RPs, and their effects on process operability, are listed in Tables 6.4-6.7. Out of the listed NEBs, the ones that are quantified are the increase in production (for RP1, 2 and 3) and the decrease of CO_2 emissions (for all RPs), which are described in Section 6.3.1 and 6.3.2. NEBs that were not quantified are discussed in Sections 7.1-7.4.

Table 6.4: Possible non-energy benefits and operability effects (denoted as +/-, meaning positive/negative effects) of retrofit proposal 1 (RP1).

Non-energy benefit	**Comment**	
Decreased CO_2 emissions	Decrease due to reduced load on furnace	
Increased production	Due to less limitation from fouling in HX1	
Reduced wear on equipment	Due to decreased load on furnace and air cooler	
Decreased noise	Due to decreased load on air cooler	
Operability factor	**Effect**	**Comment**
Flexibility	+/−	+ less fouling with parallel HX
		+ reduced load on furnace and air cooler
		− heat transfer between processes
		− increased network complexity
Controllability	+/−	+ less pressure drop with parallel HX
		+ reduced load on furnace and air cooler
		− heat transfer between different process units
		− increased network complexity
Startup/Shutdown	−	− heat transfer between processes
Reliability/Availability	+	+ less interruptions with new equipments
Practical considerations	−	− rebuilding of & installation of new equipment

Table 6.5: Possible non-energy benefits and operability effects (denoted as +/-, meaning positive/negative effects) of retrofit proposal 2 (RP2).

Non-energy benefit	**Comment**	
Decreased CO_2 emissions	Decrease due to reduced load on furnace	
Increased production	Due to de-bottlenecking	
Reduced wear on equipment	Due to decreased load on furnace and air cooler	
Decreased noise	Due to decreased load on air cooler	
Operability factor	**Effect**	**Comment**
Flexibility	+/−	+ de-bottlenecking + reduced load on furnace and air cooler − increased network complexity − pressure drop
Controllability	+/−	+ reduced load on furnace and air cooler − increased network complexity − pressure drop
Startup/Shutdown	none	
Reliability/Availability	−	− rebuilding of & installation of new equipment − high pressures and large pressure difference
Practical considerations	−	− rebuilding of & installation of new equipment − high pressures and large pressure difference

Table 6.6: Possible non-energy benefits and operability effects (denoted as +/-, meaning positive/negative effects) of retrofit proposal 3 (RP3).

Non-energy benefit	**Comment**	
Decreased CO_2 emissions	Decrease due to reduced load on furnace	
Increased production	Due to de-bottlenecking	
Reduced wear on equipment	Due to decreased load on furnace	
Operability factor	**Effect**	**Comment**
Flexibility	+/−	+ de-bottlenecking + reduced load on furnace − increased network complexity − pressure drop +/− change in steam balance
Controllability	+/−	+ reduced load on furnace and air cooler − pressure drop +/− change in steam balance
Startup/Shutdown	none	
Reliability/Availability	−	− rebuilding of & installation of new equipment − high pressures and large pressure difference
Practical considerations	−	− rebuilding of & installation of new equipment − high pressures and large pressure difference

Table 6.7: Possible non-energy benefits and operability effects (denoted as +/-, meaning positive/negative effects) of retrofit proposal 4 (RP4).

Non-energy benefit	**Comment**	
Decreased CO_2 emissions	Decrease due to reduced load on furnace	
Improved product quality	Due to de-bottlenecking	
Reduced wear on equipment	Due to decreased load on furnace	
Operability factor	**Effect**	**Comment**
Flexibility	+/−	+ de-bottlenecking
		+ reduced load on furnace
		− pressure drop
		− change in steam balance
Controllability	+/−	+ reduced load on furnace
		− pressure drop
		− change in steam balance
Startup/Shutdown	none	
Reliability/Availability	−	− rebuilding of & installation of new equipment
		− high pressures and large pressure difference
Practical considerations	−	− rebuilding of & installation of new equipment
		− high pressures and large pressure difference

6.3.1 Increased production

The value of increased production was evaluated for RP1, 2 and 3. RP4 does not lead to increased production but could affect the flexibility and achieve higher yield of more valuable products. This flexibility was not evaluated, but it is discussed in Section 7.4.

