

Diploma Thesis

Process Investigation with Focus on the Energy Optimization Potential for Mekanissa Alcohol Factory

Addis Ababa, Ethiopia

Composed for

**Institute for Process Technology & Industrial
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My sincere gratitude belongs to all those people who supported my sense of curiosity without reservation and enabled this research work in Ethiopia. The list of names would be too long as mental, financial and academic back up came from many different sides in many different ways. Only the interplay of all the assistance I was granted made this challenge an extraordinary and instructive experience.

Special thanks also to those friends and members of my family who taught me to care and reason. And to those who always encouraged me to take risks and follow distinctive ways.



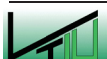
Kurzfassung

Prozessuntersuchung in Hinblick auf das Energieoptimierungspotential für „Mekanissa Alcohol Factory“ (MAF)

Die vorliegende Diplomarbeit ist Teil eines gesamten „Cleaner Production“ Projektes, welches von einer für Äthiopien entwickelten UNIDO Strategie geleitet ist. Diese Strategie hat zum Ziel, Äthiopiens Abhängigkeit von der Landwirtschaft in Richtung wettbewerbsfähiger Industriestandorte und Dienstleistungen zu lenken. Das Gemeinschaftsprojekt zwischen der „National Alcohol and Liquor Production of Ethiopia“, einem staatlichen Betrieb, und dem Institut für Chemische Verfahrenstechnik der Universität Addis Ababa versucht nun, diese Strategie in einem Pioneerprojekt umzusetzen.

Die einleitende Prozessbeschreibung gibt einen grundlegenden Überblick über die herrschenden Prozess- und Arbeitsbedingungen, um Folgeprojekte im Bereich der Prozessoptimierung entwickeln zu können. Der Hauptteil dieser Arbeit konzentriert sich auf den Energieverbrauch des Betriebes. Massen- und Energiebilanzen wurden aufgestellt als Grundlage für die Identifizierung des Energierückgewinnungspotentials innerhalb des bestehenden Systems. Die erarbeitete Lösung hält sich an die Vorgabe, den bestehenden Produktionsprozess zur Ethanolherstellung nicht zu beeinflussen. Mit Hilfe der Pinch Analyse sollte ein Konzept entwickelt werden, dass darauf basiert lediglich das bestehende Wärmetauschernetzwerk zu optimieren. Im Zuge der Problembearbeitung stellte sich jedoch heraus, dass es sich nicht um ein gepinches Problem handelt. Die entwickelten Szenarien beruhen daher lediglich auf der Vorgabe, das bestehende Netzwerk so wenig als möglich zu verändern. Eine abschließende wirtschaftliche Evaluierung gibt Auskunft, wie sinnvoll die Umsetzung des Vorschlages in finanzieller Hinsicht ist.

Die Ergebnisse zeigen, dass bei einer Investition von 27 569 EUR jährliche Einsparungen von 16 550 EUR erzielt werden können, bzw. bei der Neuinstallation eines größeren Wärmetauschers (37 505 EUR Investition) jährliche Einsparungen in der Höhe von 19 995 EUR möglich sind. Beide Varianten entsprechen einer Amortisationszeit von weniger als 2 Jahren und stellen damit eine attraktive Energierückgewinnungsmaßnahme dar.



Abstract

Process Investigation with focus on the Energy Optimization Potential for “Mekanissa Alcohol Factory” (MAF)

The presented diploma thesis is part of an overall cleaner production project initiated by an UNIDO strategy. This strategy, which aims to shift Ethiopia's dependence on the agriculture towards competitive industries and services, is now implemented by a joint project between the „National Alcohol and Liquor Production of Ethiopia“, a governmental owned company, and the Department for Chemical Engineering from the University of Addis Ababa.

The initial process description gives a basic insight into present working and process conditions of the fermentation and distillation sector to enable the identification for follow-up optimization projects. The main focus in this work is led on the energetic side of the ethanol distillation. Mass- and energy balances have been generated. Based on these data, the possible heat recovery potential to reduce the utility cost of the process within the existing system was identified. The solution proposed obeys the constraint to achieve energy savings without process modifications that will have impact on the ethanol production itself. Pinch Analysis was supposed to be applied to propose adaptations in the heat exchanger network. When investigating the retrofit system however it happened to be not a pinched problem and the recommended scenarios have been developed following a general guideline of interfering the existing network in the minimum possible way. A final economical assessment is evaluating the feasibility of implementing the designed solution.

The result shows that for an investment of 27 569 EUR annual savings of 16 550 EUR can be achieved, or for the instalment of a bigger heat exchanger (investment 37 505 EUR) annual savings up to 19 995 EUR are feasible. Both alternatives have payback times less than 2 years and are therefore attractive energy recovery measures.



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1 Preface

Ethiopia has developed a long-term strategy “Agricultural-Development-Led-Industrialization” to shift the agricultural dependence in the economical sector towards competitive industries and services. The whole scope of the strategy involves agro-industries, the development of quality management, standardization and certification for industrial competitiveness, investment and technology promotion, cleaner production practices, industrial energy efficiency and renewable energy development. An “Integrated Program for Private Sector Development, Industrial Competitiveness and Environmentally Friendly Production” [1] has singled out the main components that are relevant for Ethiopia and are now carried out in cooperation with industries and research institutions.

1.1 Tasks and objectives

The present diploma thesis “ Process Investigation with focus on the Energy Optimization Potential for Mekanissa Alcohol Factory, Addis Ababa (Ethiopia)”, aims to accomplish the targets of the component “Industrial Energy Efficiency and Renewable Energy Development” in a selected distillation plant. It represents the initial study for a long-term cleaner production project undertaken at the company. As part of the prevailing work a basic process description (including the fermentation part and distillation process itself) should stress the current working conditions and situation. It should be outlined which parts of the plant are working properly and where negligence in maintenance and investment has created the process as it is today.

The impacts of these process conditions will be investigated in later projects and will not be subject of further discussions in this work.

The focus of this study is exclusively based on the decrease of the current fuel oil consumption for the steam generation and the identification of the energy recovery potential within the heat exchanger network based on the preliminary process investigation. An awareness creation among the management in charge should be achieved by considering also the economic feasibility of an energy utilization concept.

Among the different approaches for developing an energy saving concept, the problem stated at Mekanissa Alcohol Factory (MAF) was supposed to be investigated by applying the concept of Pinch Technology. A heat exchanger network should demonstrate the maximal possible heat recovery within the examined system.



2 Cleaner Production

2.1 Definition of Cleaner Production

UNEP [2] has defined Cleaner Production as:

“... the continuous application of an integrated preventive environmental strategy to processes, products, and services to increase overall efficiency, and reduce risks to humans and the environment. Cleaner Production can be applied to the processes used in any industry, to products themselves and to various services provided in society. For production processes, Cleaner Production results from one or a combination of conserving raw materials, water and energy; eliminating toxic and dangerous raw materials; and reducing the quantity and toxicity of all emissions and wastes at source during the production process.”

Basically it is not a scientific or legal definition but more a strategy to run industrial processes in a way that minimum negative environmental impact will be caused by applying presently available technologies and considering economical constraints of the company at the same time.

The measures for Cleaner Production can in short be based on good house keeping, which includes:

- Conscious use of resources
 - Keeping record of material and energy flows
 - Drawing up material and energy balances
 - Evaluating the balances (losses, waste, changes of flows, benchmarking)
 - Identifying potential for reutilization and recycling (material and thermal)
- Avoiding of resource wasting
 - Technical reduction of losses from material and energy
 - Organizational measures and awareness building to save material and energy
- Implementation of technologies with high material and energy efficiency
 - Product change
 - Process change
 - Improvement of process control
- Internal recycling
 - Material recycling
 - Energy / heat recovery
 - Improvement of process control



Continued from previous page:

- Recycling of secondary resources
 - o Utilization of secondary resources
 - o Utilization of down-cycled – resources
 - o External processing and utilization of material
- Replacement of resources and auxiliary material
 - o Less hazardous / poisonous materials (e.g. detergents and solvents)
 - o Less ecological harmful materials
- Renewable materials and energy sources
- Technology changes
 - o Processes with higher material and energy efficiency
 - o Processes with less waste water / off gas / waste generation

2.2 Cleaner Production at Mekanissa Alcohol Factory

For the studied company MAF, the cleaner production notion has developed mainly out of the consciousness of the practiced resource wasting. So far no measures have been taken to implement and realize any of the listed steps. The declared aim is now to optimize the whole process starting from improved fermentation and distillation control in order to increase output and quality, up to a more energy efficient and less waste generating ethanol production.

The term “optimization” itself implies not just the improvement of a given system with existing technical, economical and legal constraints, the objective is moreover to achieve the highest possible technical efficiency within the prevailing limitations. To design the best system under given conditions, therefore requires the definition of the system-boundaries and the influencing factors (environment, economics, laws and other regulations etc.) that create a demand for optimization projects. Hence the “best solution” has to balance these external impacts and respond in an appropriate manner to the defined “best system” (e.g. lowest sum of capital investment, operation and maintenance costs).

Optimization of processes can have two different approaches: end of pipe measures are basically cleaning technologies at the end of a process, while an integrated environmental protection includes environmental considerations during the product development and the process design, to prevent the production of waste streams (solid, liquid, gaseous) or recycle and reuse materials and energy within the system as far as possible.



3 Process Integration

3.1 Definition of Process Integration

Process Integration, in the context of energy optimization, is defined by the International Energy Agency (IEA) [3] as followed:

"Process Integration is the common term used for the application of methodologies developed for system-oriented and integrated approaches to industrial process plant design for both new and retrofit applications.

Such methodologies can be mathematical, thermodynamic and economic models, methods and techniques. Examples of these methods include: Artificial Intelligence (AI), Hierarchical Analysis, Pinch Analysis and Mathematical Programming.

Process Integration refers to optimal design; examples of aspects are: capital investment, energy efficiency, emissions, operability, flexibility, controllability, safety and yields. Process Integration also refers to some aspects of operation and maintenance".

The more general definition used in this context is the one used by the IEA since 1993 [3]:

"Systematic and General Methods for Designing Integrated Production Systems, ranging from Individual Processes to Total Sites, with special emphasis on the Efficient Use of Energy and reducing Environmental Effects".

3.2 Introduction to Process Integration

Out of the definition of "Process Integration" it becomes obvious that design, operation and management of industrial processes with system-oriented models and methods are covered areas in the energy optimization approach.

Process Integration includes a variety of methodologies. A possible classification of the Process Integration methods is given in figure 1 [4]:



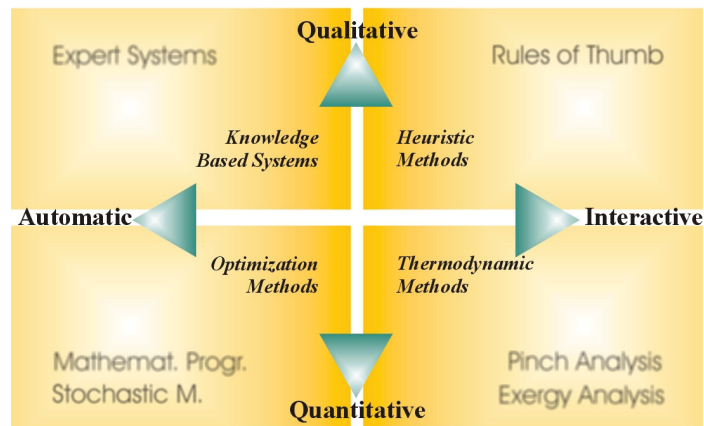


figure 1: Process Integration Methodologies

The main methodologies are pinch analysis for heat recovery, heat and power systems and wastewater minimization; exergy analysis; optimization techniques such as mathematical programming and stochastic search methods and the use of heuristics and knowledge based systems [5].

While until recently, process integration was only associated with energy integration, meanwhile a wider range of application has been created by combining thermodynamics, the advantages of mathematical programming and optimization for the improvement of overall production performances, reducing operating and controllability problems, increasing plant flexibility and minimizing undesirable emissions in all industrial sectors.

3.3 Energy related investigation of the distillation process

Directing the focus on distillation, it is well known that it is a very energy-intensive process and with rising energy costs it is more and more demanded to implement energy recovery systems in the process in order to keep expenses for energy as low as possible and produce in an economic manner. Especially since the energy crisis in the 1970s and 1980s “Process Integration” became a common design and analysis tool in the chemical industries, where energy optimization techniques started to develop.

3.4 Measurement of existing energy performance

Before any modification of the process can be realized to improve the process energy performance, it is indispensable to illustrate the energy demands and utilities of the process.

To examine the energy flows, consumptions and losses of a process, different systems can be specified to formulate energy as well as mass balances. The system (figure 2) in general has to be defined by the system boundaries that determine which part of the process is assessed and which parts are not considered. In- and output streams of material and energy



are entering and leaving the system across the boundaries and can be separated, united and transformed within the investigated system. The enclosed system can be single processes, a whole plant site or a product.

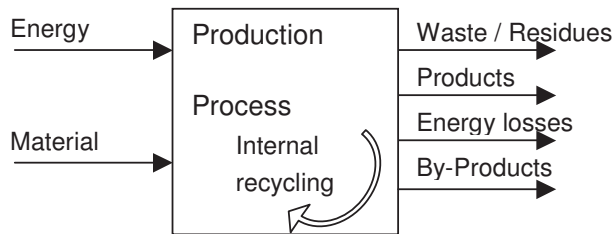


figure 2: system with boundaries and streams

Hence an adequate description of the technological process, which gives details about the studied system, defines the system, its boundaries, components and potential subsystems. Furthermore the physical properties of the surrounding (such as pressure and temperature) that have impact on the defined system and the conditions, approximations and taken assumptions complete the initial description of the system.

3.5 Energy audit of existing process

The energy audit is usually performed in two complementary ways:

- Review of plant design

Design information for the process

- Process flow sheets
- Design calculations
- Piping and instrumentation drawings
- Specifications of the equipment purchased
- Performance characteristics of the equipment
- Utility usage tabulations (design for fuel, steam and electrical usage)
- Revisions since the original installation
- Process calculations (calculations for specific process conditions)

- Audit of actual plant performance

The theoretical design data give a picture about the original conception of the plant but does not display any changes or adaptations of the process and might not meet today's requirements (changing of product quality; product range; utility, raw material, disposal costs etc.), leave alone it might give information about the actual plant performance. Based on the



background information about the original plant design it is therefore also essential to determine the present energy usage and investigate the process as it is working now.

A reliable description of the actual situation however can only be achieved if the parameters are recorded frequently and a stable process control enables constant measuring conditions and procedures.

Once this preparation work is accomplished, the optimization task can be started.



4 Thermodynamic process integration methodologies

Within the process integration methodologies two are mainly used to identify energy saving potentials: Exergy analysis and Pinch Technology.

4.1 Exergy Analysis

Exergy analysis is based on the second law of thermodynamics and is used to analyze and improve chemical processes. Exergy analysis states that the traditional process analysis, based on mass and energy balances, is not suitable as a performance indicator, as only external process losses can be detected. Especially the comparison between different energy forms (e.g. chemical and electrical) is not feasible. If the system to be investigated is simply analyzed based on the first law of thermodynamics, only the total amount of energy that is put into and will be rejected at another stage of the process will be compared, but the different qualities of energy are not taken into account. How much actual work could have been extracted from the energy source remains unknown.

Exergy analysis (also described as availability or available useful work) allows measuring the quality of energy and material streams in order to quantify internal process losses (due to fluid flow, mass and heat transfer or chemical reactions).

Wall 1977 [6] defined exergy as the work that can be extracted when a system is brought into equilibrium with its surroundings without incurring any losses.

The exergy losses can have two origins: either it is caused by irreversibility (exergy destruction) or it is simply the unused exergy share (exergy waste). Due to irreversibility exergy can by definition never be in balance ($\Delta S^{tot} > 0$) as the total exergy input always exceeds the exergy output ($E_{in}^{tot} > E_{out}^{tot}$). The exergy destruction is related to entropy generation by:

$$\Delta E = T_0 \Delta S_{tot} = E_{in}^{tot} - E_{out}^{tot} = \sum_i \Delta E_i \quad (1)$$

T_0 reference temperature

ΔS_{tot} total entropy increase

E_{in}^{tot} total input exergy

E_{out}^{tot} total output exergy

ΔE_i exergy destruction in process i

For real processes (irreversible) the non-utilized output (exergy waste) can be determined:

$$E_{waste} = E_{out}^{tot} - E_{product} \quad (2)$$



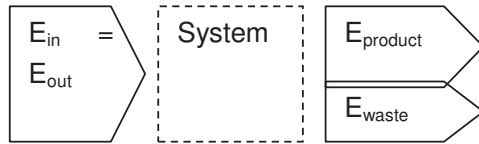


figure 3: Sankey diagram for exergy flow

Such kind of Sankey diagram (figure 3) can be generated for all process streams to visualize the exergy losses. These single exergy losses calculated for each zone have to be added up to obtain the total amount of exergy loss of the system.

A change in availability equals the amount of shaft work that can be extracted from a system flowing from an initial to a final stage [7]. This makes it possible to define the minimum work, which is necessary to achieve this change in a flowing system. The available shaft work can therefore be defined as follows:

$$-W_{shaft} = \Delta A = \Delta H - T_a \Delta S \quad (3)$$

- A availability
- H enthalpy
- T_a reference temperature (298 K)
- S entropy

The higher the value for the availability is, the higher is the work that can be extracted.

The exergy concept is often used to identify energy losses and is mainly preferred to find the best feed condition for a given separation which operates under a wide range of feed conditions. Pinch Analysis, the second widely used tool for Process Integration, is more commonly used to investigate the process streams and forms the linking of hot and cold streams in a thermodynamic optimized way, aiming to achieve the maximum possible heat recovery.

The aim of the retrofit project at Mekanissa Alcohol Factory is, to utilize preferably the existing streams for heat recovery instead of changing the layout of the plant itself. This stream matching approach, together with an accompanying evaluation of the energy savings within the heat exchanger network is offered by the Pinch Methodology.

4.2 Pinch Analysis

The Heat Recovery Pinch was identified in the 70's by Hohmann (1971), Umeda (1977-1979) and Linnhoff (1977-79) [14] who saw it as the key to process design of energy intensive processes and as answer to the then-current energy crisis. It was seen as a major



breakthrough in the development of analysis and design methods for energy efficient processes and industrial sites [8].

The original energy optimization approach was developed further over the years and Pinch Analysis now finds its application also in areas like emission targeting or water and wastewater minimization etc. User-friendly and reliable software (about 50 computer programs ranging from high standard commercial products to prototype university software [4]) have been developed as well, which allows advanced Process Integration design. The software chosen for this work is *Hint* [9], as it was available free of charge and offers all the necessary tools for a basic analysis.

Applying the Pinch method allows identifying critical elements within the heat recovery network but requires a different way of planning when approaching the problem. Instead of following a serial structured design concept (figure 4), where every element is performed independently [10], it is necessary to consider the interactions between the single design steps right from the beginning as shown in figure 5.



figure 4: serial network design

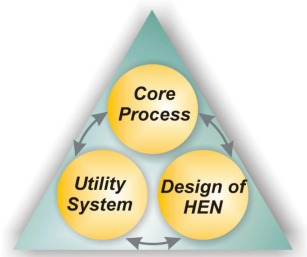


figure 5: interactive network design

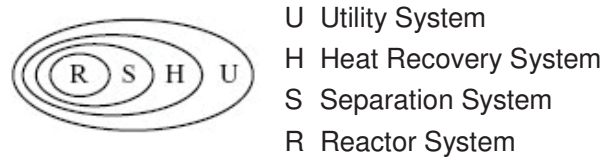


figure 6: Onion Model

The resulting conceptual design is often described with the Onion Model (figure 6). In the middle the reactor is located that specifies input-, output- and recycle- concentrations and flow rates. Based on this information the second layer of the onion model – the separators - can be designed and basic information for energy and material balances are given for the planning of the heat exchanger network (represented through the third layer). The outer layer stands for the remaining heating and cooling requirements that have to be covered by the utility system. The Pinch Analysis begins with the heat and material balance for the process and works towards inner and outer layers, by identifying energy influencing potentials in the reactor process conditions and setting targets for energy saving and utility loads at certain levels prior to the design of the heat exchanger network.



5 Pinch Technology

5.1 Introduction

Energy efficiency should be obtained with minimum capital expenditure and maximum possible energy recovery within a system to keep the annual operating costs low. To reduce the costs of utilities (e.g. steam, cooling water), heat recovery facilities have to be exploited to the maximum possible extent and higher cost utility usage (e.g. high pressure steam) should be shifted to lower cost utility usage (e.g. low pressure steam) by obeying thermodynamical principles and economical feasibility. This task can be achieved by increasing the process to process heat recovery and reducing the process – utility heat exchange. To give an example: if a process stream has to be heated and it can be achieved through using another process stream instead of using steam or electricity, the recovered heat equals savings in external energy supply. This utilization of the energy content of other process streams at the same time requires heat exchanger units that have to be implemented. This investment will only be justified in economical terms, if the cost savings in energy will exceed the capital investment for heat exchangers and additional piping and control system within a certain time period. The proper placement of heating and cooling units (exchangers) within the process therefore needs a systematic methodology, as Pinch Analysis offers it.

The optimization of heat recovery networks is based on the first and second law of thermodynamics.

The first law of thermodynamic (law of energy conservation) states that energy is conserved. It can be transformed into different forms of energy and transported but it cannot disappear.

$$dU = dQ + dW \quad (4)$$

dU total increase of internal energy U of the system

dQ total heat supplied to the system

dW total external energy increase of the system or the total work done by the system

The second thermodynamic law is also known as the law of energy degradation. The entropy change of any system and its surroundings (resulting from any real process) is positive and comes near a limiting value of zero for any process that approaches reversibility [11].

$$dS = \frac{dQ_{rev}}{T} \quad (5)$$

dS change of Entropy

Q_{rev} heat supplied for reversible processes

T temperature



Entropy in an isolated system must therefore either increase or stay constant, which means heat can only be transferred from higher levels of energy to lower levels.