For RP1, only the gasoline production increases, and for RP2 and 3 both gasoline and diesel production were assumed to increase. In RP2 and 3, half of the increased production was assumed to be gasoline and half diesel. The total production was assumed to increase by an average of 3% during 50% of the operating time for RP1 and by 5% during 20% of the operating time for RP2 and 3, leading to average production increases of 1.5% and 1% respectively, for the different proposals. The annual increase in cash flow from the increased production is presented in Table 6.8.

Table 6.8: Yearly increased earnings for increased production. Gross margins for gasoline and diesel are 0.15 €/l and 0.11 €/l, respectively.

	RP1	**RP2**	**RP3**
Increased production [% of operating time]	1.5	1	1
Increased earnings from gasoline production [M€/year]	5.25	1.75	1.75
Increased earnings from diesel production [M€/year]	-	1.32	1.32
Total increased earnings [M€/year]	**5.25**	**3.07**	**3.07**

A sensitivity analysis regarding the value of increased production for RP2 and 3 was conducted to see how it could vary depending on the assumed production increase. The result from the sensitivity analysis are presented in Figure 6.3. The annual value of increased production could differ significantly from the assumed value, as can be seen in the figure.

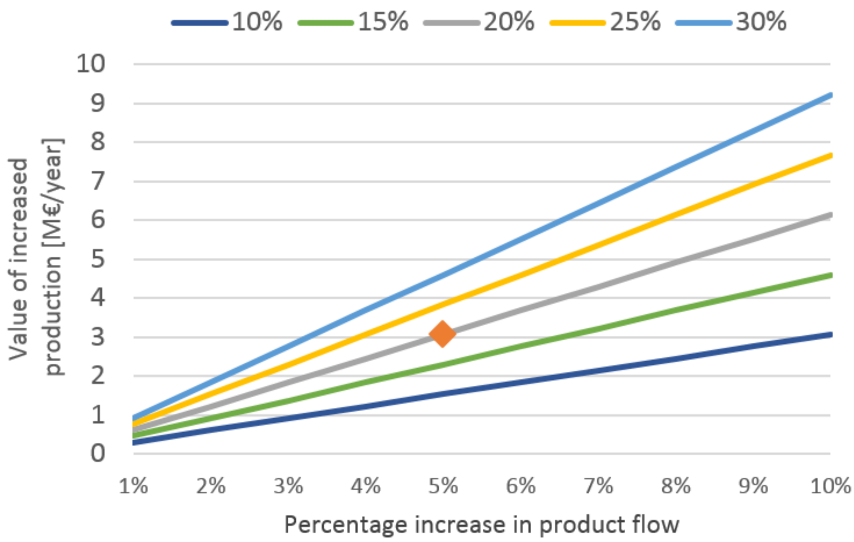


Figure 6.3: Sensibility analysis for increased production. The different lines represent how much of the time the production is increasing. The orange diamond represent the assumed value for RP2 and 3.

6.3.2 CO_2 emissions

The decrease of CO_2 emissions only occurs 75% of the year, since the other 25% of the year the refinery has more fuel gas in their system, and will need to burn the fuel gas that is saved in the furnaces. Table 6.9 shows the resulting annual cost savings related to the reduced emissions.

Table 6.9: Decrease in CO_2 emissions and savings in cost for emission permits for the retrofit proposals.

	RP1	**RP2**	**RP3**	**RP4**
Decrease in CO_2-eq emissions [kton/year]	11.10	15.07	19.70	13.66
CO_2-eq for burning of LNG [g/MJ]	71.8			
Cost of emissions [€/tonCO_2-eq]	5.5			
Savings [M€/year]	**0.062**	**0.083**	**0.108**	**0.075**

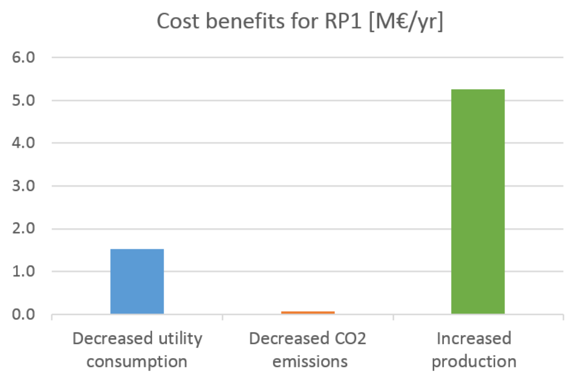
The reduced CO_2 emissions were calculated with fuel demand including the increased production for RP1, 2 and 3. The savings in cost, presented in Table 6.9, from the decrease of CO_2 emissions, assumes that the company could sell emission permits equal to their reduced emissions of CO_2-eq.

6.4 Summary and profitability

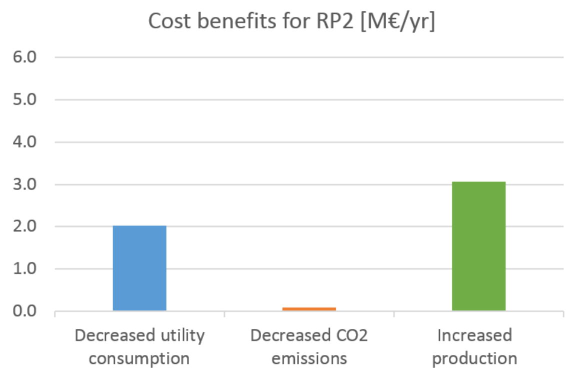
All savings in utility costs and costs of CO_2 emissions, and the earnings of increased production, are summarised in Table 6.10. Figure 6.4 shows a graphic comparison of how big the values are, for each retrofit proposal. As can be seen in both the table and the figure, increased production has the largest effect in all the RPs where it is included. The smallest effect comes from the savings in cost from decreased need for CO_2 emission permits.

Table 6.10: Summary of cost savings and earnings for all retrofit proposals.

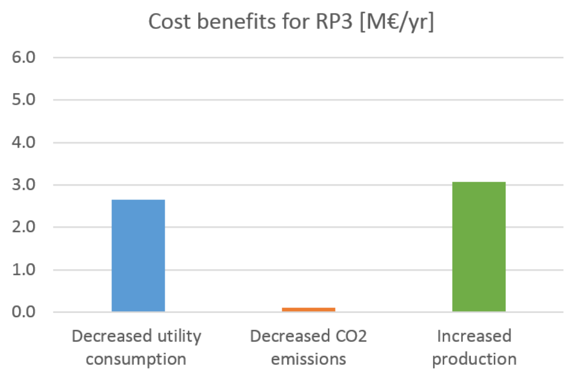
	RP1	**RP2**	**RP3**	**RP4**
Savings in utility cost [M€/year]	1.52	2.03	2.66	1.82
Savings from decreased CO_2 emissions [M€/year]	0.06	0.08	0.11	0.08
Increased earnings from gasoline production [M€/year]	5.25	1.75	1.75	-
Increased earnings from diesel production [M€/year]	-	1.32	1.32	-
Total savings/increased earnings [M€/year]	**6.84**	**5.18**	**5.83**	**1.90**



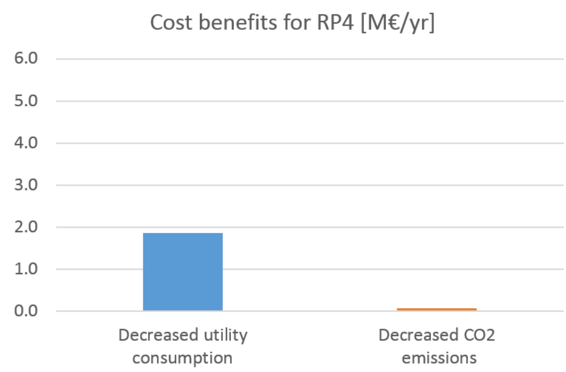
(a) Comparison of cost benefits for RP1



(b) Comparison of cost benefits for RP2



(c) Comparison of cost benefits for RP3



(d) Comparison of cost benefits for RP4

Figure 6.4: Comparison of the annual value of non-energy benefits and utility savings for each retrofit proposal.