5.2 Application of Pinch Technology

In general Pinch Analysis should be considered right from the beginning when designing a process (single units or total sites), but it can and has been used to a large extent in industries to establish the potential for energy savings in existing plants. The improvement of an existing plant (retrofit) however will lead to a different optimal heat exchanger network than the original design would have done, as certain process modifications might not be feasible on the site. Due to given practical limits, for the revamping of existing plants, as in the case of Mekanissa Alcohol Factory, the target is to manufacture the desired products with the minimum amount of energy expenditure as well as minimum additional capital cost for extra necessary heat exchanger, utility supply equipment and network.

To get an idea about Pinch Analysis and the underlying principles and methods, a brief outline should summarize the theory behind.

5.3 Basic tools and principles

5.3.1 Definitions

For a clear understanding of the used terminology, the most important expressions shall be defined beforehand.

System [11]

“A system is taken to be any object, any quantity of matter, any region etc. selected for study to set apart (mentally) from everything else, which is then called surroundings. The imaginary envelope, which encloses the system and separates it from its surroundings, is called the boundary of the system. The boundary can have two aims: to isolate the system and no interaction with the environment are feasible any more or to see the interactions between the system and its surroundings.”

Hot Stream

The term “hot stream” is used for process flows that need to be cooled and can transfer heat to streams with lower energy level.

Cold Stream

“Cold streams” describe process streams that accept heat transfer from higher energy levels.



Utility (hot and cold)

Utility includes all kind of external energy supply to fulfil the required heating and cooling demands for the process.

5.3.2 Methodology

Pinch Analysis can be divided into four phases:

1. Data Extraction
Process, stream and utility data for the specified system
2. Targeting
best performance: maximal heat recovery, minimal utility usage
3. Design
creating the Heat Exchanger Network (HEN)
4. Optimization
adaptation of best HEN with consideration of economical, safety, control and geographical distance restrictions

and includes following steps [12]:

1. Identification of hot, cold and utility streams in the process
2. Thermal data extraction for process and utility-streams
3. Selection of initial ΔT_{\min} value
4. Construction of composite curves and grand composite curve
5. Estimation of minimum energy cost target
6. Estimation of Heat Exchanger Network capital cost targets
7. Estimation of optimal ΔT_{\min} value
8. Estimation of practical targets for Heat Exchanger Network design
9. Design of Heat Exchanger Network

The sequence of this hierarchy does not display single steps that can be treated separately but require an iterative procedure that takes corrections based on acquired results.



5.3.3 Data Extraction

5.3.3.1 Data Extraction Flowsheet

A very crucial part in identifying the energy saving potential within a process and to achieve the best final design requires a good understanding of the process and a complete representation of the streams and utilities through energy and material balances as well as the identification of process heating, cooling, evaporation and condensation duties for the investigated system.

Data can be gained by following methods:

- Measurements (are often not complete and not reliable)
- Design data (are often outdated after plant modifications)
- Simulation models (may not always reflect true plant behaviour)

5.3.3.2 Data Collection Purpose

To describe the streams that shall be integrated in the heat exchanger network, process conditions and properties of the single liquid and vapour flows have to be determined. It is desirable to measure the actual temperatures, pressures and flow-rates of the streams instead of running simulations and computing with theoretical data in order to get a picture of the real situation.

The advantage of this actual process data collection is to enable an optimal practical application of the designed Heat Exchanger Network for the existing process. This is especially important for plants that were installed a few decades ago without major investment since then, but many minor adjustments due to break down of single units, as it is the case for Mekanissa Alcohol Factory.

On the other hand the disadvantage of data-inaccuracy has to be considered and additional estimations of data reliability are necessary.

The data dependability for the specific case of Mekanissa Alcohol Factory is due to manual process control rather vague and the measured data can not be taken as absolute values and therefore should possibly be compared with simulation programs or literature values. Hence the basis for the collection work is a general doubt about the collected values, as instruments are often broken, flow meters might measure flows for several usages and the flow split can only be guesstimated etc. Such data have to be treated with care and can not be taken as correct straight-ahead, as this could give a wrong computation basis for the design of the Heat Exchanger Network and the proposed energy conservation projects are doomed to fail. The mass and energy balances for the single units of the process as well as for the overall process have to be generated to get a complete matching description of the examined system.



5.3.3.3 Necessary Data collection

To obtain a valuable and meaningful process flowsheet, it basically suffices to have numeric data for solving energy and material balances. These contain temperatures, flow-rates and enthalpy changes of every stream.

Thermal Data

The required data for a Pinch Analysis include:

T_s	[°C]	Supply temperature
T_t	[°C]	Target temperature
c_p	[kJ/°C kg]	Specific heat capacity
\dot{m}	[kg/sec]	Mass flow-rate
ΔH	[kJ/kg]	enthalpy change
		$H = Q + W$ (first law of thermodynamics)
		Mechanical work W does not occur in heat exchangers:
		$H = Q = m \cdot c_p \cdot (T_s - T_t) = CP (T_s - T_t)$
Q	[kJ/kg]	heat supply or demand
ΔH_{vap}	[kJ/kg]	heat of vaporization for streams with a phase change
CP	[W/°C]	heat capacity flow rate
		whereas $CP = m \cdot c_p$,
		representing the specific heat capacity of the stream
ΔT_{min}	[°C]	Specified minimum temperature difference for heat exchange

The resulting flowsheet is called data extraction flowsheet.

Additional Data

With regard to further design and implementation of heat exchangers it is also desirable to collect data about physical properties of the streams, as heat transfer coefficients have to be estimated.

To get a complete energy consumption balance it is also indispensable to have information about the energy consumption of existing condensers and reboilers in the process. Furthermore it is also advantageous to have some further insight into the process and gather so called “non-numeric-data”. This qualitative data should give information about:

- The maximal tolerable variation of process parameters (will pressure or temperature changes influence the separation process?)
- Which restrictions have to be faced regarding safety (heat exchange between hydrocarbon streams and oxygen rich streams), controllability, flexibility, operability (start-up and shut down), product quality (potential leakage in heat exchangers) or geographical distance (costly piping)?



- In order to distil vapour and liquids, the pressure must be kept below the critical pressure
- Thermally unstable components may limit the reboiler temperature
- Feed changes require a high degree of operational flexibility

In general it is recommended to ignore those restrictions and include all streams in the Pinch Analysis at the beginning. This will have the advantage to get a full range of designing freedom and limitations can be established at any time with respect to identified limitations.

As for the case of Mekanissa Alcohol Factory it has to be mentioned, that for the time being no process modifications are subject of the investigation. The task to be accomplished is exclusively based on the given system conditions and the energetic optimization potential within these limitations, without influencing the distillation process in quality or quantity.

5.3.4 Tools of Pinch Technology

5.3.4.1 Stream table and composite curves

The gathered information will be transferred to the stream table (table 1), which summarizes the relevant information to describe the streams and to create composite curves.

table 1: stream table

Stream –Nr.	Type	T_{start} [°C]	T_{target} [°C]	\dot{m} [kg/h]	c_p [kJ/kg. °C]	CP [kJ/h. °C]
S1	Hot	150	60	0.5	0.4	2
S2	Hot	90	60	4	2	8
S3	Cold	20	125	5	0.5	2.5
S4	Cold	25	100	6	5	3.0

The data presented here have only illustration purpose and do not represent actual data from Mekanissa Alcohol Factory.



Construction of Composite Curves

Based on these thermal data the profiles of heat availability („hot composition curve“) and heat demand („cold composition curve“) of the process are plotted in a graphical temperature-enthalpy diagram (T-H diagram). To plot the curves in the T-H diagram, all enthalpies of the hot streams within a respective temperature interval are added for the hot composite curve and equivalent procedure will lead to the cold composite curve. For a detailed explanation of how this temperature-enthalpy diagram is constructed it shall be referred to literature [4] or [8] at this point.

The desired final diagram used for the Pinch Analysis is showing the “Composite Curves” (figure 7) that combine all hot streams in one hot composite curve and all cold streams in another cold composite curve. When the hot stream (red line) and cold stream (blue line) are represented in one diagram, following observations can be made:

- Heat can only be transferred from the hot stream to the cold stream at or above the specified minimum temperature difference in the interval where hot and cold streams are superimposing. This 'overlap' indicates the scope for heat recovery.
- Where the hot line exceeds the cold line heat must be rejected to a cold utility.
- Where the cold line exceeds the hot line the use of hot utility becomes evident to cover the heating demand.

Most temperature-enthalpy plots also display a narrow point between the hot and cold curves at a minimum specified temperature difference that is known as “Pinch Point” (figure 8), dividing the process into “a region above the pinch” and a “region below the pinch”. Above the pinch heat is taken in and no heat is rejected and is therefore known as heat “sink”, while the part below the pinch serves as heat “source”.

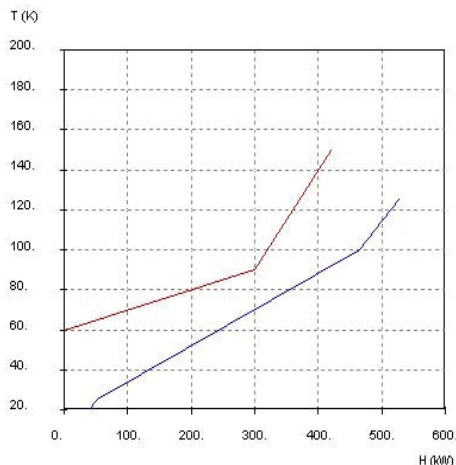


figure 7: Composite Curves

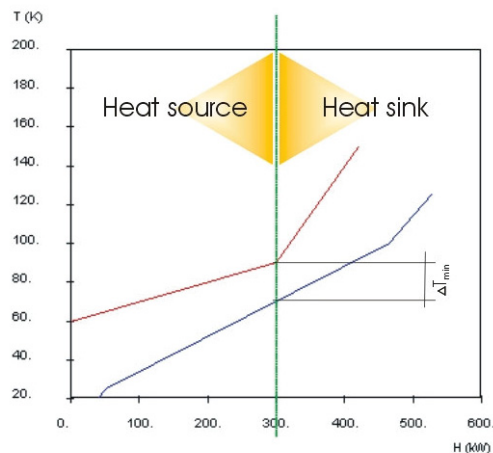


figure 8: Pinch Impact

5.3.4.2 Pinch Location

The Pinch is acting as a bottleneck where the driving forces for heat transfer become limiting. The scope for heat recovery is therefore obviously dependent on the allowable temperature approach. When the minimum temperature increases, the lines move apart, the overlap reduces and the utility demands increase.

This choice of ΔT_{\min} has therefore great influence on getting the best solution for a heat exchanger network. If ΔT_{\min} is decreased the heat recovery potential will increase and therefore the demand for external heating and cooling utilities will be minimized. At the same time the driving forces for heat exchange are getting smaller and will lead to a bigger exchange area, which will result in higher expenses for investment.

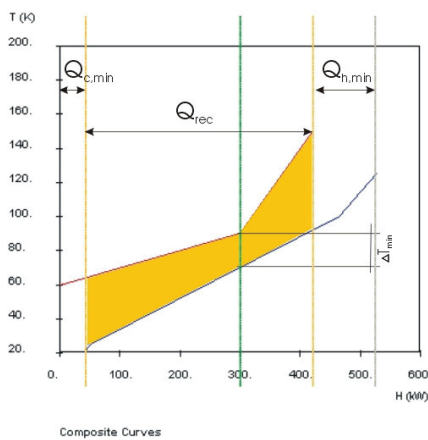


figure 9: ΔT_{\min} small

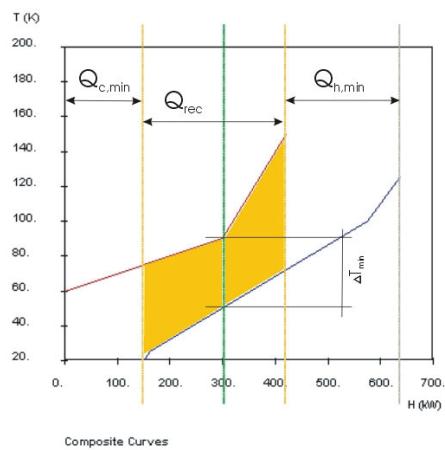


figure 10: ΔT_{\min} increased

The importance of an appropriate choice of ΔT_{\min} becomes obvious when comparing figure 9 and figure 10. To determine the optimum ΔT_{\min} different approaches exist:

- Global (experience) values for ΔT_{\min}
- Energy and capital costs as a function of ΔT_{\min}
- Consideration of individual heat transfer conditions for each process stream

Details on the determination of retrofit starting values for ΔT_{\min} are given in the targeting chapter.



5.3.4.3 Pinch impact

The occurrence of the pinch is dividing the process into two problem regions that have to be solved separately. To explain why this differentiation is necessary, the three golden rules of Pinch Methodology will show the impact of heat transfer between these two regions.

- Heat must not be transferred across the Pinch
- There must be no external cooling in the heat sink region
- There must be no external heating in the heat source region

External heating in the heat source area, instead of exploiting the existing potential for heat recovery will also lead to increased external cooling demand, as the stream that potentially could be used for heating purpose, would be cooled at the same time and would not require equivalent external cooling afterwards anymore. Neglecting this potential requires therefore external facilities for heating AND cooling. The equivalent situation occurs, if external cooling is used in the heat sink region or heat is transferred across the Pinch.

Any heat integration MUST take this fundamental Pinch decomposition into account by matching heat sources at sufficiently high temperatures with heat sinks at lower temperatures. Heat integration that does not obey this simple rule will not result in any energy saving, thus it will be a waste of capital investment. Hence the division of the process streams into a heat sink and a heat source region provides general guidelines for process modifications summarized as “Plus-Minus-Principle” (figure 11). This strategy follows following rules:

Increase the available heat above the Pinch
 Increase the heat demand below the Pinch
 Decrease the heat demand above the Pinch
 Decrease the available heat below the Pinch

The range to take influence in the existing process according to the “Plus-Minus-Principle” is very wide and can be for example: changing the pressure drop in distillation columns, changing flow-rates, setting new target temperatures etc.

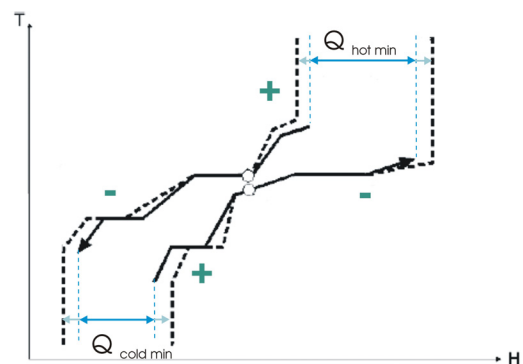


figure 11: Plus-Minus-Principle

5.3.4.4 The Pinch Principle

Characteristic for the Pinch Analysis is that the problem is not approached from the absolute hot end to the cold side but starts at each side from the pinch and is developing the Heat Exchanger Network away from the pinch location to avoid any unnecessary heating or cooling input above the minimum target.



5.3.4.5 Grand Composite Curves

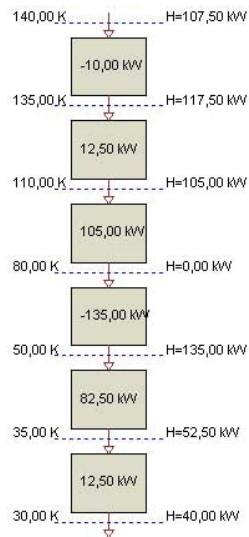
For the sake of completion another tool within Pinch Analysis that enables the identification of the potential for shifting high energy utility supply (e.g. high pressure steam) to low energy utility supply (e.g. low pressure steam) in regions where costly energy supply is not needed, should be mentioned at this point. To set these multiple utility targets, Grand Composite Curves serve as a tool, which allows spotting those intervals that demand lower energy utility supply.

The overall Q_h and Q_c have still the same value but cost optimization can be gained through introducing cheaper utility loads. This tool would help to obtain the maximal operational cost savings but has no relevance for Mekanissa Alcohol Factory as only one hot and one cold utility supply are used.

5.3.4.6 Energy Cascades

Composite curves represent the graphical solution for identification of pinch location and heat recovery potential. However for practical applications numerical methods are preferable. Based on the stream table that contains the extracted process-data, so called “Energy Cascades” (figure 12) can be generated to formulate the equivalent mathematical solution.

The algorithm starts like the procedure for constructing the Composite Curves with the division of the temperature scale into temperature intervals, in which the enthalpies are accumulated. The accumulated enthalpies for each interval are calculated as follows:



$$\Delta H_i = (T_i - T_{i+1}) * (\sum CP_c - CP_h) \tag{6}$$

i		index for temperature interval
ΔH_i	[MW]	net heat requirement in interval i
CP_c	[MW/°C]	mass specific heat of a cold stream
CP_h	[MW/°C]	mass specific heat of a hot stream
T	[°C]	temperature

figure 12: Energy Cascades



Assuming a complete heat exchange between hot and cold streams involves that there can either be a heat surplus in the respective temperature interval or a heat deficit but not both at the same time. Logically the cascading principle is also founded on the feasibility of transferring energy from higher to lower levels (second thermodynamic law). At this point it shall be referred to literature [4] or [8] again for a detailed explanation of how to compose the cascades.

As a result, after correcting the cascade where only surpluses will occur, the cascade stage with the value of zero indicates the Pinch and the first and last values in the heat cascade column are identical with the results for Q_c and Q_h from the grand composite curve.

5.3.5 Result of data collection and preparation

Heat cascading and composite curve construction show the maximal possible heat recovery potential and give a target for minimum Q_c and Q_h utility input. The aim of optimal Heat Exchanger Network design is to reach these targets through proper matching of streams.

Furthermore the location of the Pinch, which divides the streams of the process into two regions (heat source and heat sink) is identified and sets the starting point for the network design.

Minimum network costs however are not only achieved by optimal energy recovery but also need to consider the appropriate number and size (heat exchanger area) of implemented units. Installation of more than the minimum required units would lead to an increase of equipment, piping, control and maintenance costs.

Hence a controlled reduction of units (exchanger, heating and cooling equipment) needs a thorough optimization between all the different influencing parameters, which makes the targeting step a crucial point before starting with the development of the network.

5.3.6 Targeting

The Heat Recovery Pinch is the centre for the calculations of the minimum heat exchanger network costs, which consists of energy consumption (external utility supply), amount of process and utility exchangers (units) and overall heat transfer area.

5.3.6.1 Energy Target

As already shown, the minimum energy target for the process can be achieved by shifting the hot and cold composite curves, so that they are only separated by the given minimum temperature difference ΔT_{\min} . Based on this overlap the maximum possible heat recovery and the remaining heating and cooling needs can be indicated. The chosen ΔT_{\min} determines



therefore the minimum hot utility requirement ($Q_{H,\min}$) and the minimum cold utility requirement ($Q_{C,\min}$) for the evaluated process.

5.3.6.2 Number of Units

The estimation of the fewest number of units is derived from Euler's Rule of Graph Theory:

$$U = N + L - S \quad (7)$$

U	number of units
N	number of streams and utilities
L	number of heat loops
S	number of (sub-) networks

Heat loops in a network occur, when heat loads are unnecessarily shifted around in a loop. Through proper design it is possible to break these loops, so that in a first step $L = 0$ can be assumed.

Sub-networks occur when a hot stream matches perfectly with a cold stream, which is hardly ever the case. The number of (sub-) networks therefore can be taken as 1.

Hence the final targeting equation for the minimum number of units (U_{\min}) is given with:

$$U_{\min} = N + 0 - 1$$

$$U_{\min} = N - 1$$

As the Pinch principle divides the process into two thermodynamically independent regions that have to be designed separately, the minimum unit targeting equation has to be applied for both areas. If the process has not only one process pinch, but also intermediate utility pinches, the overall equation for the minimum number of units is generalized as:

$$U_{\min} = \sum_{i=1}^{n+1} (N_i - 1) \quad (8)$$

n	number of pinches
n+1	number of areas divided by pinches

Comparing the total number of U_{\min} for each region (figure 13), with the minimum unit requirement for the whole process (figure 14) a contradiction, which asks for the ignorance of the Pinch, seems to appear:



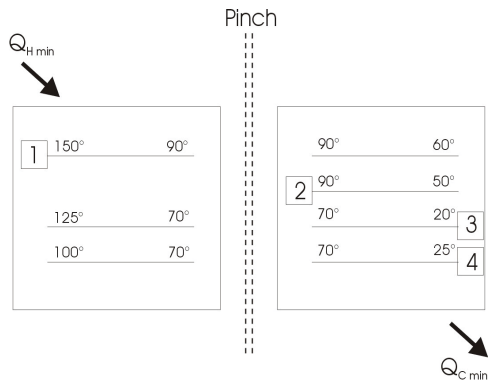


figure 13: Pinch consideration

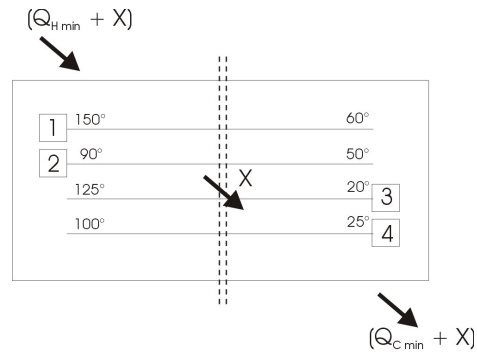


figure 14: heat transfer across the Pinch

$$U_{\min \text{ Pinch}} = U_{\min \text{ HOT end}} + U_{\min \text{ COLD end}}$$

$$U_{\min \text{ Pinch}} = [(3+1)-1] + [(4+1)-1]$$

$$U_{\min \text{ Pinch}} = 7$$

$$U_{\min \text{ total}} = [(4+2)-1]$$

$$U_{\min \text{ total}} = 5$$

Obviously a correlation between maximum energy recovery and minimal number of units occurs when targeting the Heat Exchanger Network. To achieve the overall U_{\min} , two solutions are possible:

1. The Pinch Rule, stating that no heat must be transferred across the Pinch, has to be violated, which will lead to increased heating and cooling demand, resulting in a cost trade off between units (capital costs) and energy (operating costs).
2. ΔT_{\min} has to be changed which will influence the surface area and maximal energy recovery, leading to a cost trade off between ΔT_{\min} and heat exchanger area.