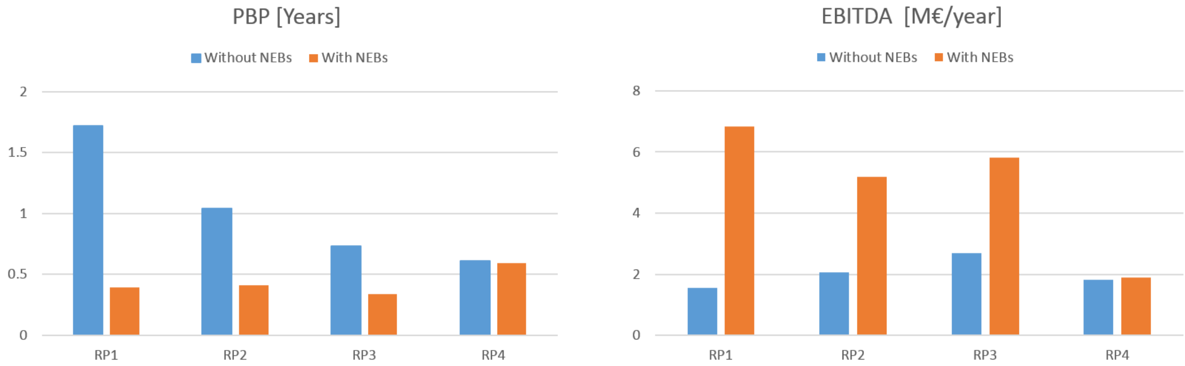
The profitability of each RP, shown as PBP and EBITDA (the net revenue) is presented in Table 6.11 and 6.12. The evaluated NEBs have, as can be seen in the tables, a great impact on both of the profitability measures. Visualisations of the results are also shown in Figure 6.5.

Table 6.11: Payback period in years, for the different retrofit proposals.

	RP1	**RP2**	**RP3**	**RP4**
PBP without NEBs	1.72	1.04	0.73	0.61
PBP with NEBs	0.39	0.41	0.34	0.59

Table 6.12: EBITDA, the annual net revenue, in M€/year, for the different retrofit proposals.

	RP1	**RP2**	**RP3**	**RP4**
Without NEBs	1.54	2.06	2.68	1.82
With NEBs	6.84	5.18	5.83	1.90



(a) Payback period in years.

(b) EBITDA, the annual net revenue, in M€/year.

Figure 6.5: Profitability of each retrofit proposal, shown as PBP and EBITDA (the net revenue). The orange and the blue bars represent the value with and without NEBs, respectively.

6.4.1 Cost optimisation of RP1

The results of the cost optimisation of RP1 is shown in Figure 6.6. As the figure shows, the PBP is shortest for the lowest ΔT_{min} for the case when only utility savings are included in the profit. For the case where evaluated NEBs (emissions and production increase) were included, the PBP decreases slightly with higher ΔT_{min}. Note that the value of the production increase is the same for all temperature differences, which causes this case to decrease with decreasing investment cost and to not vary as much as the one without NEBs. The values of the cost optimisation are presented in Appendix B.

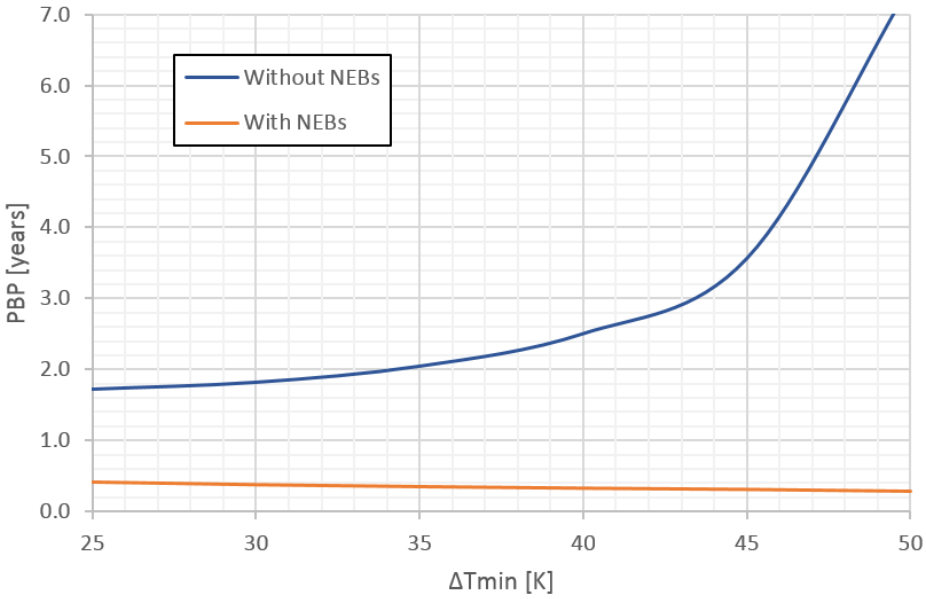


Figure 6.6: Cost optimisation of RP1 based on minimum temperature difference in the heat exchangers. Shown as PBP as a function of ΔT_{min}. ΔT_{min}= 25 K represents the value used in RP1.

7. Discussion

In this section, each retrofit proposal (RP) is first discussed and then a general discussion about method and results is presented. The features that are specific for each RP are presented in their separate sections, while effects and assumptions that concern all proposals are discussed in Section 7.5.

7.1 Retrofit proposal 1

Some issues with RP1, that were mentioned in Section 3.1, are the spatial limitations and the fact that there are two different process units involved in this proposal. The old HX, HX1, would not be an issue as it is replaced with compact plate HXs that could take up less space than the old tube-and-shell HX. For the new HX, HX2, that uses a hot stream from the CRU, there might be more problems. As it would need piping from two different process units, including a bypass, the cost of HX2 might be a bit higher than estimated. The estimation of reduced utility cost might also be affected by the interaction with another unit. As the preheating depends on the stream from the CRU, less fuel would for example be saved if there is a stop in the CRU but not in the NHTU.

One operability factor that only concerns this RP is the problem with start-up and shut-down of the processes. Since there are two process units, and they are as mentioned not always operated at the same time, this might cause problems in the equipment, especially during start-ups. This was not taken into consideration in any calculations in this thesis, but it might be a problem that can affect the feasibility of this RP. However, the issues of combining streams from two different units should not be very large in this case, as these two units are often operated at the same time.

No new pump was assumed to be needed in RP1, since there probably will be no increased pressure drop. However, a pump could be a good investment, because when HX1 is replaced, the feed will not be limited by it any more. This means that an additional pump could make it possible to increase the flow even further. Instead of a 3% increase during 50% of the operating time, as assumed in this thesis, the flow might be able to be increased with a lot more. This could be evaluated further if more calculations were done.

When the demand for cooling of the hot stream from the CRU is reduced, the load on the air cooler will be smaller. Especially during summers, when it is warmer outside, the air cooler might be limiting for the process as it cannot cool the stream enough. This leads to an excess of fuel gas in the system, that at some points cannot be consumed which leads to a decrease in possible production. There is a possibility that a decreased load on this air cooler might bring a possible increase in production, as it would limit the production less during the summer. This could also be evaluated further if more calculations are done.

One of the NEBs that could follow this RP, that has not been evaluated in this thesis, is the decreased noise from the air cooler when the load is lower. As there are regulations of how much noise the refinery can cause, less noise from an air cooler might lead to that it is possible to add noisy equipment in another place in the process. Furthermore, the work environment around the air cooler would be better with less noise.

There was a cost optimisation made for this RP, to see if an investment in smaller new HXs would bring higher profitability. As can be seen in Figure 6.6, the PBP would be longer with a smaller investment as there would be less savings in utility costs. When the NEBs are included, the high benefit of the production increase would make it slightly more profitable with a smaller investment. However, as this makes such a small difference, RP1 is considered most profitable with the HX sizes that were proposed in the previous work.

7.2 Retrofit proposal 2

Spatial limitations have not been taken into account for RP2. Since this RP also include installation of four new/expanded HXs in the process unit, spatial limitations are likely to have a large impact on the potential to implement this retrofit. The spatial limitations might cause additional costs from rebuilding the process unit, to provide space for new equipment. Rebuilding of the process unit would add additional costs that have not been taken into account in this thesis.

There are pressure limitations in the pipes in the process units today. This means that even if the additional pump that has been taken into account is installed, the pipes and other process equipment might not be able to handle the higher pressure. If so, some pipes in the process unit would have to be replaced with new ones that could handle higher pressure, which would add extra cost.

This RP includes decreased load on air coolers, similar to the case in RP1. This could lead to decreased noise from the process unit which result in better work environment and/or the possibility to install additional noisy equipment in other places in the process.