5.3.6.3 Area Targets

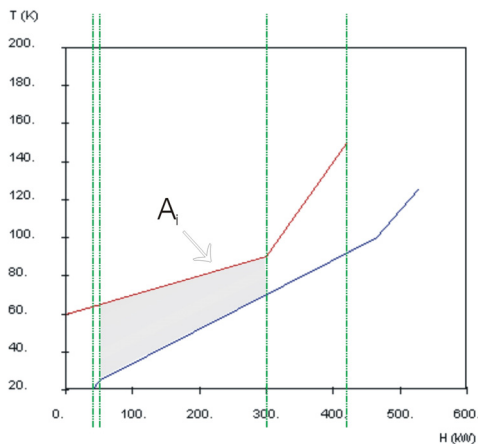
For the area targeting it is assumed that for each temperature interval a theoretical heat exchanger can be placed. The composite curve is divided into temperature intervals in a way that hot and cold composition curve do not change the slope in the respective interval

(figure 15).

In each interval a vertical heat transfer (counter-current flow) is assumed. The driving force for each hypothetical heat exchanger depends on the ΔT . For each of these intervals the



area of the theoretical pure counter-current flow heat exchangers can be calculated with following equation:



$$A_i = \frac{\Delta H_i}{U \Delta T_{LMi}} \quad (9)$$

A_i	[m ²]	Respective heat exchanger area in the interval i
ΔH_i	[W]	interval enthalpy (total heat transferred)
U	[W m ² K]	global heat transfer coefficient
ΔT_{LMi}	[K]	log. mean temperature difference

figure 15: area targeting

With the logarithmic mean temperature difference, for a counter-current-flow heat exchanger with entering (i) and leaving stream temperatures (o) of the cold (c) and hot (h) medium:

$$\Delta T_{LM} = \frac{(T_{h,i} - T_{c,o}) - (T_{h,o} - T_{c,i})}{\ln \left(\frac{(T_{h,i} - T_{c,o})}{(T_{h,o} - T_{c,i})} \right)} \quad (10)$$

This vertical arrangement equals a pure counter-current area within the overall network and gives only a target value of the minimal required area. In reality of course the aim can only be to achieve this value as far as possible.

5.3.6.4 Cost targeting for heat exchanger area

The basic heat exchanger cost equation is

$$\text{Cost} = a + b \cdot A^c \quad (11)$$

a, b, c constants in heat exchanger cost law (dependent on type and material of heat exchanger)

A area

5.3.6.5 Cost targeting for Heat Exchanger Network

The minimal Heat Exchanger Network cost targeting combines target figures for the minimum number of units and the area estimation:



$$C_{\min, HEN} = N_{\min} \left[a + b * \left(\frac{A_{\min}}{N_{\min}} \right)^c \right] \quad (12)$$

$C_{\min, HEN}$	min. HEN costs
N_{\min}	min. number of units
A_{\min}	min. area

5.3.6.6 Total Energy Cost (operational costs)

$$C = \sum_{U=1}^U Q_u * C_u \quad (13)$$

Q_u	[kW]	duty of utility U
C_u	[\$/kW, yr]	unit cost of utility U
U		total number of utilities used

5.3.6.7 Total Annual Cost

By combining all these single targeting figures and considering additional economic factors such as payback time or interest rate and operating hours, the total annual costs can be estimated. Still it has to be kept in mind that there are several uncertainties, like for instance unknown distribution of the area to the single heat exchangers, which will have impact on the final result.

5.3.7 Retrofit Targeting

Targeting for an existing process has to consider further design impacts, as the optimum of different modification alternatives and their influence on the existing process is more difficult to assess than a grassroots design.

For revamp energy optimization projects limitations have to be tolerated that avoid the exploitation of maximum heat recovery:

- Layout and structure of existing plant
- Installed heat exchanger equipment

These constraints require a slightly different targeting approach for retrofit projects, as it is usually preferred to achieve best energy recovery facilities, while keeping structural changes as low as possible.

To keep investment costs as low as possible, the integration of existing units into the newly designed network has to be considered, which has an important impact on targeting the minimum exchange area.



To compare the ideal heat exchanger network with the one from the existing plant, the energy-area plot (figure 16) for heat exchanger networks illustrates the potential and errors that can occur in optimizing an existing HEN.

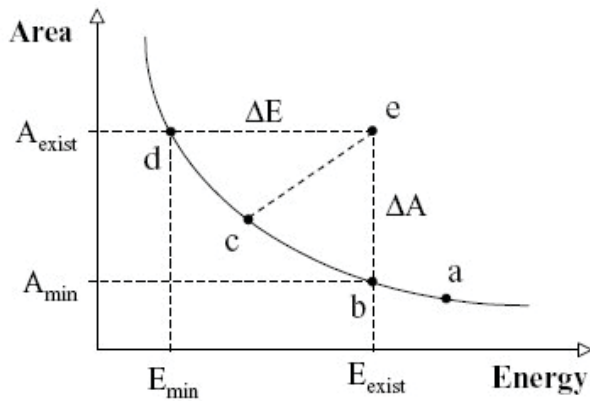


figure 16: energy-area plot

The existing energy consumption and the already installed heat exchanger network are represented at point e. The curve is plotted after calculating minimum target values for energy and area with different levels of ΔT_{\min} and shows the minimum heat exchanger area for a certain energy demand. To achieve the maximum heat recovery and minimum external heat supply, point e is supposed to come closer to the minimum target curve to make use of the potential ΔA and ΔE savings. However it has to be considered that reaching point c, although it might be the optimal point in a grassroots design, means wasting of existing area in a retrofit project. Point e will rather be moved to the left, to keep the installed equipment and gain energy savings through changes in the existing network instead of creating a completely new network. As certain installation of new technical equipment cannot be avoided the slope of the retrofit curve will be flat in the beginning (starting with moderate investment like re-piping) and gradually become steeper with the implementation of new heat exchangers.

For retrofit projects the concept of area efficiency (figure 17) gives a guideline for the slope of the curve. It is assumed that the new installed heat exchanger area will at least have the same efficiency as the existing one. Although this seems to be a very conservative assumption, the necessary data can easily be obtained from the existing design to target the area for the new heat exchanger network.

$$\alpha = \frac{A_r}{A_{ex}} = \frac{A_1}{A_2} \quad (14)$$



- α Area efficiency
 A_t Minimum area for the current level of heat recovery (target surface area for the new design at the existing energy consumption (E_{ex}))
 A_{ex} Total heat transfer area in existing network
 A_1, A_2 equivalent retrofit areas

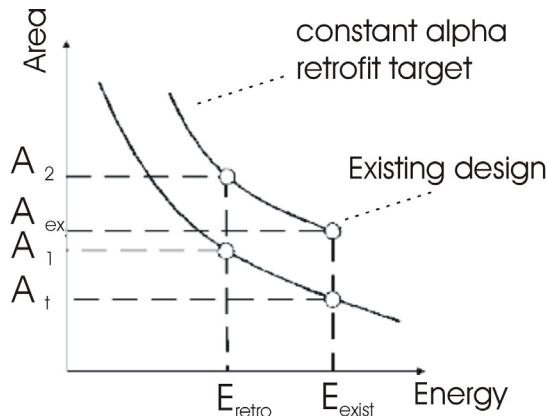


figure 17: energy-area retrofit approach

Another alternative to target the energy savings and corresponding area demand can - independently of area efficiency assumptions - simply be chosen on given payback values, as the payback time is defined as follows:

$$PB = \frac{\text{Investment Cost}}{\text{Annual Savings}} \quad (15)$$

5.3.8 Further considerations

Finding the optimal solution for retrofit projects must not only include analysis of energy saving potential and the corresponding investment costs but also the impacts on the existing process, as every modification of the process might change the operating conditions for existing heat exchangers and only additional evaluation will show how flexible exchangers can be adapted.

Examples for modifications are [4]:

- Addition of a new heat exchanger
- Additional area to an existing unit
- Change of internals in heat exchangers
- Modification of piping on one side of the exchanger



- Modification of piping on both sides of the exchanger
- Moving a heat exchanger to a new location

5.3.8.1 Retrofit targeting based on ΔT_{\min}

The proper choice of ΔT_{\min} is crucial for the development of the whole network. It has to balance driving forces in the heat exchanger, level of heat recovery and for retrofit projects it also has to ensure that the minimum heat exchanger approach temperature is not exceeded for existing equipment.

5.3.8.2 Experience ΔT_{\min} values

For grassroots projects the scenarios are usually different compared to the design opportunities for retrofit tasks, as more influencing parameters deriving from the existing network have to be considered and assessed. Thus practical experience data can serve as guideline for a ΔT_{\min} starting value in a first instance. Linnhoff March [13] is providing typical ΔT_{\min} values for various types or processes and recommends:

Steam against process stream	10 - 20 °C
Cooling water against process stream	15 – 20 °C



5.4 Design

After targeting values for ΔT_{min} , numbers of utilities and heat exchanger area the design phase of the heat exchanger network can be started. As the pinch point is dividing the problem in two different areas, also the design needs to be done for both regions separately. A clear representation and good overview of the existing streams to create an optimal heat exchanger network is provided by the stream grid (figure 18).

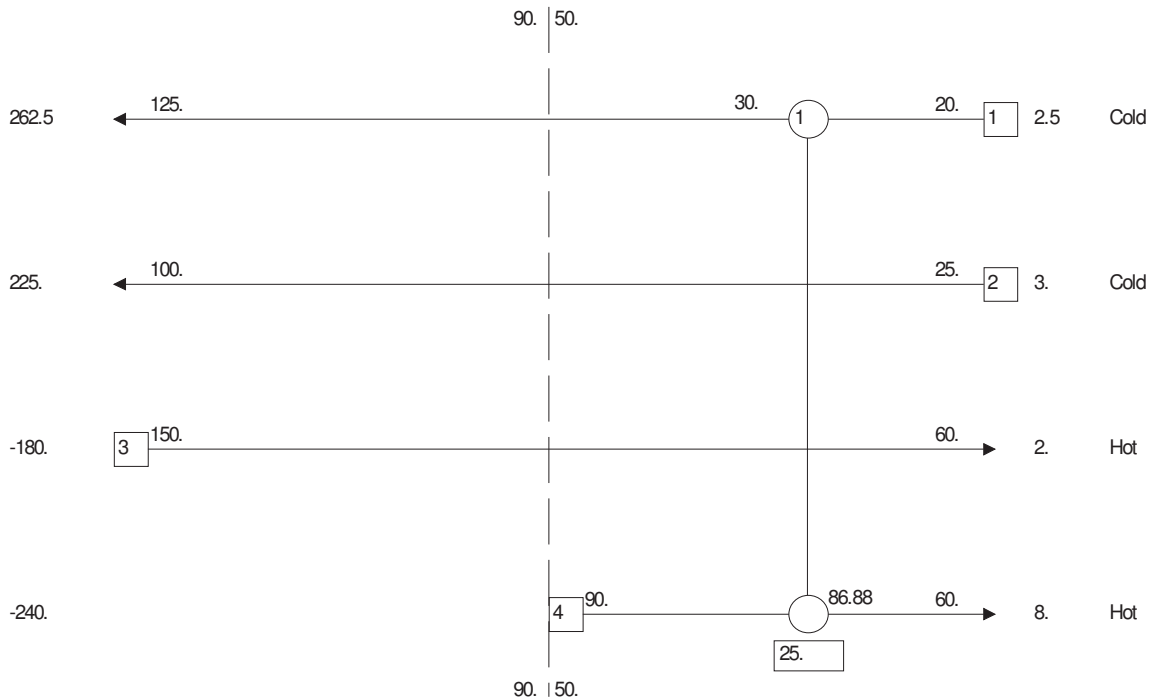


figure 18: stream grid

Hot streams are represented through arrows running from the left to the right, while cold streams start at the right side of the drawing board. The respective temperatures are written above the streams, while the heat load H [kW] can be read in the left column. The numbers in the squares at the beginning of the streams identify the single streams. The Pinch location is positioned as a vertical line dividing the problem into the region above (left side) and below (right side) the pinch. The column right to the stream grid shows the affiliated mass specific heat of the streams, CP [kW/K].

On this sketch heat exchangers and utilities can be placed and will be indicated with single circles for heat or cooling supply, or two connected circles, each on the corresponding streams that will be matched in the heat exchanger. The heat exchanger duty is displayed in a box below the heat exchanger.



The task is now to create a heat exchanger network by placing heat exchangers and utilities in a manner that the targeted values will be achieved.

5.4.1 Pinch Design Method

At the Pinch, which is specified as the point with ΔT_{\min} , the driving forces are limited. To create the maximum heat recovery, the design has to start at this point and develop away from there, to ensure sufficient driving forces for stream matching. The challenge in matching the streams is to find a sequence of placing the exchangers to create a network with maximum heat recovery and minimum units, while obeying the design rules.

5.4.1.1 Pinch Design above and below the Pinch

The region above the Pinch is characterized as heat sink, where hot streams should be utilized as far as possible to keep the demand for external heat supply low. To figure out which streams can be matched right at the Pinch, feasibility criteria developed by Linnhoff [14] have to be applied:

Limiting temperature difference

ΔT_{\min} is the limiting driving force and has to be maintained for every stream match

Number of hot and cold streams

The Pinch rules forbid to implement cooling utilities above the Pinch, therefore the number of cold streams has to be equal or exceed the amount of hot streams. A minimum utility design can only be created if every hot stream can be matched with a cold stream.

Mass specific heat of the streams (CP)

Above the Pinch the hot streams are supposed to be cooled with cold streams which requires a CP relation of $CP_{\text{Hot}} < CP_{\text{Cold}}$. This rule has to be applied for the overall sum of streams as well as every single stream matching.

$$\sum CP_{\text{cold}} \geq \sum CP_{\text{hot}} \quad (16)$$

$$CP_{i,\text{cold}} \geq CP_{j,\text{hot}} \quad (17)$$

The following CP flowcharts (figure 19 and figure 20) give an overview about the feasibility criteria for the design above and below the pinch [15]:



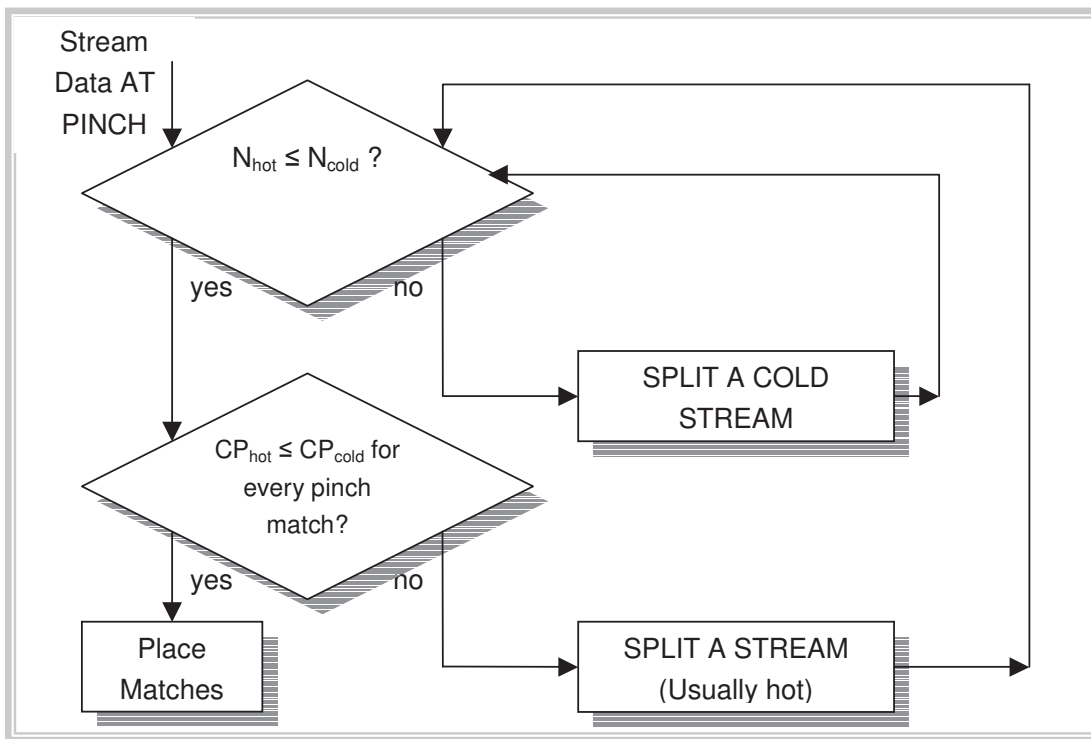


figure 19: rules above the pinch

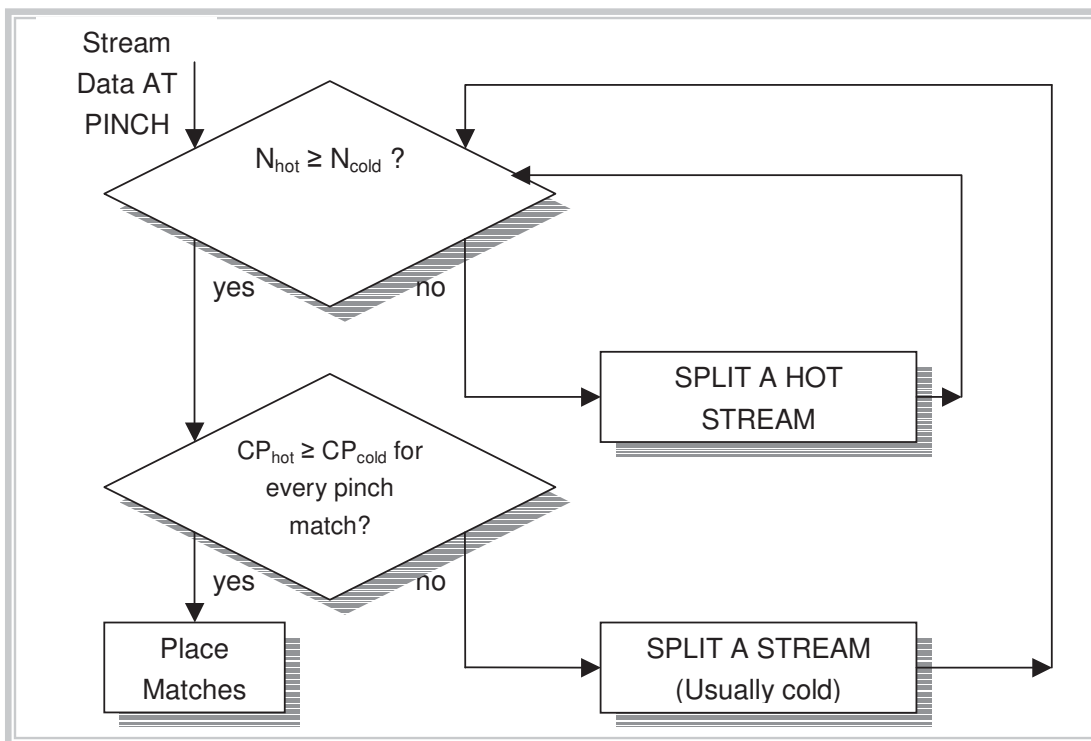


figure 20: rules below the pinch

5.4.1.2 Stream splitting

If the amount of hot and cold streams does not fulfil the feasibility criteria, either a cold (above the pinch) or a hot stream (below the pinch) has to be divided. The number of hot or cold streams will increase with the additional created branches and the requirements are met. Similar rule has to be applied if the comparison between the mass specific heat of the streams (CP) shows a violation of these rules. To facilitate stream matching, the change of CPs by stream splitting might be necessary.

5.4.1.3 Tick off heuristic

An achievement of minimal units to be implemented is only feasible if the match load at the Pinch is maximized, which leads to the application of the tick-off-heuristic. This means that either complete streams or residuals are matched with other streams in a way that no further heat exchanger has to be placed on this specific stream.

5.4.1.4 General procedure to complete the Heat Exchanger Network

Linnhoff March [13] summarizes the general further procedure as follows:

“Network design for a grassroots project:

a) Develop a Minimum Energy Requirement (MER) network

- Divide problem at the Pinch
- Start at the Pinch and move away
- Start with biggest stream "in"
- Observe CP_{OUT} , CP_{IN} , splitting streams where necessary
- Place all pinch matches first
- Maximize loads on all pinch matches to minimize number of units (the "tick-off" rule)
- Fill in the rest
- Merge above and below the pinch designs

b) Evolve the MER network for network simplicity and capital-energy trade-off

- Exploit heat load loops and heat load paths”

What heat loops are, has already been explained. To avoid this unnecessary heat load shifting within the network, those loops are supposed to be broken in a way that the loads are relaxed along a heat load path, without violating temperature driving forces or exchanger capacities. For a detailed description of this procedure it shall be referred to literature [4] or



[8] at this point, instead a simple case study diagram [4] shall illustrate the problem of a heat loop (figure 21) and its solution through heat loop relaxation (figure 22):

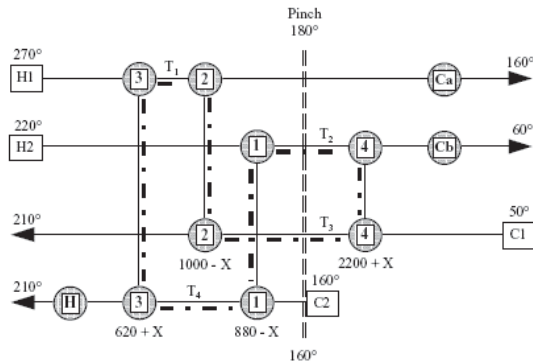


figure 21: heat loop

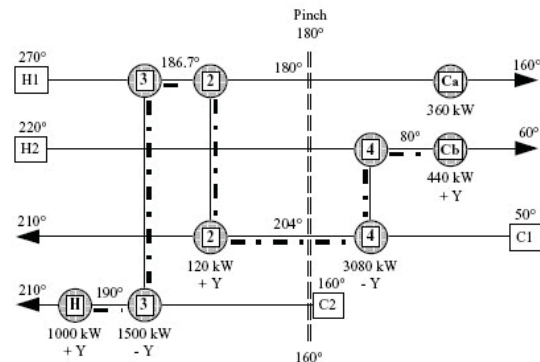


figure 22: loop relaxation

For revamping projects Linnhoff March [13] proposes one of the following three methods:

- “Pinch Design Method with maximum re-use of existing exchangers
- Correction of Cross-Pinch exchangers (starting from the existing network, to “shift” heat exchangers away from a cross Pinch situation)
- Analysis of exchanger paths”

5.4.1.5 Selection of alternatives

Usually several alternatives can be presented and a decision for one has to be made by balancing energy recovery, number of units and area needed.