Another thing that has not been accounted for is the decrease in flexibility due to higher network complexity. If the flow rate of the streams that are used to preheat the feed into the MHC are too small, the desired temperatures will not be reached. Then the furnace would have to heat that feed extra to reach the desired temperature into the reactors. In that case the energy savings would not be as high and more fuel would be used. It is also possible that the stream used to preheat the feed would be cooled too much, which could cause problems downstream in the process unit. These problems will not necessary occur when operating with the configuration from RP2, but in some cases the increased complexity can lead to these problems.

Regarding the increased production in this RP, it was assumed that only the production of gasoline and diesel would be increased and that the proportion between them is 50/50. The production could however affect other products and the proportion between the increased products may be different. The size of the total increased production is also an uncertainty. It is possible that the production could increase even more and there is also a possibility that the production cannot be increased as much, as better catalysts could be used in the future. With better catalysts the production could be kept at a higher level during more time of the year.

7.3 Retrofit proposal 3

RP3 is similar to RP2, which means that problems regarding spatial limitations and pressure limitations would also be similar. In this RP only three new/expanded HXs would be installed. However, these HXs would be larger than the ones in RP2. This means that the HXs in RP3 would take up as much space as in RP2, which could complicate installation of new equipment. The increased production for RP3 has the same uncertainties as RP2 and only a simplified mix with gasoline and diesel was accounted for.

There is often a surplus of LP steam in the refinery in the current configuration and operating conditions. It is this excess LP steam that is used to preheat the feed into the MHC. However, there is not always a surplus of LP steam and when there is a shortage more LP steam would have to be produced. The increased production of LP steam leads to an increase of fuel consumption in the steam boilers. In this thesis, a scenario where some additional fuel is needed was investigated, but it could be both more or less additional fuel demand during other operating scenarios, leading to either higher or lower utility cost.

7.4 Retrofit proposal 4

For the estimation of investment costs for RP4, u-tube HXs with a high pressure factor were considered. Since the pressures in the ICR are high, special made high pressure HXs would be needed, which means that the actual cost might be even higher than the level considered in this thesis. The same goes for the pump, even though high pressure was considered in the evaluation, there might be additional cost factors as all piping etc. needs to be able to handle the high pressures. As in all the other RPs, there are also spatial limitations, which can complicate installation of new equipment.

For the utility cost, a scenario where the fuel and electricity balances are unaffected was considered. A different scenario might lead to increased cost of fuel in the steam boilers to make up for the lost steam, or increased cost of electricity to make up for lost turbine work with work from electric motors. The main scenario that was considered, is from an operating point where the production is stable and normal, so the assumption is still valid. However, there are scenarios that affect the electricity and fuel use, and if further calculations were done it would be good to consider how the utility costs would be affected by other scenarios.

The value of the evaluated NEBs of RP4 differs a lot from the other proposals, as it does not include any increased production. The reduced energy demand in the furnace would bring an increased flexibility, as the furnace capacity would no longer limit the potential for heating the process stream. The increased available capacity could be used to achieve a higher temperature at the inlet of the separation unit. This could give either higher yields of diesel and lighter products, which are more valuable than the heavier UCO, or increased purity of the products. Additional firing in the furnace to reach higher temperatures in the separation would of course lead to less savings in fuel costs. The additional fuel demand would then have to be subtracted from the reduced utility costs. It is not clear what the value of this increased flexibility could be, as there are other limitations affecting the desired production in this separation as well. For example, purer products from the separation might not be profitable since they are mixed with other product streams before becoming final products. If further evaluation was made, the value of flexibility and possibility to receive more valuable or purer products would be an interesting benefit to look at.

7.5 General discussion

The sensitivity analysis of the investment costs for RP1 shows that the choice of cost function makes a big difference in the cost estimations. The analysis was only made for one RP, but it indicates that there is probably a difference even when other types of equipment are concerned. Since a larger or smaller investment cost would affect the PBP as well, the profitability results would also be affected by this. The difference that the inclusion of NEBs makes for the profitability would also change, but the effect would still point in the same direction.

The cost estimations for pumps include a lot of assumptions about how the pressure drop would increase due to the new equipment. This means that the real increase in demand for pump work can differ significantly from these estimations. Also, if a new pump would be installed, a larger capacity than to only cover for the additional pressure drops in the new HXs would probably be invested in. More detailed calculation of the pressure drops could be made if there are future studies.