5.4.1.6 Heat transfer coefficient – Area (UA) Analysis

A very simple method to compare the different options without detailed cost calculation is the UA analysis. For pure counter-current heat exchangers, the product of heat transfer coefficient and heat exchanger area can easily be calculated for each exchanger before and after any retrofit modification: $UA = Q / \Delta T_{LM}$. Following table gives an illustrative UA-value (kW/°C) overview for two alternative Retrofit Designs [4].



table 2: UA analysis

Heat Exchanger	Existing Design	Retrofit A	Retrofit B
1	17.49	44.12	17.49
2	33.91	89.20	89.20
3	-	9.07	27.99
4	-	29.52	35.68
total	51.40	171.91	170.36
Recovered energy	1500 kW	2720 kW	2800 kW

This form of table contains all the relevant information to evaluate the different alternatives and avoids rigorous mathematical cost function programming.

5.4.2 Note: Pinch Analysis focused on Distillation Columns

The energy consumers in the investigated project are the distillation columns. Pinch Analysis offers also methods and principles for appropriate modifications of the distillation columns itself and the proper integration of standalone columns into the overall process.

The standalone column modifications include options like feed stage location, reflux improvement, feed preheating or cooling and intermediate condensing or reboiling. The integration of the column into the total network links the column heating and cooling duties with the process heating and cooling requirements.

This extended energy optimization technique has not been applied when tackling the problems of MAF, as the proposed solution should have been appropriate for the resources of the company. This means economical and controllability limitations would have restricted the application of such highly interlinked networks anyway and the task of optimization within the given limits would not have been fulfilled.



5.4.3 Summary of the key steps in Pinch Technology:

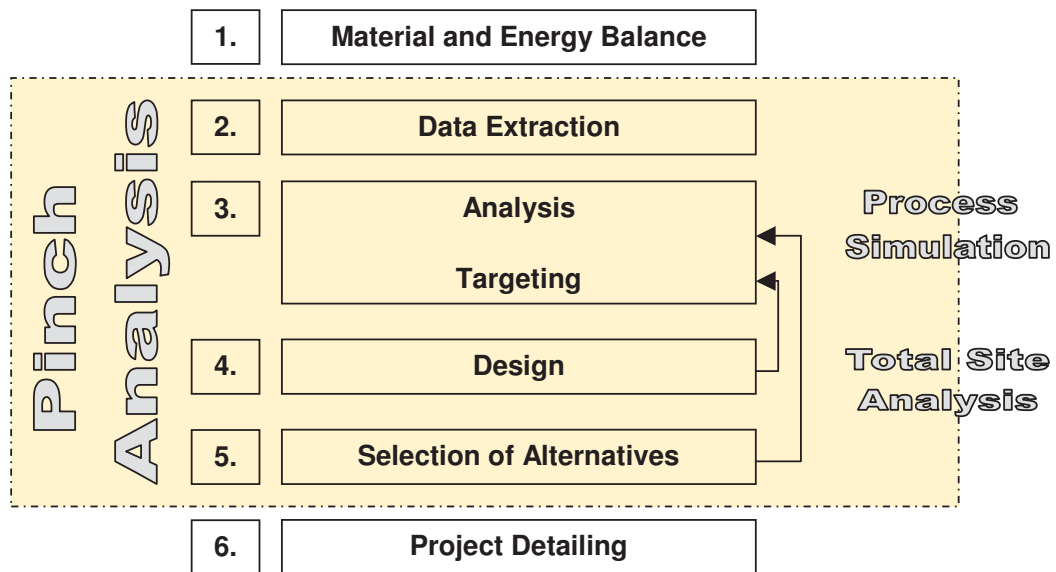


figure 23: Pinch Technology – key steps

6 Initial situation at Mekanissa Alcohol Factory

For the energy recovery task the main focus was laid on the high energy consuming areas: distillation and steam generation. Totally excluded from the investigation were electrical consumers, such as pumps, stirring and aeration equipment for the fermentation sector. The reason for these constraints is basically the lack of availability of specifications for the particular units. Furthermore frequent power breaks from the electricity supplier side lead to diesel-generator activity and an investigation and break down of electricity consumption for the single appliances would have been a time-consuming and irrelevant effort for the company as the clearly defined aim was to reduce the fuel oil consumption per litre ethanol.

Mekanissa Alcohol Factory faces monthly average values of 0.9 – 1.1 l fuel oil per litre produced ethanol (96,6 %vol), while benchmarking with other private Ethiopian distillation plants showed performances up to 0,33 l oil / l ethanol. This big difference was mainly explained with outdated technology usage and insufficient knowledge about heat recovery facilities. Additionally it was admitted that the original existing energy recovery facilities were not maintained and replaced after breakdowns or malfunctions of different equipment.

For decreasing the litre oil per litre ethanol ratio basically three approaches can be considered:

1. To keep the oil consumption constant while increasing the ethanol output
2. To keep the ethanol output constant and search for reduction potential of fuel oil consumption
3. A combination of alternative 1 and 2: to decrease oil consumption and increase ethanol production at the same time.

Discussions with MAF chemists and operators have shown that the rising of ethanol production is limited by the plant capacity itself and is already exhausted. The controllability of the output is also restricted by irregular maintenance work and scale formation. The practiced maintenance procedure is to shut down the plant and remove the scale layers whenever product quality and quantity do not fulfil the standards anymore. Planned, overall maintenance that includes also the cleaning of condensers and reboilers is only scheduled in a 2-years-intervall.

Facing this situation the only option left to reduce the fuel oil consumption per litre ethanol was to search for options where the energy input of fuel oil can be reduced by more efficient equipment (boiler & heat exchanger) or were it can be replaced by other energy means, i.e. unused waste heat that is rejected to the environment. An investigation of the existing heat exchanger network should also reveal potential for optimization in external heating and cooling duty.

Additional improvements, which are not subject of this thesis but tasks of other process optimization projects, can also be expected through an enhanced fermentation process.



Proper handling and pre-treatment of the raw material, which will for instance have impact on the scale formation in the columns, will also affect the heat transfer (energy losses). Modifications of the reflux ratios, column pressure, condensation and reboiler temperatures might also lead to increased product quality and quantity and are influencing the energy consumption at the same time.

Prior to the MAF process description a very short theoretical introduction to basic ethanol production should point out for the management of the company, where MAF does not comply with state of the art systems and conditions. It is basically an opportunity to figure out focus areas for coming process optimization projects.



7 Process description

The first step for analyzing any plant performance is to study the process itself to develop an understanding for the processing sequence, the equipment, the operating conditions (including the control decisions and operator's expertise) and not at last sample taking locations. This familiarization should contain:

- Process flow diagrams (PFD)
- Piping and instrumentation diagrams (PID)
- Operating instructions and time-sequence diagrams
- Unit installation
- Operator perspectives and expertise

The main difficulty faced at MAF derives basically from the almost total lack of manuals or design data. The only flowsheet available was a basic drawing at the (not working) control board, showing the single components of the distillation sector, including: columns, condensers, reboiler and the piping system connecting the single units, as well as the original placed pumps. However the investigation of the process showed that the original design has been modified over the years and parts were dismantled or reconnected in a slightly different way. In order to start with reliable information about "to-be" data, a request was sent to Frilli (Italy), the company that was erecting the plant in 1977 and provided Mekanissa Alcohol Factory with new condensers some years back. Unfortunately the response was rather negative, as absolutely no information according design and specification of the original plant was seemingly available anymore.

Due to this reason the process description is exclusively based on the present state of the system and the data that could be extracted from the existing process.

The fermentation and distillation plant at Mekanissa Alcohol Factory was originally designed to process grain, wine and molasses. However, the only raw material used is blackstrap molasses from local sugar factories. The molasses is brought via truck transport to the factory compound, where it is stored in a housed basin before it is transferred to the actual fermentation sector.

The initial layout of the factory includes clarification equipment for the molasses to separate the solid particles from the rest of the mash in order to avoid unwieldy handling during pumping, aeration, stirring and most important, unwanted scale formation during the distillation process. The clarification of molasses would principally follow a procedure of adding flocculent to help unwanted inorganic impurities to settle as sludge in the clarifier tank and a subsequent pasteurizing in form of heating to destroy unwanted bacteria. Due to lack of spare parts and neglecting to invest in proper maintenance on time this pre-treatment facility is not functioning for several years now and is simply skipped as a result.



7.1 Materials

Input streams

Raw Material:	Molasses
	Water
Additives:	Sulfuric acid (95 %)
	Diammoniumphosphat (DAP)
	Yeast (commercial baker's yeast)
Auxiliaries:	Zeolites (water softener unit)
	Lubricants
Energy:	Electricity
	Fuel oil light

Output streams

Products:

- Ethanol (96,6 %vol) - used for further liquor production in another branch.
- Denatured ethanol (94-96 %vol) ¹

Waste streams

- Drained molasses from the fermentation sector
- Drained molasses from the distillation sector
- Separated water from the distillation sector
- CO₂ release to atmosphere during fermentation process
- Atmospheric release of other volatile substances, such as acids as by-products of bacterial metabolism (Acetic-, Propionic-, Isobutyric-, Butyric-, Isovaleric- and Valeric acid) (about 2300 – 3400 ppm), essential oils, various ethers and also alcohol to some extent.
- Heat release to atmosphere during fermentation process
- Heat release to atmosphere during distillation process
- Flue gas from the boiler section

Other losses

- Spilling of oil during filling process into storage tanks
- Used lubricants (discharged into nearby river)
- Cooling water from air generator activity

¹ The term “denatured alcohol” refers to the indication that the alcohol is not supposed to be used for drinking purpose later on. Denaturants are added to give the ethanol a disagreeable taste, odour or distinctive colour. This measurement derives from the need to distinguish alcohol for beverages and for industrial usage. At Mekanissa Alcohol Factory, the denatured alcohol is coloured and mainly used in hospitals for disinfection purpose and in private households as fuel ethanol.



7.2 Basic ethanol production

7.2.1 Introduction

Ethyl alcohol is a colourless, limpid, volatile liquid, which is flammable and toxic in its pure form. It is one of the most important organic chemicals, soluble in water and most organic liquids. It is used as solvent, extractant, antifreeze, intermediate in the synthesis of numerous organic chemicals and is also an essential ingredient of alcoholic beverages.

7.2.2 Production

Synthetic production as well as industrial fermentation are used to produce ethanol. The group of the synthetic production processes contains following processes [16]:

Catalytic hydration of ethylene

Ethylene (produced by petroleum cracking and from natural gas) is treated with water at high temperatures in the presence of acidic catalysts to produce ethyl alcohol.

Sulfuric acid hydration of ethylene

Ethylene is treated with concentrated sulfuric acid to produce ethyl hydrogen sulfate and diethylsulfate, which are then hydrolyzed to ethyl alcohol and diluted sulfuric acid.

Fischer-Tropsch process

Ethyl alcohol is the major by-product in the synthesis of methanol by the reaction of carbon monoxide and hydrogen over iron catalysts.

In earlier years (until about 1947) industrial alcohol was mainly produced by fermentation, but now the hydration of ethene has become the main source of industrial produced alcohol. Mekanissa Alcohol Factory however is working in the industrial fermentation sector, which is the reason why the focus will be led on this technology.

For industrial fermentation different feedstock can be used. Mainly molasses (blackstrap molasses), grain (barley malt), sulfite waste liquor (from softwoods) or wood waste (the polysaccharides of wood are converted to fermentable sugars by acid hydrolysis) is processed to alcohol.

Industrial ethanol fermentation basically consists of two main processes: fermentation and distillation. Fermentation describes the process, where microorganism (e.g. yeast) under



suitable conditions convert substrate (e.g. sugars) into metabolic products (e.g. ethanol and CO₂). Plants, which do not directly yield simple sugars (e.g. starch or cellulose), must be converted to sugar to allow yeast to ferment it. The resulting product is called “beer” and has to be distilled to obtain ethanol. Distillation is the most commonly used method for the separation of homogenous fluid mixtures. The different boiling points or volatility is exploited to divide the components of the mixture in a column-type process. Vapour and liquid phases flow counter currently within the mass transfer zones in a distillation column, which either contains trays or packings to enable maximum contact between the two phases. To yield a good performance the overhead vapour is (partially) condensed and reenters the column as liquid reflux. At the bottom a vapour stream is provided through a reboiler.

7.2.2.1 Feedstock preparation

Feedstock preparation in general depends on the raw material used and varies between plants that produce simple sugars directly (e.g. sugarcane) or preliminary starch (e.g. corn). Different plants with different sugar content require different treatment to create optimal conditions for enzymes¹ that convert starch to complex sugars (“liquefaction”) and break the complex sugars further down to simple sugars (“saccharification”). The resulting mash finally contains glucose, which is metabolized into ethanol by yeast. The raw material used in Mekanissa Alcohol Factory is molasses and does not require special pretreatment, so this part will not be discussed in detail.

7.2.2.2 Molasses

The cane molasses which is used as feedstock in Mekanissa Alcohol Factory is also referred to as blackstrap molasses and is a thick viscous by-product of sugar production, which still contains about 25 % water, 15 % inorganic salts, 10 % organic matter and 50 % sugar that is not crystallisable anymore but can be used for fermentation. This high sugar content makes molasses very valuable for ethanol production through fermentation and distillation, without pre-treatment in form of cooking and enzymatic activity to hydrolyze starch into fermentable sugars. Nevertheless to increase the efficiency of the fermentation process and reduce the suspended solids that might cause scaling and a blocking of distillation columns, a preliminary treatment of molasses is indispensable.

Proper handling and storage conditions are also important influencing factors to gain the maximum amount of ethanol out of the molasses, as the fermentable sugars are decreasing due to wrong storage conditions. If the molasses is stored between 30-35 °C it will lose 2-3 % per year of its fermentable sugars. A storage temperature increase of 10 °C would lead to

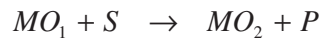
¹ Enzymes are defined as proteins with catalytic properties (specific activation of biochemical processes).



a quadrupled loss [17]. At MAF the molasses storage tank is not cooled, but simply exposed to the environmental temperature, which means daily and seasonal fluctuations ranging from 5 to 30 °C.

7.2.2.3 Fermentation of molasses for ethanol production

The general equation for fermentation is described as:



MO₁ Microorganism

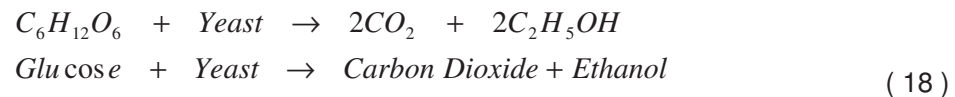
S Substrate

MO₂ increased amount of microorganism

P metabolic products

Yeast growth in particular requires aerobe conditions and glucose as substrate for reproduction.

The consecutive reaction for ethanol production from glucose by fermentation (anaerobe process) is given by the Gay-Lussac (1815) equation:



This equation gives only a summary of various complex and parallel proceedings and it has to be considered that during fermentation not only ethanol and carbon dioxide are produced, but also other substances such as glycerol, succinic acid, acetaldehyde, acetic acid, formic acid, higher alcohol called fusel oil and others, which leads to about 95 % yield out of the maximal theoretical yield under optimal conditions.

To create an ideal environment for yeast and the fermentation process, several influencing factors such as presence of nutrients, moisture content, oxygen relationships, degree of acidity (pH), temperature and presence of other microorganism have to be controlled.

Temperature level and pH

The temperature range of the mash determines to a great extent the growing rate of microorganism. Different species have different requirements regarding T_{max} and T_{min} . The desired temperature level for optimum yeast conditions in the fermentation tank is between 32°C and 33°C. The fermentation process itself is an exothermic reaction (88 kJ/mol). Therefore, depending on the weather conditions in the environment, cooling equipment might be necessary to keep the optimal temperature level in the fermentation



tank. Especially in tropical regions the heat cannot be released to the surrounding and heat exchangers have to be installed, as above 36 °C the vitality of yeast starts to decrease, while the conditions are becoming better for other (undesired) bacteria.

Microorganism are also sensitive to acidity and the populations will grow or shrink depending on the existing pH. For the growth of yeast, the pH should be obtained in the range of 4 – 4.5 which will also restrict the development of other bacterial species and limit bacterial contamination of molasses.

Moisture

Microorganism require sufficient available moisture for their activity, to absorb nutrients in liquid form and discard waste materials through the cell walls. The specific moisture requirements are dependent on several influencing factors (e.g. nutrients, oxygen and temperature) but it is important that in the water hydrophilic colloids, salt, sugar or other solutes are not solved in high amounts. Blackstrap molasses can contain up to 80 °Brix¹ sugar and therefore need to be diluted to below 25 °Brix to avoid cell dehydration caused by a great osmotic pressure through salts and sugars outside the microbial cell.

Nutrient use in molasses fermentation

To create conditions for proper growth and reproduction of yeast, essential elements like carbon, hydrogen, oxygen, nitrogen, phosphorus, sulfur, calcium, iron and magnesium and certain trace elements are necessary.

Particularly nitrogen and phosphorus might have to be added to the diluted molasses to obtain optimum results out of yeast activity. To equalize the deficiencies for these nutrients diammonium phosphate (DAP) is usually added to the molasses solution. Special care has to be taken not to add nitrogen in form of ammonium sulfate as this will cause scaling problems in the further process. Calcium sulfate (gypsum) is the main component that forms scale in the distillation columns. It is formed when sulfuric acid is added to control bacterial contamination or nitrogen is mixed into the mash in form of ammonium sulfate to satisfy the nutrient demand of yeast, through converting of calcium salts to calcium sulfate, which is fairly insoluble especially at higher temperatures. The calcium sulfate therefore precipitates when the molasses beer is fed to the pre-heaters or steam-heated stripping column and causes scaling problems, which reduces the distillation capacity.

¹ °Brix: A measure for the soluble dry substance of a liquid. It is a measure of what the sugar content of a liquid would be if all the dissolved and suspended solids were sugar. The Brix scale is usually used to determine the ripeness of fruits. e.g.: 10 °Brix measures 10 percent of Sucrose in a 100g cane sugar solution.



Impact of ethanol production on yeast

Through activity of the microorganism the mash will be turned into a “beer” of 8-12 % alcohol and then will stop, as ethanol is damaging the cell membrane and so becomes toxic to yeast and inhibits growth of the microorganism. A method to yield the upper ethanol content in the mash is incremental feeding. The sugar concentration in the mash is limited with 25 °Brix, which is only sufficient to yield 7-8 %vol of ethanol in the fermented beer. Lower starting Brix values (e.g. 18 °Brix) allow the yeast population to start growing very rapidly. Through fermentation process the sugar content will gradually decrease. At the level of about 12 °Brix, less diluted molasses (~35 °Brix) is added and beer ethanol levels of up to 12 % can be attained after 60 hours.

Thus a common practice is to set up two different mash dilutions, one for yeast propagation for inoculation and one for the fermentation process itself.

7.2.2.4 Pretreatment for inoculation and yeast propagation

To facilitate fast yeast growth the sugar concentration in the mash has to be kept at a low level. Usually concentrations used for yeast propagation lie between 0.5 and 1.5 %. To inhibit undesired bacteria growth, pH adjustments (pH 4 – 4.5) and sterilization for about 1 hour at a temperature of 120 – 130 °C are widely practiced.

To inoculate or “pitch” the main mash, a “starter” out of a pure yeast-culture has to be prepared. To achieve the increasing of the yeast cells to the utmost extent, aerobic conditions (artificial aeration) are necessary, unlike for the fermentation process itself, which occurs in anaerobic environment.

7.2.2.5 Pretreatment for fermentation and molasses fermentation

Fermentation does not aim to produce as much yeast as possible but to produce ethanol through micro-biological activity. The concentrations are mostly kept between 10 – 18 % sugar content. Molasses with concentrations above this level are diluted with water, before pumping into the fermenter. Higher shares would have the reverse effect and reduce the activity of yeast, which results in longer fermentation times and improper sugar utilization. Lower concentrations on the other hand have the consequence of being uneconomic as fermenting space is not fully used and proportionally higher amounts of fermented mash have to run through the distillation column to achieve a certain quantity of ethanol.

The pretreated mash is finally inoculated by the starter and the anaerobic fermentation process will start. About one percent of alcohol is usually carried out with the evolved carbon dioxide, but can be recovered by scrubbing it with water. In addition, trace quantities of ethyl acetate, isobutyl alcohol, and isoamyl alcohol are also produced.



7.3 Fermentation sector at Mekanissa Alcohol Factory

The molasses specification itself is not constant throughout the year. Variety of cane, composition of soil, climatic conditions, harvesting practices, the sugar manufacturing process itself as well as the handling and storage are factors that are affecting the composition of the molasses.

According to literature typical composition of molasses from sugar cane grown on the African continent is [18]:

Total solids:	75 to 88 %wt
Total reducing sugars:	44 to 60 %wt
Unfermentable sugars:	4 to 5 %wt
Fermentable sugars:	40 to 55 %wt
Total inorganics:	8 to 12 %wt
Settable dry sludge:	< 3.5 %wt
Specific gravity:	1.38 to 1.52 kg/l
Titrate volatile acidity:	3,000- 20,000 ppm
pH at 40 deg. dilution:	4.5 to 5.6
Caramel (OD):	0.2 to 0.6

The average sugar content of the raw material used at Mekanissa Alcohol Factory is about 45-50 %wt and shows °Brix between 75 and 80.

For the processing description see parallel the flowsheet “fermentation sector” in the appendix.

The first step to start the fermentation activity of microorganism is that one part of the untreated molasses will be diluted with water to 30 °Brix and another part to 15 °Brix for yeast propagation. To adjust the pH to the required level of 4 – 4.5 approximately 0.015 kg sulfuric acid per litre produced ethanol (96.6 %vol) is added. Parallel to this dilution step, the yeast initiation is prepared in the laboratory. About 1 kg of dry yeast is dissolved in about 2 l warm water ($T_{H_2O} = 28 - 30 \text{ }^\circ\text{C}$).

About 0.5 kg DAP per 3000 l yeast solution is added as fertilizer. This yeast solution will be transferred into the propagation tank and 10 °Brix molasses (pH 4 – 4.5) is incrementally fed into the tank. Continuous aeration and stirring of the mash creates a favourable environment for yeast growing. For optimal conditions the incremental feeding is controlled in a way that the °Brix of the mixed solution will be kept at a level of 10 °Brix. This is achieved by constant °Brix measurement. As sugar is the main nutrient for yeast a degradation of about 0.5 °Brix per hour can be observed. Once the batch is filled up to the top it therefore takes about 4 hours until the molasses reaches about 8 °Brix. This degraded molasses will be transferred into another tank where it is mixed with the prepared 15 °Brix molasses in a ration 1:1.



During this initiation process the mixture will again be degraded to 10-11 °Brix before it is transferred into the final fermentation tank, where the 30 °Brix molasses will be continuously added by keeping a constant °Brix level of 17 - 18. The fermentation process itself requires anaerobe conditions and generates heat, which should not lead to temperatures above 35 °C. The practical fulfilment of these prerequisite is achieved by constant measurement of the temperature inside the fermentation tank. When the temperature level will exceed 35 °C, the tank will be cooled from outside by manual activated direct water sprinkling.

Although the fermentation process is supposed to take place under anaerobic conditions, the tanks at Mekanissa Alcohol Factory are not covered due to uncontrollable foam generation, as there is no antifoam supply. Hence the generated CO₂ is not captured and converted into marketable products, such as dry ice, liquid CO₂ for soft drinks, fire-fighting foams, filtration products or other various industrial uses.

Once the fermentation tank is filled to the top, the mash will have about 15-16 °Brix and 35-37 °C. The final fermentation tank does not aim to increase the amount of yeast anymore, but to produce ethanol through biological activity of yeast. After 36-48 hours the °Brix level will be decreased to 10-11 °Brix while the ethanol content will rise up to 7-8.6 %vol.

The fermented mash will now be pumped into a decantation tank where the remaining sludge is supposed to settle within 60 hours. There is no control mechanism to assess the amount of sludge settled, a drainage valve will simply be opened manually and the consistence and colour of the drainage will be monitored. If the sludge part of the fermented mash is separated, the valve will be closed again and the remaining liquid will be stored in an intermediate tank before it will be further distilled. The decanted part is discharged into the nearby river without any further treatment.

7.4 Distillation sector

See appendix for Process Flow Diagram (PFD) and Piping and Instrumentation Diagram (PID)

The flow diagram includes basically three interlinked networks and represents the actual process, as compared to the original PID some reboilers have been dismantled and pipes blinded:

- Steam production and steam flow
- Ethanol distillation (raw material input and product, byproduct and waste output)
- Cooling water cycle

7.4.1 Steam production and steam flow

The steam production starts with the water softener unit in the boiler section to pretreat the feed water in order to minimize scale formation in the boiler. Three softener units with each



0,6 m³ water content are manually charged with zeolites, removing the positively charged ions such as calcium (Ca²⁺) and magnesium (Mg²⁺) ions by exchanging sodium ions for calcium and magnesium ions¹).

A formed scale layer causes overheating and failure of boiler tubes, as the thermal conductivity of porous boiler scale is similar to insulating brick and therefore acts in a counterproductive way for heat exchange. This phenomenon prevents an efficient transfer of heat through the tubes to the circulating water and lower boiler efficiency will be the result. Another important aspect is that corrosion may occur under the boiler scale.

In general boiler scale causes:

- Increased fuel consumption by decreasing the operating efficiency
- Thermal damage
- Unscheduled down-time
- Increased cleaning time and cleaning costs
- Reduced working life of a boiler

At Mekanissa Alcohol Factory insufficient attention has been paid to these impacts of untreated water in the past and forced the boiler operators to dismantle some of the heat exchanger tubes inside the boiler. As a consequence the nominal steam production and fuel oil consumption (see table 3) were not representative for the actual process anymore.

The technicians and chemists of the company have already identified this boiler inefficiency, and a new boiler was purchased to replace the two old ones. For the time being however, the new boiler is placed in a casing on the compound and has not yet been installed. A decision about the installation time was not yet made and the investigation of the boiler section as part of the overall energy optimization concept was kept briefly, aiming to get at least comparative figures between the old system and the new one, once it will be replaced.

¹ Ion exchange can be written as reversible reaction involving chemically equivalent quantities. e.g.:
$$Ca^{++}(aq) + 2NaR \Leftrightarrow CaR_2 + 2Na^+$$

R represents a stationary univalent anionic site in the polyelectrolyte network of the exchanger phase.

General Hardness (GH) measures the cations (+ charge) for calcium and magnesium. Carbonate Hardness (CH) refers to only the bicarbonate, and carbonate anions (- charge); it does not measure the sulfates and other anions. The common units found in the literature are degrees of general hardness dGH (GH) from the German system (1dGH = 10 mg/liter CaO) or ppm Ca from CaCO₃.



To preheat the boiler feed water, some of the generated steam is directly led into the preheater tank and raises the steam temperature up to about 70 °C.

The two boilers are producing saturated steam, which is led into the distillation sector through one joint pipe.

table 3: Design specification VAPOMAT – steam boiler

Model	Year	Steamproduction	Nominal Output	Fuel Oil Consumption	Efficiency
		[kg /hr]	[kcal /hr]	[kg /hr]	[%]
3.20	1977	2.000	1,170.000	138	87
3.25	1979	2.500	1,146.000	172	87

The steam production is referred to a pressure of 12 bar and a feeding water temperature of 80 °C. The fuel consumption is referred to fuel oils having a lower calorific value of 9800 kcal / kg. It is demanded that the fuel oil should have:

Lower calorific value:	9.800 kcal/kg
Max density:	0.94
Max viscosity:	20 °Engler at 50 °C
Pour point:	0 °C

The specification of the actual used Fuel Oil Light (inland code: 98 181 FC) shows that the used oil fulfils the requirements. The details can be found in the appendix.

Table 4 shows the real boiler capacities and clearly points out, that the previously listed specification data are far away from what is achieved now. A detailed efficiency calculation will be presented later on.

table 4: actual boiler capacities

	fresh water temperature	feed water temperature	average steam flowrate boiler 3.20	average steam flowrate boiler 3.25	total average steam flowrate
	[°C]	[°C]	[l/hour]	[l/hour]	[l/hour]
average	24.90	69.17	1080.96	1198.13	2279.09

The average oil consumption for both boilers is 3,979.32 l/day.



The total steam is provided at 9 bar. The pressure will be automatically reduced to 2 – 2.5 bar as soon as the main steam pipe enters the distillation area and will be further reduced by manual valve adjustment. The steam inlet pressure varies for the different columns:

Distillation column	170 mbar
Rectification column	125 mbar
Oil column	88 mbar

All working columns are operated with direct steam supply; additionally oil and rectification column have also reboilers for vapour generation.

The method of steam distillation and combined reboiler usage might seem to be somewhat unreasonable at first sight, but has energetic and qualitative advantages.

Steam distillation is commonly used for the extraction of essential oils and essences. The objective of the steam distillation at MAF is to remove relative small quantities of high boiling substances (e.g. fusel oils) from a large quantity of product. Using general higher distillation temperatures would call for more energy input to heat the large amount of product up to the higher temperature. For small amount of volatile substance to be distilled it therefore might pay off to use large amounts of steam and still save steam, which might have been necessary to raise the temperature level of the total liquid stream. This can be explained with the large partial pressure that is introduced into the system by blowing steam through. A liquid boils when the total vapour pressure of the liquid equals the external pressure of the system, so the original high vapour pressure of the liquid will be reduced to low partial pressure and the energy needed to evaporate the high boiling point components will be reduced significantly.

The mixture will boil when the vapour pressure of the volatile component reaches $P-p_s$ (where p_s is the introduced steam pressure and P is the original total system pressure), which is lower compared to the original system where no steam is present: $(P-p_s) < P$.

However the steam should not be introduced to evaporate the other part of the liquid, as this would introduce water into the liquid phase and affect the distillation adversely. Hence the heat in the distilled steam will be wasted in the condenser, if no heat recovery mechanism is present. This additional water input explains the complementing use of reboilers, as they are producing the upflowing vapour stream without introducing additional water that would have to be condensed and therefore would cause increased utility demand in terms of cooling water as well as steam.

7.4.2 Ethanol distillation

Column and heat exchanger description as well as the process flowsheet, can be found in the appendix.



The demethylizing column is not operating at the moment. The reason for the decoupling of this column lies in the raw material used. Fermented sugar cane molasses does not contain methanol. In case of corn or wheat base, the demethylizing column would be essential to fulfil the quality requirements of the final product. For molasses usage however, there is no need to reactivate the final column at the moment.

All columns except the demethylizing column are supplied with direct steam. Additional reboilers are installed at the rectification and fusel oil column, which aim to avoid further dilution of the ethanol streams. At the distillation column two more reboilers were placed, which are disconnected due to leakage. The reboilers are also operated with steam and the condensed water of the reboilers was originally meant to preheat the boiler feed water. Presently only a small water stream from the oil column reboiler actually flows to the boiler section. This single stream has insignificant influence on the temperature of the boiler feed water. The condensed steam from the rectification column has to be drained manually every 24 – 36 hours as the pump for pumping the condensed water to the boiler feed water preheater is not working and was not replaced.

On top of filter, rectification and fusel oil column are each two vertical condensers placed to condense the overhead vapours and create reflux. In the first condenser usually only partial condensation occurs, while the second one condenses the remaining vapour.

The distillation column has an acidity condenser for the top vapour, one feed preheater and two more condensers, whereas the feed preheater and first condenser are already condensing the whole vapour stream withdrawn from the last steel section of the distillation column, hence the third condenser does not receive any vapour anymore.

The upper four sections of the distillation column and the acidity condenser are made of copper, aiming to remove the sulfuric acid from the vapour.

Sulfuric acid + Copper → copper sulphate + Hydrogen

The copper sulphate precipitates at the copper and will be removed manually (without chemicals) during maintenance work together with the scale layer that is formed due to other organic and inorganic matter in the feed molasses. Within the process investigation period however the company faced some problems and irregularities with this section. Especially the copper-condenser was malfunctioning, so that this part had to be excluded in further analysis.

7.4.2.1 Mode of operation

Usually the fermented molasses is degassed before it is pumped into the distillation column, at MAF however the molasses is not passing a degasser before it comes to the distillation sector. The feed molasses is preheated from 25 °C up to about 60 – 70 °C (desired level is 70 °C) by the overhead vapours of the distillation column and then fed into the distillation



column. Direct steam is used to heat the solution and to drive off the alcohol vapour. An ethanol-water mixture containing about 56 %vol ethanol is extracted from the molasses. This diluted ethanol vapour will be refluxed to the filter column, while at the bottom of the distillation column, stillage is removed, which contains about 0.32 %wt of ethanol. This stillage could be utilized for animal feed or has to be disposed, as it is the case for MAF, where it is simply discharged into the nearby river.

In the filter column the ethanol is separated from the fermentation byproducts, such as aldehydes, ethers, methanol and higher alcohols. To separate the ethanol from the rest, the different boiling points of the components are exploited. At the top of the column the more volatile elements (compared to ethanol) are removed (aldehydes, ethers and methanol). After the second filter condenser denatured alcohol is withdrawn and directly send to the product cooler before it will be stored.

Fusel oils (higher alcohols) however are less volatile than pure ethanol and remain in the withdrawn ethanol stream. The diluted ethanol stream, containing 41,3 %Vol ethanol from the filter column will be fed into the distillation and rectification column.

In the rectification column, the ethanol stream will be concentrated to its final specification of 96.6 %vol. Higher concentrations of ethanol can only be achieved through azeotropic distillation methods. The product stream is withdrawn from the top section of the column and goes through the product cooler before it will reach the storage tank.

The discharged bottom liquid phase from the rectification column was tested on alcohol content but showed 0 % Ethanol. The vapour phase is used to preheat the feed stream to the oil column and will be fed to the oil column as well.

Although the fusel oils are less volatile than pure ethanol, they are still more volatile compared to the ethanol in solutions with high concentrations of water, so that they can be steam-distilled, too. At the lower trays of the rectifying column two more side streams in form of high and low tail fraction (commonly described as fusel oil carrying streams) are withdrawn, whereas the low tail pipe is closed at the moment due to pump breakdown.

Fusel oil is an oily, colourless liquid with a disagreeable odour and taste. It is a mixture of alcohols (largely amyl alcohols) and fatty acids, formed during the alcoholic fermentation of organic materials. After imperfect distillation of these fermentation products it becomes an impurity in the distilled liquor. Fusel oil can be used as a solvent in the manufacture of certain lacquers and enamels (it dissolves nitrocellulose). It has a detrimental effect on the human system [19] and their presence in alcoholic beverages is known to be a cause of headaches and hangovers.



The fusel oil composition for molasses feed is usually approximately in following range [20]:

1-Propanol	13.2	wt%
1-Butanol (n-butylalcohol)	0.2-0.7	wt%
2-Methyl-1-propanol (isobutyl alcohol)	15.8	wt%
2-Methyl-1-butanol (active amyl alcohol)	28.4	wt%
3-Methyl-1-butanol (isoamyl alcohol)	37.4	wt%

The fusel oil column aims to separate these low and high boiling fractions in order to recover ethanol that is recycled to the filter column. The oil fraction that is collected in the decanter will be discharged about every second month.

Side streams in form of denatured alcohol are only withdrawn from the second condensers of the filter and fusel oil column. A third source for denatured ethanol would be the acidity condenser, which is, as already mentioned, not working for the time being. Before the products (ethanol as well as denatured alcohol) are finally reaching the storage tank, the pipes are going through the product cooler, which is using water as a cooling medium.

7.4.2.2 Operational problems

Manual control of steam and cooling water supply are causing irregular vapour-release to the atmosphere. As a result of these inconstant process conditions, temperatures, pressures and flowrates are also underlying certain fluctuations, which will be discussed in detail in the mass and energy balance chapter.

7.4.3 Cooling water cycle

The cooling water pipes are passing the oil-water-separation equipment before they are merging again on the third level, where the condensers are placed. From this joint cooling water pipe the single cooling water streams for each condenser are withdrawn. Water regulators, which should control the amount of water flowing into the condensers, are not working any more. Thus the distillation operators are adjusting the valves manually by observing the vapour release to the atmosphere as well as the temperatures in the second condensers of each column. The cooling water is used to condense all overhead vapours, except in the molasses feed-preheater condenser, above the distillation column and to cool all the product streams (96,6 %vol ethanol and denatured ethanol) in one final heat exchanger. The cooling water itself is cooled by an air-cooling tower (see appendix for nominal specifications).



8 Data collection

8.1 Relevant data

The relevant data that are required to compute complete mass- and energy balances include following areas:

- Stream identification
- Temperatures
- Pressures
- Mass-flowrates (measured and calculated)
- Energy-flowrates (calculated based on \dot{m} , T, p and thermodynamic properties)

8.1.1 Stream identification

Compare with generated flowsheet

8.1.2 Flowrates

For the essential flow-rates, functioning flowmeters were mostly available, except for the cooling water cycle and the boiler-water consumption.

The actual amount of water circling in the cooling water grid could not be determined directly, as the added amount of water depends on the inlet and outlet water temperatures at the cooling tower itself. These temperatures are checked in a 2-hours-interval. If the cooling tower outlet temperature of the water exceeds 30-35 °C, a valve to add fresh water is opened manually and closed again, after the required temperature range is reached. The cooling water mass-flowrate was therefore simply calculated by the required cooling loads in correlation with the actual cooling water temperatures.

$$\sum m_i c_{p,i} \Delta T_i + \sum m_i \Delta H_{v,i} = m_{H_2O} c_{p,H_2O} \Delta T_{H_2O} \quad (19)$$

m_i	massflowrate of stream i to be cooled
$c_{p,i}$	specific heat capacity of stream i
ΔT_i	temperature interval of stream i (before and after cooling)
$\Delta H_{v,i}$	latent heat of stream i
m_{H_2O}	total amount of water in the cooling water cycle
c_{p,H_2O}	specific heat capacity of water
ΔT_{H_2O}	temperature difference of cooling water before and after the cooling tower.



For the steam production the amount of water pumped into the two boilers could be determined by using an installed flow-meter. Having this data, a difficulty occurred, as the distribution of the steam into the single columns and respective reboilers is only controlled manually and a part of the steam is also used to preheat the boiler water, which is again adjusted by the boiler operators themselves according to the fresh water temperature.

8.1.3 Alcohol content

The alcohol content of the fermented molasses is frequently measured by using a Maligand Ebulliometer¹. For pure ethanol-water mixtures an alcohol hydrometer was used. The alcohol content of product streams and feed molasses are permanently checked by the chemists, whereas for the intermediate streams (between the columns and condensers) samples had to be taken and analyzed separately.

8.1.4 Temperatures

For measuring the temperatures, thermometers were installed only inside the columns, inside the second condenser of each column and right after / before the cooling tower.

The stream temperatures itself mostly could not be determined directly, except a leakage at the flange or valve of a pipe allowed to get direct access to the fluid flowing inside the pipe. For the remaining streams a digital thermometer was purchased by the company, to measure at least the surface temperature of the stream carrying pipe. As the operating conditions could not be kept totally stable throughout the data collection, the gained data were somehow varying and the average values show sometimes high deviation rates. On the other hand however this displays the actual working procedures at the company and deliberately modified process conditions would have lead to a distorted picture of the process.

Where possible, the temperatures were taken at leakage spots to get the actual temperature right away. For the other pipes the surface contact area was cleaned and properly brushed with a wire brush.

The temperatures were taken several times on different days and different times and the energy balance is finally based on the average temperatures.

¹ Ebullioscopy: a technique for finding molecular weights of substances by measuring the extent to which they change the boiling point of a solvent; Ebulliometer is a common tool used to determine alcohol content by volume by measuring differences in boiling points between water and alcohol.



For the condenser area, the column overhead vapour, the reflux stream out of the first partial condenser and the remaining vapour going to the second total condenser are approximated to be at the same temperature level.

8.1.5 Pressures

Within the steam grid are five manometers installed: one in the main steam line before pressure reduction, one right after reducing the pressure and one each at the steam inlet of distillation, rectification and fusel oil column. For all the column-overhead streams atmospheric pressure was assumed.

The result of the data collection is summarized in the appendix.

8.2 Additional required calculations

8.2.1 Calculation of density of fluid mixtures

As the flow rates for the single streams in the distillation process at MAF could only be measured on a volume-basis, the densities of the streams at the respective temperatures and pressures have to be determined in order to solve the energy- and mass balance.

The calculation for the density of liquid mixtures proposed by Amagat [21] is:

$$\rho_m = \sum x_i \rho_i \quad (20)$$

ρ_m Density of mixture at respective temperature

x_i Mole fraction of component i

ρ_i Density of pure component at respective temperature

Regarding the true density of the molasses it generally lies in the range of 1.4 – 1.5 kg/l. Fine air bubbles can be entrained during the filling process, which escape only very slowly. This included air ranges between 5-17 %vol and makes it indispensable to determine the exact density by weighing.



8.2.2 Calculation of thermodynamic stream properties

Specific heat capacity for liquid mixtures [20]

$$c_{p,l,mix} \cong \sum x_i c_{p,l,i} \quad (21)$$

$c_{p,l,mix}$	liquid heat capacity of the mixture
x_i	mole fraction of component i
$c_{p,l,i}$	liquid heat capacity of component i

Specific heat capacity for vapour mixtures

For the calculation of $c_{p,vapour,mixture}$ an ideal gas system is assumed. This implies that the heat capacity is only temperature dependent. Pressure independence is justified, as high-pressure rates do not occur.

To consider the temperature following equation is used:

$$c_{p,v,i} = A + BT + CT^2 + DT^3 \quad (22)$$

$c_{p,v,i}$	vapour heat capacity of component i
T	respective temperature of gas
A, B, C, D	constants that can be found in literature

The specific heat capacity for vapour mixtures is calculated similar to $c_{p,liquid,mix}$

$$c_{p,mix}^{id} = \sum y_i c_{p,v,i}^{id} \quad (23)$$

$c_{p,mix}^{id}$	heat capacity of ideal gas mixture
y_i	mole fraction of component i
$c_{p,v,i}^{id}$	heat capacity of ideal gas component i



Heat of vaporization of mixtures

Method of Pitzer

$$\frac{\Delta H_v}{RT_c} = 7,08(1 - T_r)^{0.354} + 10,95 \cdot \omega \cdot (1 - T_r)^{0.456} \quad (24)$$

which can be applied for

$$0.6 < T_r < 1$$

whereas

$$T_r = \frac{T}{T_c} \quad (25)$$

T_r	reduced temperature
T_c	critical temperature
ω	acentric factor
R	gas constant
ΔH_v	heat of vaporization

If the required reduced temperature range is not fulfilled, an estimation that delivers accurate results is given by [21]:

$$\Delta h_v(T_{r2}) = \Delta h_v(T_{r1}) \left(\frac{1 - T_{r2}}{1 - T_{r1}} \right) \quad (26)$$

$\Delta h_v(T_{r1})$	heat of vaporization at reduced Temperature 1
$\Delta h_v(T_{r2})$	heat of vaporization at reduced Temperature 2

For the mixture the single heats of vaporization of the components at the respective temperature are added:

$$\Delta h_{v,mix} = \sum \Delta h_{v,i} \cdot x_i \quad (27)$$

$\Delta h_{v,mix}$	heat of vaporization of the mixture
$\Delta h_{v,i}$	heat of vaporization of component i
x_i	molar fraction of component i



9 Mass and component balance

9.1 Measured and calculated results

9.1.1 Overall mass and component balance

The overall mass and component balance (table 5) is restricted to the three main components in the system: water, ethanol and solids in the molasses. All the trace elements like fusel oils etc. are not considered, as it was not possible to get detailed analysis. All further calculations and simulations are based on these key components and will not consider any other elements.

table 5: overall mass and component balance

		total	water	ethanol	solid
		[kg/hr]	[kg/hr]	[kg/hr]	[kg/hr]
SYSTEM INPUT	feed (7.2 %vol alc.)	3,093	2,651	179	263
	steam	2,200	2,200	-	-
sum		5,293	4,851	179	263
SYSTEM OUTPUT	drainage (ethanol only from distillation column drainage)	4,451	4,147	12	289
	ethanol losses to atmosphere - estimated (from total ethanol input)			2	
	steam to preheat boiler water		191		
	max. amount of drained H ₂ O at rect. reboiler		21		
	96.3 %vol ethanol	14	1	13	-
	94.3 %vol ethanol	5	0	5	-
	96.6 %vol ethanol	119	6	113	-
sum		4,589	4,366	144	289
difference abs.		704	486	34	- 27
difference [%]		13.30	10.01	19.30	-10.18

Factors that influence the deviation:

- The drainage flowrate could only be determined by applying the bucket method and therefore implies higher uncertainty rates compared to the other flowrates, where a flowmeter could be used.
- The water that has to be drained from the reboiler at the rectification column due to malfunction can only be estimated. Emptying the reboiler about once a day equals max. 500 l condensed steam per day, considering the geometry of the reboiler (a specification of the reboiler was not available anymore).