When calculating the savings in fuel cost and the decreased CO_2 emissions, the fuel was assumed to be LNG. However, the marginal fuel that is used in the refinery could vary, and other fuel than LNG could be used. If other fuels are used, the savings in utility cost and the decrease of CO_2 emissions could be different.

In the calculations of value of increased production, the parts of the process units that are mentioned in this thesis were considered to be the only limitations in the production. No consideration was given to other limiting equipment in the units, or in the rest of the process. If there are other limitations, that would mean that the production cannot be increased as much as estimated.

One thing that has not been taken into account in the calculations of this thesis is the cost of stopping the production for the installation of new equipment. The refinery has large planned stops for larger maintenance and new installations/rebuilding of the process approximately every sixth year. If the RPs can be fit in to one of these planned stops, there would be no additional loss in production due to the installation. However, if additional stops are needed, this would then decrease the value of the increased production.

Regarding the calculations for savings from CO_2 emissions, it was assumed that the company would sell their emission permits. The company could however decide to keep the permits and use them another year (as long as they do not exceeds the limitation of banking in the ETS), when they have too few, instead of buying more then. Either way, the company would profit from the decrease in CO_2 emissions. In this thesis the effects from decreased CO_2 emissions are quite small. This NEB could however give a larger effect if the emission permits were more expensive, which they might be in the future.

Some of the NEBs that have not been evaluated in this thesis are: reduced wear on equipment and operability issues. Lower load on furnaces and air coolers would reduce wear on the equipment and could bring higher flexibility and controllability to the process units. The reduced wear on equipment could be accounted for in different ways, such as less need for maintenance or longer life time of the equipment. Less interruptions due to maintenance will also increase the reliability of the process units.

Changes in the steam network can be both positive and negative for the operability. If there is a large surplus of LP steam in the refinery, and that steam can be used for heating process streams, the flexibility and controllability of the processes can increase. However, if the use of LP steam leads to a very small surplus of LP steam or even a shortage, this would have negative effects on flexibility and controllability.

When looking at the profitability, NEBs shorten the PBP and increases the net revenue for all the RPs. This shows that including of NEBs in the profitability evaluation matters for all the energy efficiency measures that have been evaluated. Even though there are some uncertainties due to assumptions made in the calculations, and the fact that price of fuel and process equipment can vary, the results in this thesis all point in that same direction.

Besides the monetary value of the evaluated NEBs, pointing out the value of NEBs can also give a higher strategic value of the RPs. Except for the NEBs that have been quantified, the other mentioned benefits can also bring an increased strategic value of the measures. This could be of use when proposing investments in energy efficiency measures.

8. Conclusions

Four different RPs for heat exchanger networks at Preemraff Lysekil have been evaluated in this thesis. The RPs were all constructed in previous work, with the aim to increase energy efficiency and also add other effects to the process. Investment cost, utility costs and earnings from selected NEBs have been estimated for all the four RPs. Several possible NEBs were found in the evaluation, but only two were selected to be quantified. These selected NEBs were reduced CO_2 emissions and increased production. Including the value of NEBs in the evaluation of the RPs gave more then 50% shorter PBPs for the three RPs were increased production was included.

The results of this thesis show that NEBs of energy efficiency measures can bring both higher profitability and possibly add more strategic values. In three out of four evaluated retrofit proposals, the value of quantified NEBs is larger than the value of savings in utility costs. If more NEBs with different characteristics were evaluated, such as improved product quality in RP4 or reduced wear on equipment in all RPs, the profitability might increase even further.

Possibly, the including of NEBs in evaluation of profitability, could lead to more investments in energy efficiency measures and thereby more energy efficient processes in industry. However, further studies are needed to verify how large impact NEBs can make on investments in energy efficiency.

9. Future studies

If further studies of the subject are made, it would be interesting to look into some additional aspects that have been mentioned in this thesis. As mentioned in the discussion, some of the NEBs that have not been evaluated economically in this thesis could be interesting to investigate further.

For RP1 the production increase could be even larger since the reduced load on the air cooler possibly removes limitations during summertime. The production could also be increased a lot more if a new pump were to be installed in the process unit. Both of these possible effects could be investigated in further work.