- The ethanol losses to atmosphere could not be determined. Additionally they are fluctuating, as cooling water and steam input are controlled manually. Hence the assumption of 1 % losses might differ considerably from reality.
- The ethanol content of the feed stream is not determinable with the available means; the measured %vol alc. includes higher alcohols, ketones, aldehydes etc.; Literature [17] states that the content of produced byproducts highly depends on the fermentation process and cannot be predicted without closer investigation of the fermentation process. The fermentation sector at MAF is facing qualitative problems and parallel projects for this part have been started already.

9.1.2 Single column mass and component balance

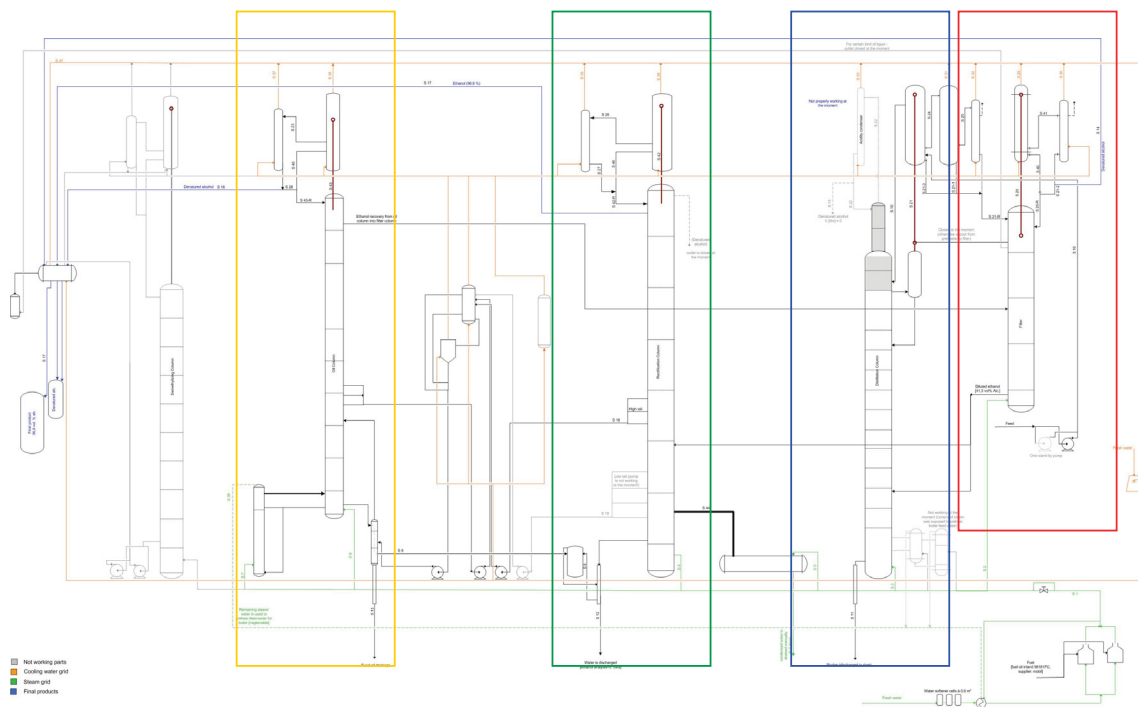


figure 24: overview columns

The following equations describe mass and component balances for each column (see figure 24 for system boundaries; for stream numbers see flowsheet “Distillation Sector” in the appendix), where S stands for the mass flowrate of the respective stream and w denotes the mass fraction of ethanol in the respective stream.



□ Filter Column

1.) Condenser area

$$I. \quad S_{40} + (S_{21_2} - S_{14}) = S_{20_R}$$

$$II. \quad S_{40} * w_{40} + (S_{21_2} * w_{21_2} - S_{14} * w_{14}) = S_{20_R} * w_{20_R}$$

$$III. \quad S_{21_2} = 2 * S_{14} \quad [ASSUMPTION; S_{21_2} \text{ refers to stream after withdrawal of } S_{14}]$$

2.) Total Column

$$I. \quad S_{21_R} + S_{20_R} + REC_{oil} + STEAM_{filter} = S_{20} + BOTTOM_{filter}$$

$$II. \quad S_{21_R} * w_{21_R} + S_{20_R} * w_{20_R} + REC_{oil} * w_{oil} = S_{20} * w_{20} + BOTTOM_{filter} * w_{B,filter}$$

→ Bottom_Filter

→ Steam_Filter

□ Distillation Column

Total Column

$$I. \quad S_{10} - S_{21} + BOTTOM_{filter} * F + STEAM_{dist} = DRAINAGE_{dist}$$

$$II. \quad S_{10} * w_{10} - S_{21} * w_{21} + BOTTOM_{filter} * F * w_{B,filter} = DRAINAGE_{dist} * w_{Dr,dist}$$

Whereas F is the distribution factor for the filter bottom stream which is divided into one stream flowing to the distillation column and one to the rectification column.

It is assumed that 0.4 of the total massflowrate is led into the distillation column and the remaining 60 % will be fed to the rectification column.

→ Drainage_Distillation

→ Steam_Distillation



□ Rectification Column

Total Column

$$I. \quad STEAM_{rect} + BOTTOM_{filter} * (1 - F) = S17 + HighTail + DRAINAGE_{rect}$$

$$II. \quad BOTTOM_{filter} * (1 - F) * w_{B,filter} = S17 * w_{17} + HighTail * w_{HighTail}$$

The drainage part that leaves the system at the rectification column was calculated iteratively over the whole balancing system, after the steam inlet into the oil column was fixed.

→ High Tail

→ Steam_Rectification

□ Oil Column

1. Condenser Area

$$I. \quad (S28 - S16) + S45 = S43_R$$

$$II. \quad (S28 * w_{28} - S16 * w_{16}) + S45 * w_{45} = S43_R * w_{43_R}$$

$$III. \quad S28 = 2 * S16 \quad [ASSUMPTION; S28 \text{ refers to stream after withdrawal of } S16]$$

2. Total Column

$$I. \quad S43_R + OIL2 + STEAM_{oil} = S43 + REC + OIL1 + OIL3$$

$$II. \quad S43_R * w_{43_R} + OIL2 * w_{oil2} = S43 * w_{43} + REC * w_{REC} + OIL1 * w_{oil1}$$

$$III. \quad OIL2 - OIL1 = HighTail \quad [APPROXIMATION]$$

The direct steam input into the oil column was calculated by taking the overall steam that enters the system, minus 10 % losses through leakages and valves, minus the steam injection into the other 3 columns and minus the steam used in the reboilers. The steam usage in the reboilers was again found iteratively by parallel calculation of the molar flow-balance.

→ Oil_3_out (equals drainage part from this column)

→ Oil_1_out



9.1.2.1 Result for the mass and component balance

table 6: results for mass balance

			Mass flowrate	ethanol content
section	calculation sequence	Stream	kg/hr	weight share
filter		S21_R	622	0.48
		S20_R	285	0.92
		REC from oil	34	0.76
	2	STEAM filter	265	0.00
		S14	14	0.94
		S20	298	0.92
	1	BOTTOM filter	907	0.35
distillation		S10 (feed)	3,093	0.06
		S21 (=S21_R)	622	0.48
		BOTTOM filter	907	0.35
	4	STEAM dist	799	0.00
	3	DRAINAGE dist	3,637	0.00
ectification		S42	823	0.95
		S42_R	823	0.95
		S17	119	0.95
		BOTTOM filter	907	0.35
	5	High Tail	89	0.83
	6	STEAM rect	155	0.00
		DRAINAGE rect	487	0.00
oil		S43	347	0.77
		S43_R	342	0.77
		REC from oil	34	0.76
	8	OIL 1 out	52	0.01
		OIL 2 in (high tail + OIL 1 out)	141	0.22
		High Tail	89	0.83
	7	OIL 3 out (to rect. Drainage)	326	0.00
		STEAM oil	276	
		SUM STEAM	1,496	
		SUM DRAINAGE	4,450	

NOTE:

ASSUMPTION: steam losses in the piping system = 10 %; 175 kg steam go back to preheat the boiler feed water. As a result 363 kg / hr steam are utilized in the reboilers. This steam usage in the reboiler however could not be verified, as no specification was available and therefore hides a certain inaccuracy factor.



9.1.2.2 Result verification

The results were crosschecked in two ways: component balance (table 7) and molar flow balance (table 8).

table 7: result – component balance

section	Stream	COMPONENT BALANCE			
		Ethanol content	water content	total	input-output
		[kg]	[kg]	[kg]	[kg]
filter	S21_R	301	321	622	
	S20_R	262	22	285	
	REC from oil	25	8	34	
	STEAM filter	0	265	265	
	S14	13	1	14	
	S20	275	23	298	
	BOTTOM filter	314	594	907	0
distillation	S10 (feed)	186	2,907	3,093	
	S21 (=S21_R)	301	321	622	
	BOTTOM filter	127	240	367	
	STEAM dist	0	799	799	
	DRAINAGE dist	12	3,625	3,637	0
rectification	S42	780	43	823	
	S42_R	780	43	823	
	S17	113	6	119	
	BOTTOM filter	187	353	540	
	High Tail	74	15	89	
	STEAM rect	0	155	155	
	DRAINAGE rect	0	487	487	0
oil	S43	268	79	347	
	S43_R	263	79	342	
	REC from oil	25	8	34	
	OIL 1 out	1	52	52	
	OIL 2 in (high tail + OIL 1 out)	31	110	141	
	High Tail	74	15	89	
	OIL 3 out (to rect. Drainage)	0	326	326	
	STEAM oil	0	276	276	0



For binary water-ethanol systems constant molar flowrates are assumed throughout the columns, which means for every mole steam that condenses, one mole of ethanol is supposed to be evaporated, unless the latent heat of the steam is not an energetic loss.

table 8: result – molar flow balance

section	Stream		molar liquid in	molar liquid out	molar vapour in	molar vapour out
filter	S21_R		24,341			
	S20_R		6,929			
	REC from oil		1,010			
	STEAM filter					
	S14				14,737	
	S20					7,247
	BOTTOM filter				39,768	
	REFLUX RATIO	SUBCOOLED & LOSSES	7,490			7,490
		TOTAL	39,769	39,768	14,737	14,737
		DIFFERENCE		0.004%	-	
distillation	S10 (feed)		151,359			
	S21 (=S21_R)					24,341
	BOTTOM filter		16,106			
	STEAM dist				44,326	
	DRAINAGE dist			186,945		
			SUBCOOLED & LOSSES	19,985		
		TOTAL	187,450	186,945	44,326	44,326
		DIFFERENCE		0.270%	-	
rectification	S42					19,335
	S42_R		19,335			
	S17			2,804		
	BOTTOM filter		23,662			
	High Tail			2,440		
	STEAM rect				8,632	
	DRAINAGE rect			27,049		
			REBOILER DUTY		10,704	10,704
		TOTAL	42,997	42,997	19,335	19,335
		DIFFERENCE	-	0.000%	-	
oil	S43					10,224
	S43_R		10,091			
	REC from oil			1,010		
	OIL 1 out			2,877		
	OIL 2 in (high tail + OIL 1 out)		6,778			
	High Tail					
	OIL 3 out (to rect. Drainage)					18,082
	STEAM oil				15,324	
		REBOILER DUTY		12,982	12,982	
		TOTAL	16,869	16,869	28,306	28,306
		DIFFERENCE	-	0.000%	-	



Comparing the molar flowrates within the vapour and the liquid stream system reveals information about how much steam has to be condensed to fulfil constant molar flow for both (liquid and vapour) subsystems, respectively how much additional vapour has to be generated by the reboilers in order to reach this balance within the columns.

The grey cells in table 8 represent the vapour moles that are supposed to be condensed within filter and distillation column, respectively are the liquid moles that have to be vaporized in rectification and oil column reboilers to achieve the correct molar stream balance. The released latent heat of the vapour condensation covers the energy to heat up possibly subcooled liquids as well as energy losses from the columns to the environment.

The reboiler duty which is presented here can not be compared with other data, as absolutely no information about the reboiler specification was available and as additionally of course also the energy for rectification and oil column heat losses has to be provided, which will increase the reboiler duty.

However the difference in percent, which is calculated under the total balance, illustrates the deviation, as liquid and vapour balance have to be fulfilled at the same time. It can be clearly seen, that the deviation is maximal 0.27 %, which seems to indicate a very good accuracy rate achievement.

The energy balancing chapter will show how far these results really comply with other results from the energetic point of view.

As already mentioned above, the process conditions at MAF are not very stable and a long term database for composing reliable mass balances is not yet available. To get some clue whether the measured and calculated balances are representative for the real situation, comparative column simulation models should show if the gained data are in an acceptable accuracy range or not.



9.2 Column simulation

9.2.1 Vapour-Liquid-Equilibrium (VLE) calculation

The initial step in calculating or analyzing a binary distillation system is the computation of the vapour-liquid equilibrium curve.

The determination of the polynomial equation describing the vapour-liquid-equilibrium is an iterative procedure, starting with the fixation of the column pressure drop. An arbitrary number of 1000 is chosen to describe the equilibrium stages in the column.

The saturated pressure of each pure component existing in the system is calculated with the Antoine equation:

$$\begin{aligned} P_{\text{ethanol,sat}} &= 10^{[A-B/(T+C)]} \\ P_{\text{water,sat}} &= 10^{[A-B/(T+C)]} \end{aligned} \quad (28)$$

$P_{\text{ethanol,sat}}$ and $P_{\text{water,sat}}$ denote the saturated vapour pressure [mmHg] for the pure substances of the binary system ethanol-water. A, B, and C are the Antoine constants which are component specific and available from reference literature (e. g. [21]) and T [°C] describes the corresponding temperature.

The mole fractions of ethanol (x_e for liquid phase and y_e for the vapour phase) are found using the total system pressure P_{total}

$$x_e = \frac{P_{\text{total}} - P_{\text{water,sat}}}{P_{\text{ethanol,sat}} - P_{\text{water,sat}}} \quad (29)$$

$$y_e = \frac{x_e P_{\text{ethanol,sat}}}{P_{\text{total}}} \quad (30)$$

These equations would be sufficient for ideal systems, for a non-ideal system however activity coefficients (γ), which express the interactions between the molecules of the mixture, have to be considered in calculating the liquid and vapour compositions.

For ideal systems ($\gamma = 1$) Raoult's Law can be applied:

$$p_i = x_i p_i^s \quad (31)$$



In reality however it is common to calculate with ideal gas phase and real liquid phase for systems with not too high pressures. Following equations can be used with sufficient accuracy:

$$p_i = \gamma_i x_i p_i^s \quad (32)$$

To calculate the activity coefficient γ_i , the Wilson equation has been applied. The Wilson method is based on the concept of local composition and recommended for liquid mixtures that contain polar or associating components and therefore show deviations from ideal behaviour, as it is the case for alcohol-water mixtures.

$$\begin{aligned} \ln \gamma_1 &= -\ln(x_1 + \Lambda_{12}x_2) + x_2 \left(\frac{\Lambda_{12}}{x_1 + \Lambda_{12}x_2} - \frac{\Lambda_{21}}{x_2 + \Lambda_{21}x_1} \right) \\ \ln \gamma_2 &= -\ln(x_2 + \Lambda_{21}x_1) - x_1 \left(\frac{\Lambda_{12}}{x_1 + \Lambda_{12}x_2} - \frac{\Lambda_{21}}{x_2 + \Lambda_{21}x_1} \right) \end{aligned} \quad (33)$$

$$\begin{aligned} \ln \gamma_1^\infty &= -\ln \Lambda_{12} - \Lambda_{21} + 1 \\ \ln \gamma_2^\infty &= -\ln \Lambda_{21} - \Lambda_{12} + 1 \end{aligned} \quad (34)$$

$$\Lambda_{ij} = \frac{v_{0j}^l}{v_{0i}^l} \exp\left(-\frac{\lambda_{ij} - \lambda_{ii}}{RT}\right) \quad (35)$$

Λ_{ij} Wilson equation constants

Index	Component	Molar volume (v) [for T = 293 K] [cm ³ /mol]	Wilson Parameter	
			$\lambda_{12} - \lambda_{11}$ [J / mol]	$\lambda_{21} - \lambda_{22}$ [J / mol]
1	Water	18.07	3815	1755
2	Ethanol	58.68		

R gas constant (8.31447 J / K·mol)

The molar fractions of the single components in liquid and vapour phase are finally calculated using following correlations:

$$x_i = \frac{P_{total} - p_2^s \gamma_2}{p_1^s \gamma_1 - p_2^s \gamma_2} \quad (36)$$

$$y_i = \frac{P_{total} - p_2^s \gamma_2 * \frac{p_1^s \gamma_1}{P_{total}}}{p_1^s \gamma_1 - p_2^s \gamma_2} \quad (37)$$



As the temperature profile is unknown at the beginning, this calculation scheme has to be solved iteratively until the sum of the partial pressures divided by the total pressure prevailing at the respective stage equals 1:

$$\frac{\sum_i p_i}{P_{total}} \equiv 1 \quad (38)$$

The matlab algorithm to generate the vapour-liquid-equilibrium line (VLE), adapted for the real column pressure at MAF is VLE_Wilson.m, which is attached on CD.

9.2.2 Vapour-Liquid-Equilibrium (VLE) interpolation

Having enough x_i and y_i data for the binary equilibrium system, the single points can be plotted and a trendline can be generated for further calculations. The trendlines have been developed for each column according to the actual pressure drop profile, but as expected the pressure has negligible influence in the equilibrium behaviour, so one function describing the vapour-liquid-equilibrium line for all columns was taken. Analyzing the equilibrium curve shows that a polynomial representation of the data points, makes it necessary to fractionate the whole curve into sections in order to get reasonable accuracy for the whole curve.

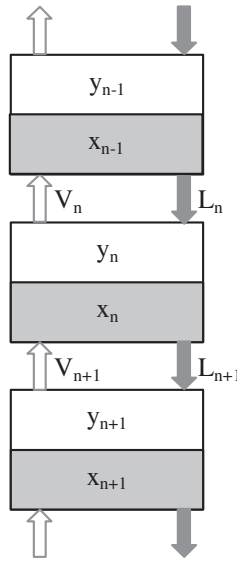
As a result a division into three sections, leading to three polynoms of sixth order, describe the equilibrium curve, to achieve correlation values of $R^2 = 1$ for each section.

The matlab algorithm to generate the function for the equilibrium line is VLE_interpolation_Wilson.m, again attached on CD.

9.2.3 McCabe-Thiele model

Within the column there is permanent mass and heat transfer between the raising steam and the down-flowing liquid. The fractional distillation is a multistage concept, where the equilibrium between vapour and liquid is supposed to be reached at every stage of the column, so that through counter-current flow of the vapour and the liquid stream a separation of components with different boiling points can be achieved. This equilibrium stage model (figure 25) is used to formulate the material and component balance across a single stage. Hereby constant molar flows are assumed, which is reasonable for ideal mixtures when the components have similar molar vaporization heats, which is given for the system ethanol-water (ethanol 39 kJ / mol; water: 40.7 kJ / mol).





L describes the liquid molar flow rate, while V represents the vapour molar flowrate. y_i is the mole fraction of species i in the vapour and x_i is the mole fraction in L . The number of moles of component i on stage n is denoted as $N_{i,n}$ and can be expressed with the material balance for component i at stage n as follows:

$$\frac{dN_{i,n}}{dt} = (L_n x_{i,n-1} - V_n y_{i,n}) - (L_{n+1} x_{i,n} - V_{n+1} y_{i,n+1}) \quad (39)$$

For a steady-state operation: $\frac{dN_{i,n}}{dt} = 0 [mol s^{-1}]$. (40)

figure 25: equilibrium stage model

Rewriting the equation into a form that the vapour composition ($y_{i,n}$) becomes a function of $x_{i,n+1}$, results in the curve known as operating line.

$$y_{i,n} = \frac{L_n}{V_n} x_{i,n-1} + \frac{V_{n+1} y_{i,n+1} - L_{n+1} x_{i,n}}{V_n} \quad (41)$$

9.2.4 Mass balance

The overall mass balance for a simple distillation column (figure 26) includes three streams:

- F feed
- D distillate
- B bottom product

$$F = D + B \quad (42)$$

The component balance for component i is given as:

$$F z_{i,F} = x_{i,D} * D + x_{i,B} * B \quad (43)$$

- $z_{i,F}$ mole fraction of component i of feed-stream
- $x_{i,D}$ mole fraction of component i of distillate-stream
- $x_{i,B}$ mole fraction of component i of bottom product-stream



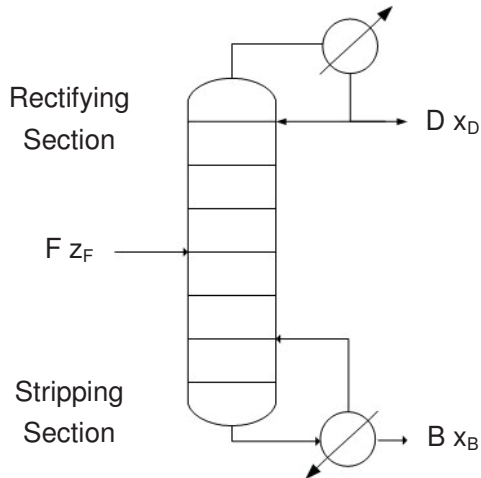


figure 26: simple distillation column

The model describing the column consists basically of two sections: the part above the feed is the rectifying (enrichment) section, while the stripping (removal) part is located below the feed. Both areas have to be investigated separately.

The overall component balance for the rectifying section can be written as:

$$V_n y_{n,i} = L_n x_{i,n-1} + D x_{i,D} \quad (44)$$

Assuming constant molar flow, the reflux ratio R can be defined as:

$$R = \frac{L}{D} \quad (45)$$

NOTE:

Although most of the distillation columns are designed in a manner that the reflux is a saturated (bubble point) liquid, in real columns this must not always be the case, as the reflux might enter the column as subcooled liquid. Especially if the condenser arrangement starts with a partial condenser and is followed by a total condenser, the liquid stream leaving the second condenser can be below the bubble point temperature, depending on how the cooling medium is adjusted. If such subcooled reflux is entering the top stage of the column it causes upwards flowing vapour to condense. As the condensation process releases latent heat, it provides the sensible heat needed to heat up the subcooled reflux to the bubble point. Additionally this causes more liquid flowing down the column, so the “internal reflux ratio” is higher than the “external”, which is only determined by the condenser. Hence in case of subcooled liquid the external reflux ratio has to be replaced by the internal reflux ratio, otherwise the computed number of equilibrium stages will be higher than actual required.



Combining the overall balance and the reflux ratio gives following relation between the entering vapour and liquid flows for the rectification section and an equivalent equation for the stripping section:

Rectification section	Stripping section
$y_{i,n} = \frac{R}{R+1} x_{i,n-1} + \frac{1}{R+1} x_{i,D} \quad (46)$	$y_{i,m-1} = \frac{L}{V} x_{i,m} - \frac{B}{V} x_{i,B} \quad (47)$

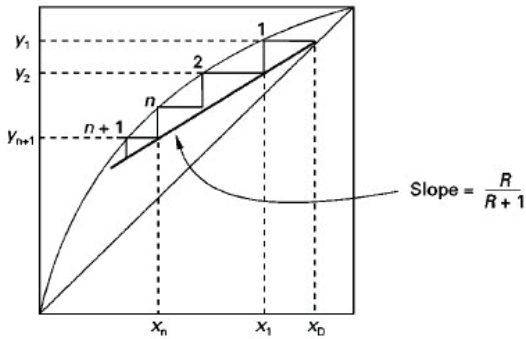


figure 27: mass balance – rectifying section

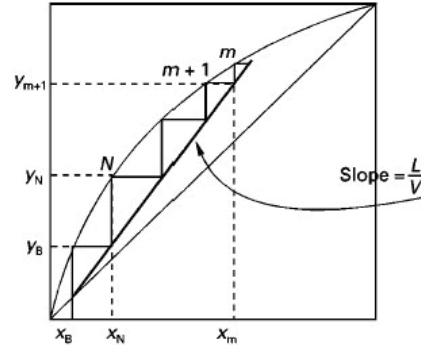


figure 28: mass balance – stripping section

The rectifying and stripping lines intersect at the feed stage. The location of this point depends on the conditions of the feed (sub-cooled, saturated liquid, partially vaporized, saturated vapour or superheated vapour).

Hence with the feed condition, the actual intersection point between the stripping and rectifying operation lines is given and the system is determined. To describe the thermodynamic state of the feed stream the variable q is introduced:

$$q = \frac{\text{heat required to vaporize 1 mol of feed}}{\text{molar latent heat of vaporization of feed}} \quad (48)$$

sub-cooled	$h_F < h'$	$q > 1$
saturated liquid	$h_F = h'$	$q = 1$
partially vaporized	$h' < h_F < h''$	$0 < q < 1$
saturated vapor	$h_F = h''$	$q = 1$
superheated vapor	$h_F > h''$	$q < 1$

Combining equation (44) with (46) and (47) leads to the equation for the q -curve:



$$y = \frac{q}{q-1}x - \frac{1}{q-1}z_F \quad (49)$$

Plotting the operating line and the equilibrium line (VLE) in one diagram (figure 29), forms the basis of the McCabe-Thiele approach, which allows to determine and calculate the number of stages and the corresponding stage compositions if the flows in the system are known.

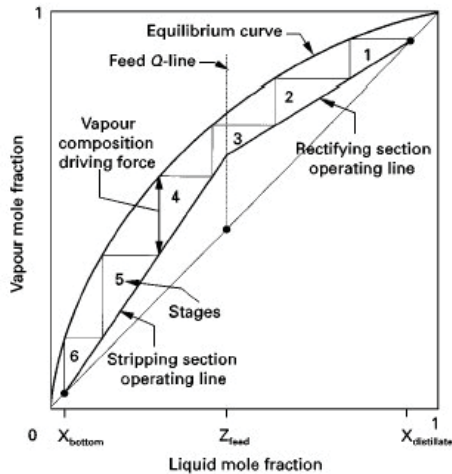


figure 29: typical McCabe-Thiele diagram

With given operating lines, an algorithm has to be found to handle the graphical display and counting of equilibrium stages in the simulation program in a convenient way. For this purpose the particular intersection points are determined and linear equations between the distillate, intersection and bottom points are formulated. An intersection point occurs, either at a feed stage or a side stream stage and is given as follows:

$$\begin{aligned} y_{\text{stripping section}} &= y_{\text{rectification section}} = y_{\text{intersec t}} \\ x_{\text{stripping section}} &= x_{\text{rectification section}} = x_{\text{intersec t}} \end{aligned} \quad (50)$$

$$\begin{aligned} y_{\text{rect}} &= \frac{L}{V}x + \frac{D}{V}x_d = \left(\frac{R}{R+1}\right)x + \frac{x_d}{R+1} \\ y_q &= \left(\frac{q}{q-1}\right)x - \frac{x_F}{(q-1)} \\ \Rightarrow x &= \frac{x_F - (1-q)y}{q} \end{aligned} \quad (51)$$



$$\Rightarrow y_{rect} = \left(\frac{R}{R+1} \right) \left(\frac{x_F - (1-q)y}{q} \right) + \frac{x_d}{R+1}$$

$$y_{rect} \equiv y \equiv y_{intersect}$$

$$\Rightarrow y_{intersect} = \left(\frac{R}{R+1} \right) \left(\frac{x_F - (1-q)y_{intersect}}{q} \right) + \frac{x_d}{R+1} \quad (52)$$

$$\Rightarrow y_{intersect} = \frac{Rx_F + x_d q}{(R+q)}$$

$$y_{rect} = y_q$$

$$\left(\frac{R}{R+1} \right) x + \frac{x_d}{R+1} = \left(\frac{q}{q-1} \right) x - \frac{x_F}{(q-1)}$$

$$\Rightarrow xd \equiv x \equiv x_{intersect}$$

$$\Rightarrow x_{intersect} = \frac{\left(\frac{x_F}{1-q} - \frac{x_D}{R+1} \right)}{\left(\frac{R}{R+1} + \frac{q}{1-q} \right)}$$

9.2.5 Multiple feed and side-streams, direct steam injection

Plotting the operating lines and the equilibrium line in one diagram, the McCabe-Thiele method can now be used for binary systems to “step off” the number of necessary equilibrium stages to meet the specified top and bottom product compositions.

So far only one feed and one withdrawn stream (at the bottom) have been considered. In the case of the distillation plant at MAF however several feed and side streams, as well as open steam injection, have to be implemented in the model.

The slope of the operating lines is generally given by the ratio between liquid and vapour streams, which is different for the stripping and rectifying section as the feed condition (q) determines whether L will change or V or both streams. For each intersection between a feed or side stream a separate operating line has to be created, which is based on the actual vapour and liquid flow rate from this specific part of the column onwards. So the intersection calculation as described previously has to be applied at each feed or sidestream stage, as the slope of the operating line will change from this point onwards.

If a reboiler is used to create the vapour stream, the composition of the vapour at the bottom stage is a function of the leaving bottom liquid stream (according to the equilibrium condition). However if pure steam is injected, the starting point of the operating line at the bottom is given by $x_b = x_b$ and $y_b = 0$, which leads to a situation where the operating line actually crosses the 45° line.



9.2.6 Stage and column efficiency

The McCabe-Thiele method assumes that the two phases leaving each stage of the column are in equilibrium with respect to both heat and mass transfer. For the temperature change it is often justified to assume that vapour and liquid phase are leaving at the same temperature, as long as there are no large temperature changes prevailing between the stages. But especially the assumption for the equilibrium with regard to mass transfer between vapour and liquid phase is not always justified in industrial counter-current, multistage distillation columns, as residence time and contact between the phases might not be ideal. Thus the theoretical predicted equilibrium concentration is higher than the actual stage changes. To consider these real conditions, the stage efficiency has to be implemented. The stage efficiency can be defined of either phase and, for a given component, equals the change in actual composition in the phase, divided by the predicted equilibrium change.

For the efficiency term three types of plate efficiencies can be used [22]:

- Overall efficiency (for the entire column)
- Murphy efficiency (based on a single plate)
- Local efficiency (for a specific location on a single plate)

The overall efficiency simply describes the theoretical required (ideal) stages compared to the actual used number of plates in the whole column. Determining the overall efficiency can easily be achieved, by taking samples at the top and bottom of the column and comparing the theoretical required stages according to the McCabe-Thiele method and the actual used plates at the industrial site.

The Murphy efficiency for the vapour phase is defined by:

$$\eta_M = \frac{y_n - y_{n+1}}{y_n^* - y_{n+1}} \quad (54)$$

y_n actual concentration of vapour leaving plate n

y_{n+1} actual concentration of vapour entering plate n

y_n^* concentration of vapour in equilibrium with liquid phase from plate n

The one used for the simulation files for the columns at MAF is based on the liquid phase:

$$\eta_M = \frac{x(n) - x(n+1)}{x(n) - x_{eff}(n+1)} \quad (55)$$

$$\Rightarrow x_{eff} = x(n) - \frac{x(n) - x(n+1)}{\eta_M}$$



$x(n)$	composition of liquid at stage n
$x(n+1)$	composition of liquid at stage $n+1$ predicted by the equilibrium
η_M	Murphy efficiency
x_{eff}	actual composition achieved at the stage.

The estimation of plate efficiency is usually empirically based (see [22]) and the actual performance data have to be obtained from the industrial site.

According to literature [23], the overall stage efficiency has been found to be a complex function of

- Geometry and design of the contacting trays
- Flow rates, paths and patterns of vapour and liquid streams
- Composition and properties of vapour and liquid streams

In well-designed trays the flowrates are usually near the capacity limit, so that the efficiency is mainly influenced by the vapour and liquid stream properties. In that case four methods can be used to give a prediction [23]:

- Comparison with performance data from industrial columns for the same or similar system
- Use of empirical efficiency models derived from data on industrial columns
- Use of semi-theoretical models based on mass and heat transfer area
- Scale-up from data obtained with laboratory or pilot plant columns

For binary distillation systems, especially of close-boiling mixtures and a low liquid viscosity, stage efficiencies for well designed trays and optimal operating conditions are usually in the range between 70 – 100 %.

At MAF the actual plate efficiency could not be determined with the available means due to limitations in sample taking from intermediate stages.



9.2.7 McCabe-Thiele models for MAF columns

Two different types of McCabe-Thiele models have been developed in the programming software *Matlab 6.1*. All *Matlab* files mentioned in the following paragraphs are attached on CD.

9.2.7.1 Illustration of the actual situation (stagewise model – see appendix)

distillation_real.m
filter_real.m
rectification_real.m
oil_real.m

The only gained result from this column modelling was that the feed location of single streams is not optimized. The data verification for mass flowrates and concentrations of the single streams was not feasible, as the column efficiency could not be estimated in a satisfactory manner. For this purpose another simulation file was created, that considers the optimal feed location¹, enabling the calculation of a minimum required overall efficiency under given tray numbers for each column, assuming the feed stages would comply with the internal stage compositions.

9.2.7.2 Consideration of the correct feed location

McCabeThieleDISTILLATION.m
McCabeThieleFILTER.m
McCabeThieleRECTIFICATION.m
McCabeThieleOIL.m

Initially an estimated overall efficiency of 70 % was implemented by using the Murphee efficiency model. Out of the models with the correct feed location and the number of existing stages in each column, the minimum required efficiency of the trays that has to be achieved, if the in- and output stream specifications are supposed to be kept, was calculated. This efficiency value however does not give valuable information, as it cannot be checked with the existing column situation. Also the efficiency usually varies from tray to tray and a separate investigation of the columns within the process optimization approach is necessary for a profound statement.

¹ The feed should in general be introduced into the column where the temperature and composition roughly match the column profile. Modelling an existing column, the feed stage in the diagram should correspond as closely as possible to the actual feed stage in the column. For badly dislocated feed stages the effectiveness of the stage is not reached at all as the leaving streams will not be in equilibrium to each other, instead will rather remain close to the operating line.



Result

The column modelling could as already pointed out not be used for the verification of the mass balance. What came out clearly is improper feed stage location. Especially the filter column shows that without modification a simulation with the equilibrium stage model is not applicable (compare figure 30 and figure 31 for efficiency $EFF=0.325$). For the details of how the algorithm has been modified to avoid the exceeding of the equilibrium line it shall be referred to the appendix at this point.

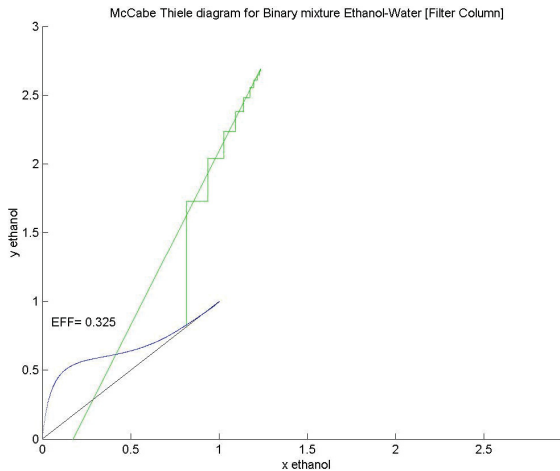


figure 30: McCabe Thiele – filter column real

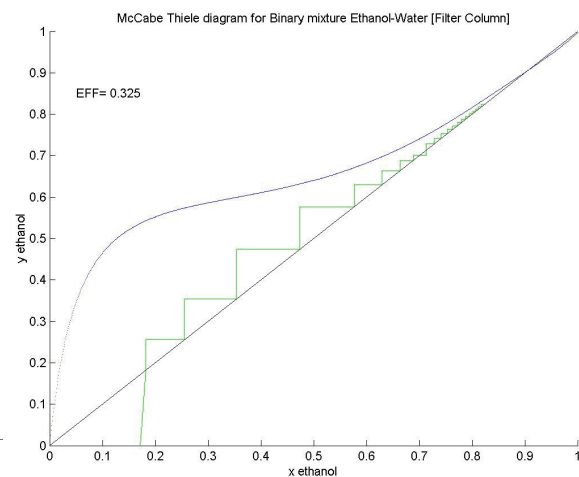


figure 31: McCabe Thiele – filter column modified

Nevertheless the algorithm can find usage in the future for predicting bottom composition with varying input data. This however is only reasonable once proofed efficiency data are available and optimized feed stages are established.

The importance of the efficiency task shall be briefly illustrated by comparing two different efficiency values that were used in the real oil column modelling (table 9).



table 9: Oil column modelling – different efficiency assumptions

Oil_real (Efficiency = 0.7)			Oil_real (Efficiency 0.165)		
stage	x_e	y_e	stage	x_e	y_e
1	0.5658	0.5689	1	0.5658	0.5689
2	0.3357	0.3417	5	0.3378	0.3439
3	0.1346	0.1433	10	0.1568	0.2038
4	0.0499	0.0597	15	0.0733	0.1295
5	0.0185	0.0287	20	0.0357	0.0962
6	0.0071	0.0175	25	0.0191	0.0814
7	0.0031	0.0672	30	0.0117	0.0749
8	0.0049	0.0688	35	0.0085	0.0720
9	0.0056	0.0694	40	0.0070	0.0699
10	0.0058	0.0696	45	0.0064	0.0695
11-35	0.0059	0.0697	50	0.0061	0.0693
36-54	0.0059	0.0692	54	0.0060	0.0692

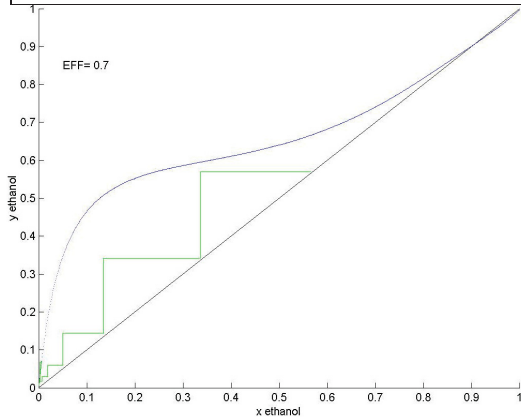


figure 32: Oil column – efficiency 0.7

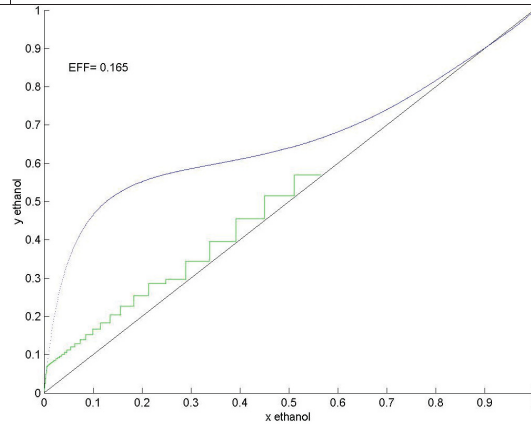


figure 33: Oil column – efficiency 0.165

For efficiency 0.7 the stream composition is almost not changing from stage 10 onwards, while similar results with plate efficiency 0.165 are only reached from stage 50 onwards. In both cases top and bottom specifications are the same, but the internal equilibrium processes are varying a lot. Theoretical, local efficiencies could have been implemented at the withdrawn sidestream stages, as the stage composition of the stream and the sidestream have to be the same. This however would have only been another theoretical value, without practical justification, as due to improper feed location the overall system does not have the same efficiency values throughout the column anyway.

To summarize the modelling part it has to be said that data verification, as initially planned, was not feasible. It became however obvious that there is a general need for further investigations regarding proper feed location.



10 Energy balance

10.1 Steam system

The fuel oil consuming process at MAF is the generation of steam, which is used in the distillation sector.

The investigation of energy optimization potential in the steam system starts with the breakdown of the system into three parts:

- Steam generation Boiler
- Steam distribution system Pressure reduction stations, steam traps, piping and insulation
- Steam utilization system End use equipment that consumes steam

Each of these subsystems reveals its own improvement potential and will be discussed separately in the following.

10.2 Steam generation

10.2.1 Boiler efficiency

Boiler efficiency in general describes the ratio between energy input and energy output. For industrial boilers with high annual fuel usage, even seemingly small improvements in the efficiency can achieve substantial savings.

To assess the energy efficiency of industrial boilers it has to be understood right at the outset that efficiency is a highly boiler-specific characteristic and an accurate comparison has to be based on proven performance.

The specifications and guarantees in terms of efficiency for new boilers are generally based on certain design conditions (fuel, load, temperature, pressure). Any deviation of these specified characteristics will have impact on the actual boiler efficiency so that the theoretical value given by the boiler manufacturer cannot be achieved.

Altering operating conditions will therefore greatly impact the efficiency terms and will definitely result in different efficiency figures.



The main influencing factors determining the energy efficiency of a steam boiler system are:

- Fuel characteristics
- Combustion system limitations (firing system and heat input)
- Equipment design (heat transfer)
- Steam system operation requirements

Having these influencing parameters to increase the efficiency of a boiler, there are still constraints that will limit the improvement of an already installed boiler, as technical possibilities do not necessarily have to be justified in economical terms. To change the equipment design itself for instance, is therefore usually not reasonable and in general task of the boiler designer and manufacturer. For an existing industrial site the steam requirements are given as well, and improvement potential within the steam system involves a whole process analysis.

10.2.1.1 Fuel type

Once a boiler is installed it is often not convenient and cost effective to adapt an existing system in order to utilize different fuels. This however does not mean that the fuel is not an influencing parameter in the approach to increase the efficiency of a prevailing steam generation facility. A variation of fuel composition [moisture (H_2O), carbon (C), hydrogen (H), nitrogen (N), chlorine (Cl), sulphur (S), oxygen (O) and ash] due to changing of suppliers and/or insufficient standards that guarantee a certain quality will directly affect the boiler efficiency.

As for the case of Mekanissa Alcohol Factory the choice of oil to purchase is limited and practically only one type of oil (fuel oil light) can be used. Fuel oil light is imported from Djibouti and provided by several suppliers in Ethiopia. The actual decision which trader is contracted is exclusively made on price benefits and not on quality fluctuations. If under very rare circumstances a shortage of oil supply occurs, heavy fuel oil has to be purchased and mixed with naphtha on the companies' compound.

10.2.1.2 Combustion system limitations (firing system and heat input)

Combustion system limits are especially given by environmental legislation, which demands to keep NO_x -, SO_x -, POPs (persistent organic pollutants) or other critical emissions at a low level. Ethiopia has not yet developed laws that prohibit off-gas emissions to the atmosphere, so monitoring equipment for air-pollutants is not installed. This however does not only have environmental impacts but also influence on the combustion efficiency. A big potential for increasing the combustion efficiency lies in the optimization of air for combustion:

Lowering the amount of combustion air leads to incomplete combustion and consequences in increased unburned-fuel loss. Maximum combustion efficiency on the other hand



correlates with rising excess air and parallel increased flue-gas temperatures, which results in higher losses through the stack. This interplay between sufficient but not too much excess air entails the control of the flue gas to keep it out of the combustible range via monitoring the CO level, which indicates the completeness of any combustion, and at the same time to keep the oxygen content in the flue gas at an acceptable level. The measurement of oxygen shows the amount of excess air used, and should not be above the recommended level, which is given from the supplier based on given installation, type of burner and fuel fired. If it exceeds this specification, the heat content from the fuel is wasted to heat up this extra air instead of going into the production of steam. A monitoring system that regulates the air-to-fuel load is therefore an investment that does not only benefit the environment but also pays off in terms of fuel savings and control mechanism for keeping the process conditions constant.

When discussing the flue gas stream, another potential for fuel-saving in existing heaters evolves: the utilization of heat contained in the flue gas. The installation of heat exchangers to recover the heat loss carried away with the flue gas is a common practice for either air heating systems or feed preheating purpose. This energy potential however cannot be exploited to the maximum extent, due to high dew points of the flue gas caused by SO_3 formation (corrosion problems) from the sulphur in the fuel (figure 34).