For RP2 and 3 there are limitations in the process unit due to pressure and spatial limitations, which could make these proposals difficult to implement. Adjustments in these RPs might therefore be needed to make them more feasible, and the change in profitability of the RPs would then be interesting to evaluate. For these RPs it would also be interesting to make additional investigation of which products would be affected and how much the production increases.

For both RP3 and 4 the steam network is affected, and in this thesis only one scenario of the steam network model is used to estimate the effects. If further evaluation is made, it would be interesting to use more scenarios to see how the fuel and electricity balances would change during different operating conditions.

For RP4 it would be interesting to evaluate the value of the increased flexibility due to the increased available capacity in the furnace. Both the possible increase of production of the lighter, more valuable, products and possible increased purity of the product streams could be evaluated.

Furthermore, additional RPs could be evaluated to see the effects from more NEBs with different characteristics. The same type of study could also be carried out in other refineries or other types of industries. It could then be evaluated whether the value of NEBs would be as high even in other types of proposals and processes.

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A. Calculations - Investment cost

Table A.1: Results of estimation of heat transfer areas and investment cost for *retrofit proposal 1*. Cost functions from Ulrich & Vasudevan (2004)

	U-value [W/m^{2}K]	Area [m^2]	Purchased Cost base case [k€]	Purchased Cost incl. f_M & f_P [k€]	Investment cost, incl. installation [k€]
HX1	800	3x701	240.0	607.2	2 283.1
HX2	800	2x107	39.4	99.7	374.8
Total					**2.66 M€**

Table A.2: Results of estimation of heat transfer areas and investment cost for *retrofit proposal 2*. Cost functions from Ulrich & Vasudevan (2004)

	U-value [W/m^{2}K]	Area [m^2]	Purchased Cost base case [k€]	Purchased Cost incl. f_M & f_P [k€]	Investment cost, incl. installation [k€]
HX1	350	419	41.1	148.0	556.3
HX3	350	524	48.4	174.2	655.1
HX5	350	304	32.7	107.9	405.7
HX6	350	385	38.6	139.0	522.5
Pump		17.5 kW	13.4	61.1	220.6
Total					**2.36 M€**

Table A.3: Results of estimation of heat transfer areas and investment cost for *retrofit proposal 3*. Cost functions from Ulrich & Vasudevan (2004)

	U-value [W/m^{2}K]	Area [m^2]	Purchased Cost base case [k€]	Purchased Cost incl. f_M & f_P [k€]	Investment cost, incl. installation [k€]
HX1	350	614	54.5	196.2	737.7
HX3	350	802	67.0	241.2	906.9
HX5	600	210	25.4	83.8	315.2
Pump		17.5 kW	13.4	61.1	220.6
Total					**2.18 M€**

Table A.4: Results of estimation of heat transfer areas and investment cost for *retrofit proposal 4*. Cost functions from Ulrich & Vasudevan (2004)

	U-value [W/m^{2}K]	Area [m^2]	Purchased Cost base case [k€]	Purchased Cost incl. f_M & f_P [k€]	Investment cost, incl. installation [k€]
HX1	350	2x460	88.0	295.7	1 111.8
Pump		92 kW	27.6	236.0	851.9
Total					**1.96 M€**

B. Cost optimisation of RP1

Table B.1: Results of estimation of heat transfer areas, investment cost and utility cost for the cost optimisation of RP1. The case with ΔT_{min}=25 °C represents RP1 as it is presented in Section 3.1.

ΔT_{min} [°C]	Areas [m^2] HX1	HX2	Investment cost [M€]	Utility cost Fuel saved [MW]	Savings [M€/year]
25	701 x3	107 x2	2.66	6.71	1.54
30	595 x3	81 x2	2.32	5.53	1.27
35	518 x3	59 x2	2.05	4.36	1.00
40	458 x3	41 x2	1.84	3.19	0.73
45	411 x3	24 x2	1.65	2.02	0.46
50	371 x3	10 x2	1.44	0.84	0.19

As in RP1, the value of increased production is calculated to 1.94 M€, and the PBP with and without NEBs is shown is Table B.2.

Table B.2: Payback periods with and without NEBs for RP1, for different values of ΔT_{min}.

ΔT_{min} [°C]	Payback period [year] without NEBs	with NEBs
25	1.73	0.42
30	1.82	0.38
35	2.05	0.35
40	2.52	0.33
45	3.56	0.31
50	7.42	0.29