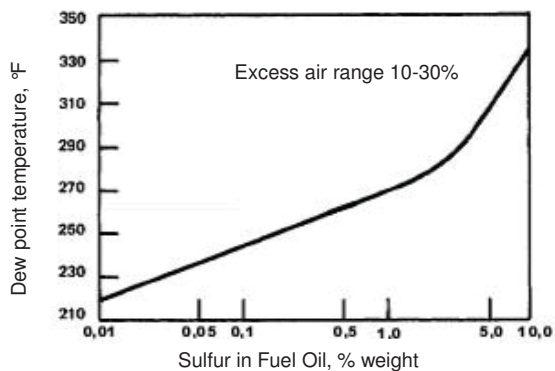


figure 34: maximum flue-gas dew point versus percent of sulphur in typical oil fuels [24]

The fuel oil used at Mekanissa Alcohol Factory contains max. 0.3 %wt. sulphur, which allows the utilization of the energy in the flue-gas until about 127 °C. State of the art industrial boilers are equipped with an economizer, which is extracting the off-gas-heat to preheat the feedwater.

At MAF the amount of excess air entering the boiler is one of the unknown parameter. Furthermore the fuel control (load control) occurs by manual adjustment. It therefore could not be determined what the actual optimization potential of the old boiler section would be. Literature [25] provides comparative figures that point out that old stokers with no air controls and fixed airflow, can reach even 200 % excess air. Considering this extremely high



combustion air ratio a significant drop of boiler efficiency will be the consequence and can even go down to 60 %.

However as the old boilers will be replaced by a new facility in the coming years anyway, this energy recovery potential has been skipped for the time being and the only investigation undertaken was the determination of the fuel oil consumption to get a picture about the current situation, without identifying where the actual losses occur.

10.2.1.3 Steam system operation requirements (water-steam circulation, preheating, superheating)

Steam-water circulation describes the part where water or steam or a mixture of it flows through the boiler, aiming to generate steam by carrying off heat from the tube metal. This heat absorption must be kept at a level that ensures sufficient cooling of the furnace-wall tubes during operation conditions. Is the circulation flow of the water-steam system not adjusted to the actual situation, malfunctions due to overstressing, overheating or corrosion will occur. Scale formation and maintenance status are the critical factors that deteriorate the heat transfer itself, which is mainly influenced by the heat flux and the turbulence. A scale layer of only 0.8 mm in a boiler can decrease the efficiency up to 6-7 % [26].

At Mekanissa Alcohol Factory one of the boilers has been extremely affected by scaling and as consequence some of the clogged up tubes had to be removed. Now more attention is paid on regular quality checks of the water after the softener unit and it is made sure, that the hardness of the water is kept at the demanded level.

10.2.2 Efficiency term

When talking about boiler efficiency, a distinction between different terms that are used to qualify efficiency in the context with boilers has to be made. Boiler, thermal, combustion or fuel-to-steam efficiency are often used to describe the efficiency.

Simply efficiency or boiler efficiency does not give valuable information about the performance of the boiler as it is unknown which losses are actually included in the calculations. Thermal efficiency is an expression usually found for heat exchangers, but has less significance for the comparison of boiler activity. It can be used to measure the ability of the exchanger to transfer combustion heat to the water and/or steam in the boiler. However this single part of the boiler does not represent the overall boiler efficiency and is not a useful indicator for fuel usage.

The relevant efficiency terms used for boilers are fuel-to-steam efficiency and combustion efficiency.



Combustion efficiency is an indicator for the fuel burner and is determined as ratio between the total heat released in combustion minus the heat loss in the flue gas (FG), divided by the total heat released. This term is mainly influenced by unburned fuel and excess air in the flue gas.

$$Q_{FG} = m_{FG} * c_{p,FG} * (T_{FG} - T_{ambient\ air}) \quad (56)$$

$$\eta_{combustion} = \frac{Q_{in} - Q_{FG}}{Q_{in}} \quad (57)$$

η	efficiency
Q	energy flowrate
L_{RC}	factor that considers the losses through RC
RC	radiation and convection
FG	flue gas
in	input
out max	maximal output

It is obvious that low flue gas temperatures will lead to higher efficiency results. Also the influence of the ambient air becomes apparent now and has even stronger impact on radiation and convection losses from the boiler itself, when comparing boilers outside or in a boiler room.

Fuel-to-steam efficiency gives the actual ratio of energy input that is finally converted to steam and therefore gives the required information for assessing a steam boiler. The fuel-to-steam efficiency is calculated like the combustion efficiency but additionally considers the RC-losses: the heat radiating from the boiler (R-losses) and heat losses caused by air flowing around the boiler (C-losses). Throughout the firing range of a boiler the RC-losses are constant and variations depend mainly on boiler type, size and operating pressure.

To calculate this fuel-to-steam efficiency term two methods can be applied [27]: the Input-Output Method and the Heat Loss Method.

$$\eta_{fuel\ to\ steam} = \frac{Q_{in} - Q_{losses}}{Q_{in}} \quad (58)$$

$$\eta_{fuel\ to\ steam} = \frac{Q_{in} - Q_{RC} - Q_{FG}}{Q_{in}} = \frac{Q_{steam}}{Q_{in}} \quad (59)$$

$$Q_{RC} = L_{RC} * Q_{out\ max} \quad (60)$$



Measuring losses of the boiler includes the inquiry of following points:

- Sensible heat in the flue gas
- Unburned carbon monoxide in the flue gas
- Radiation
- Safety valves (cannot properly be measured)
- Blowdown¹⁾

Blowdown does not only cause heat losses, but also water losses if vapour and water are not collected and reused after re-treatment. If the blowdown water is simply discharged it therefore accumulates costs for the water itself, for the fuel oil that was used to produce the steam and also the costs for the water treatment in the softener unit.

At MAF blowdown is carried out every twelve hours. How much water is actually discharged could not be determined, as the blowdown enters the drainage pipe, which leads to the nearby river.

Out of these losses that determine the boiler efficiency none could be measured or calculated completely, hence only the input-output method was feasible and the evaluation of the boiler performance was based on following gathered information:

- Fuel consumption
- Steam output
- Feed water temperature
- Temperature, pressure and quality of the steam

An additional factor that has to be considered at MAF is the water preheating system. To preheat the feed water, the original plant layout included the utilization of condensed reboiler steam from the distillation sector. Due to breakdowns of pumps however this energy potential is not used and direct steam injection is used to preheat the water from 25 °C to 69 °C.

The temperature rising from 69 °C to boiling point (92 °C at the plant location due to the altitude) is taking place in the boiler. Saturated steam will be provided at 9 bar (175 °C).

¹ Blowdown is generally described as a scale-producing and sludge-bearing liquid and may give troubles on heat exchanger surfaces. Blowing down describes the process of removing boiler water, which has a maximum acceptable level of concentration on a regular basis in order to control the level of total dissolved solids.



Water preheating

$$\begin{aligned} \dot{m}_{H_2O, dist\ steam} &= (\dot{m}_{H_2O, feed} - \dot{m}_{H_2O, preheat\ steam}) \\ (\dot{m}_{H_2O, feed} - \dot{m}_{H_2O, preheat\ steam}) * c_{p, H_2O} * \Delta T_1 &= \dot{m}_{H_2O, preheat\ steam} * h_v + \dot{m}_{H_2O, preheat\ steam} * c_{p, H_2O} * \Delta T_2 \\ (2200\ kg - \dot{m}_{H_2O, preheat\ steam}) * 4.179\ kJ / kg\ K * (342\ K - 298\ K) &= \\ &= \dot{m}_{H_2O, preheat\ steam} * 2030\ kJ / kg + \dot{m}_{H_2O, preheat\ steam} * 4.207\ kJ / kg\ K * 44\ K \\ \dot{m}_{H_2O, preheat\ steam} &= 175.07\ kg \\ \dot{m}_{H_2O, dist\ steam} &= 2024.93\ kg \end{aligned}$$

$\dot{m}_{H_2O, feed}$	boiler feed water (total water entering the boiler; measured with flowmeter)
$\dot{m}_{H_2O, dist\ steam}$	steam going to distillation section
c_{p, H_2O}	specific heat capacity water
ΔT_1	temperature difference fresh water T – boiler feed T
$\dot{m}_{H_2O, preheat\ steam}$	steam required to raise temperature of boiler feed water
h_v	specific enthalpy
ΔT_2	temperature difference steam condensation T – boiler feed T

Boiler

Utility steam generator circulation systems are generally categorized as once-through flow. This means that under normal operating conditions the total feed-water flow equals automatically the generated steam flow, as the feed-water is absorbing heat from the tube surface until it is completely converted to steam. The water consumption and steam generation could be examined separately for both boilers but the fuel oil consumption could only be measured by observing the level change in one of the oil storage tanks by closing the rest of the feeding pipelines and keeping the level in the intermediate storage bin constant. The actual oil feeding rate is adjusted manually by the operators but is usually kept constant after the start-up phase.

Fuel oil consumption and energy input

The measured fuel oil consumption [at 25 °C] is: 3,979.32 l/day

$$\begin{aligned} Q_{in} &= V_{oil} * \rho_{oil} * calorific\ value \\ Q_{in} &= 3,979.32\ l / day / 24 * 0.890\ kg / l * 43,031\ kJ / kg \\ Q_{in} &= 6.349,932\ kJ \end{aligned}$$



10.2.3 Calculated boiler efficiency

$$\eta_{SG} = \frac{Q_{out}}{Q_{in}}$$

$$Q_{out} = m_{steam} * (h_{steam} - h_{feedwater})$$

$$Q_{out} = 2,200kg * (2,773.9 - 288.9)kJ / kg$$

$$Q_{out} = 5.467,000kJ$$

$$Q_{in} = 6.349,932 kJ / kg$$

$$\eta_{SG} = \frac{5.467,000}{6.349,932} * 100$$

$$\eta_{SG} = 86.1 \%$$

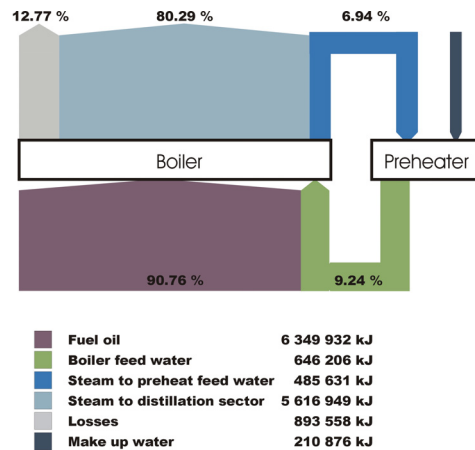


figure 35: Energy – Sankey Diagram for boiler section

figure 35 gives an overview about the energy flowrates of the single streams in the boiler section.

The new boiler that should be installed in the coming years has nominal values of about 90 % efficiency and will produce 6 000 kg steam per hour. This increased capacity of the boiler goes along with the planned reactivation of a part of the factory that is out of work at the moment. A smaller distillation plant that will also produce ethanol (96.6 % vol), should be repaired again and the new boiler should cover the required steam supply as well.

The new boiler (Generator Steam Matic “SM”) is equipped with an adjustable fuel control valve as well as a control valve for the desired air ratio, which will help to come closer to the nominal efficiency. It is recommended to monitor the emitted CO₂ level in the flue gas (should be in the range 10-13 %) to find the optimal air supply amount, which makes it obvious that an off-gas monitoring system has to be installed as well, if good boiler efficiency should be achieved. Also a recording steam meter should be installed to enable the assessment of the boiler performance and to improve the distillation process control itself, too.

10.3 Approach for steam saving within the distribution and utilization system

The general approach to avoid wastage of steam (energy respectively) proposed by Oliver Lyle [28], follows a very logical pattern:

- Prevent the escape of heat – avoid any leakage in the system
- Reduce the work to be done – optimize the consuming process in a way to use minimum energy input for the required quality and quantity of the product
- Use the heat over again – energy recovery as far as economical justified



The prevention of heat losses can be analyzed by categorizing two kind of losses: primary heat losses are including radiation, leakages, safety valve blows and by-pass traps and secondary heat losses are dealing with flash steam, condensate, blowdown and hot waste effluent. If primary heat losses are identified, there is no argument that excuses the ignorance of the elimination of these losses. Proper insulation, avoidance of leakages and proper adjusted valves are not only an energy saving method but a general sign of a well kept plant and “good housekeeping”. When dealing with secondary heat losses, a situation might occur, where recovery might not pay-off in economical terms and the additional equipment cost or amount of additional work involved may exceed the benefits of heat re-utilization. This investigation therefore always requires a parallel economical consideration.

The reduction of work to be done contains a large range of influencing parameters, which are process specific and vary from site to site. For a distillation plant relevant areas to be scrutinized in the analysis of a binary system are [23]:

- Feed flow rate, composition, temperature, pressure and phase condition
- Desired degree of separation between the two components
- Operating pressure (relevant for the choice of the operating pressure are the available cooling medium and the vapour pressures of the components that have to be separated)
- Vapour pressure drop
- Minimum reflux ratio and actual reflux ratio (in industrial plants efficiency and low running costs are important parameters for keeping the factory profitable. In the case of distillation columns this can be achieved by working at the minimum possible reflux ratio, which requires good mass-flow control devices or reflux divider).
- Stage efficiency (minimum number of equilibrium stages versus actual number of stages)
- Type of condenser (total, partial or mixed condenser)
- Degrees of subcooling of the liquid reflux
- Type of reboiler (total or partial)
- Type of contacting (tray or packing column)
- Dimensions of the column (height, diameter, stage spacing)
- Feed entry stage

10.3.1 Steam distribution system

The steam distribution system is mostly a neglected part of the boiler system and therefore often hides significant potential for improvements.



The steam distribution system consists of:

- piping and its insulation
- valves
- pressure-reduction stations
- steam traps
- moisture separators

The steam pipes at MAF are continuously insulated with an asbestos fibre cover except the valves, pressure-reduction station and about 3 m from the junction of the main steam line to the distillation column. It would be advisable to remove the unused asbestos insulation from the line to the dismantled reboiler at the distillation column as it is of no use there unless the reboilers are reactivated again and to insulate the pipe to the distillation column instead.

Pressure-reduction stations effect saturated steam (condensation at the dew point) in a way that, through pipe expansion for instance, the pressure will be reduced while no work is done by the steam, so the actual heat content carried by the steam remains the same although pressure and temperature are decreased. Comparing the actual energy content of the steam at the two different pressure and temperature levels with the expected data from steam tables reveals that at fallen pressure and temperature – with still the same amount of energy carried by the stream - the steam is now classified as superheated.

This phenomenon to superheat saturated steam simply by reducing the pressure can be very useful in certain circumstances. Especially if the steam is very wet, it can cause big nuisances when it creates extra condensate on heat exchanger surfaces, as it has little heating value, but covers the exchanger surface with an additional water film, that has to be handled by the steam trap. To prevent the forming of unwanted condensate the superheat due to expansion will help to dry it and therefore increases the amount of steam reaching the utilization system.

At MAF the main pressure reduction station reduces the 9 bar steam to about 2 – 2.5 bar, still supplying saturated steam after the reduction, which was verified by temperature measurements and steam table comparison.

Condensate generated in the steam distribution system:

Condensed steam from the steam system is often simply drained, which is a considerable loss of energy and can be avoided by returning as much condensed (uncontaminated) steam as possible. Additionally the condensate has a low conductivity with few contaminants and has not to be pre-treated again, which allows direct recirculation into the boiler feed water. If the steam however is part of the process as it is the case at Mekanissa Alcohol Factory due to direct steam distilling, a separate steam energy recovery is not possible. Still the existing recovery opportunities from the reboiler facilities should be exploited to the maximum extend.



Pure condensate within the distribution (piping) system at MAF is not collected, as the amount of condensed steam from the steam pipes is negligible.

For an isolated (Asbestos) DN 100 pipe with the pressure of 9 bar and a length of 23 m (measured from the boiler house until the pressure reduction station), following steam loss can be calculated [29]:

The condensate load generated in insulated steam pipes for saturated steam in a 100 mm pipe and a pressure of 9 bar is given with 34.8 kg/h per 100 m pipe length and for an average surrounding temperature of 20°C. This means that for the distance of 23 m at MAF 8 kg/h condensate are generated which equals 0.3 % of the hourly steam production. Projecting this figure on an annual loss, 47 063 kg/a results for an average operating time of 245 days per year, which equals not even the steam production of one 24 hour operating day.

10.3.2 Steam utilization system

Factories are often not aware of how much steam is consumed by single units of the total process, respectively also do not know how much should be used. The investigation of the steam utilization system generally does not only reveal better knowledge about the energy consumption distribution but leads to a better understanding for the process in general and can also give insight into better process control.

At Mekanissa Alcohol Factory the steam utilization system consists of following areas:

- Direct steam distillation in four columns
- Additional reboilers at the rectification and oil column
- [Indirect steam distillation in one column (Demethylizing Column is not operating)]
- Boiler feed water preheating purpose (already described)

It has been already pointed out that this thesis is not dealing with operational changes and influences on the distillation conditions, such as pressure, temperature or reflux variations. Part of getting a proper energy picture of the utilization system however is, to determine at least the energy losses, occurring due to lack of column insulation.

Out of the mass balance and molar flow balance the theoretical steam demand of the single columns, reboiler duties as well as steam losses have already been identified. A crosscheck by determining convective and radiative column heat losses should indicate how far the two different approaches for the utilization losses calculation are corresponding.



10.4 Column heat losses

10.4.1 Modes of heat transfer

There are three fundamental types of heat transfer: conduction, convection and radiation, which all can occur simultaneously. The definition of those types is as follows [24]:

“**Conduction** is defined as the transfer of heat from one part of a body in another part of the same body, or from one body to another in physical contact with it, without appreciable displacement of the particles of the body.

Convection is the transfer of heat from one point to another within a fluid, gas, or liquid by the mixing of one portion of the fluid with another. In natural convection, the motion of the fluid is entirely the result of differences in density resulting from temperature differences; in forced convection, the motion is produced by mechanical means.

Radiation is the transfer of heat from one body to another, not in contact with it, by means of wave motion through space. “

When investigating the heat losses from the distillation columns, two types of losses from the column surface to the environment occur: losses by radiation and by convection. These two losses released to the surrounding equal the heat flux through the column walls, which offers now two approaches to calculate the total heat losses: either to determine the amount of heat transferred through the column wall (heat conduction) or to compute the convection and radiation losses.

10.4.1.1 Heat conduction

Fourier's law is the fundamental differential equation for heat transfer by conduction

$$\frac{dQ}{d\Theta} = -kA \left(\frac{dT}{dx} \right) \quad (61)$$

$\frac{dQ}{d\Theta}$	[W]	rate of heat flow (heat per unit time)
k	[W/m·°C]	thermal conductivity
A	[m ²]	area at right angles to the direction in which the heat flows
$-\frac{dT}{dx}$	[°C/m]	the rate of temperature change with the distance in direction of the heat flow



The values for the thermal conductivity as a function of temperature are listed in literature tables and may also be obtained from the material suppliers. Impurities however influence the thermal conductivity considerably (up to 50 - 75%), which is a relevant factor to be considered when calculating the heat flux.

For a cylindrical shaped body, like the distillation columns (disregarding top and bottom losses), following equation can be derived after solving the integral:

$$Q_{cylinder} = -kA \frac{dT}{dr} = -k(2\pi rL) \frac{C_1}{r} = -2\pi kLC_1 = 2\pi kL \frac{T_i - T_o}{\ln(r_o / r_i)} \quad (62)$$

Hence following parameters are required:

Dimensions of the column (length, inner and outer radius; L, r_i, r_o)

Temperature of the column surface inside the column, T_i

Temperature of the column surface outside the column, T_o

Thermal conductivity, k

Trying to apply this method at MAF for the calculation of the heat losses created the problem that a direct measurement of the inside wall temperature was not feasible with the available means, and an estimation would not have been accurate due to scale forming layers which influence not only the temperature but also the thermal conductivity.

The only option left was therefore the calculation of the losses via heat convection and radiation, which do not interfere with each other, so that both can be calculated separately and summed up afterwards.

10.4.1.2 Heat convection

Natural convection occurs when a solid surface is in contact with a fluid of different temperature from the surface and forces the fluid to move due to the imposed density differences, as the density of the "heated" air next to the hot surface is less than that of the unheated air at a distance from the body. The temperature difference between the hot surface and the air in the surrounding room causes a transfer of heat by conduction into the current of air next to the wall, and the stream carries the heat away by convection in a direction parallel to the surface.

The convection heat transfer coefficient h over a surface area A is defined by

$$\begin{aligned} \dot{q} &= h \Delta T \\ h &= \dot{q} / \Delta T \end{aligned} \quad (63)$$



- \dot{q} heat flux
 h convection heat transfer coefficient, often also α
 ΔT temperature difference

For constant wall temperature of the body T_w and a constant temperature of the surrounding fluid T_∞ , the temperature difference is calculated as follows:

$$\Delta T = T_w - T_\infty \quad (64)$$

The heat transfer is described with the dimensionless heat transfer coefficient of the Nusselt number:

$$Nu = \frac{h \cdot l}{k} \quad (65)$$

- l characteristic length (equals column height for vertical cylinders), also often D
 k thermal conductivity of the fluid, also often λ

The Nusselt number for natural convection is a function of Grashof and Prandtl number:

$$Nu = f(Gr \cdot Pr) \quad (66)$$

Grashof number

$$Gr = \frac{g l^3 \beta \rho_{fluid} \Delta T}{\nu^2} \quad (67)$$

Prandtl number

$$Pr = \frac{c_p \mu}{k} \quad (68)$$

- ν [m²/h] kinematic viscosity of the fluid
 β [1/K] coefficient of thermal expansion of fluid at constant pressure
 μ [cP] absolute viscosity of the fluid
 c_p [kJ/kg·K] specific heat of the fluid at constant pressure
 k [W/m²·K] thermal conductivity of the fluid
 ρ_{fluid} [kg/m³] density of the fluid
 g acceleration of gravity (9,81 m/s²)

$$\beta = \frac{1}{\rho} \left(\frac{d\rho}{dT} \right)_p \quad (69)$$



Developing the series of $\rho(T)$ around the reference point of T_∞ (surrounding temperature) and neglecting the terms of higher order leads to:

$$\beta\Delta T = \frac{\rho_\infty - \rho_w}{\rho_\infty} \quad (70)$$

which is valid for liquids as long as $\beta\Delta T < 1$ is fulfilled.

For ideal gases the coefficient of thermal expansion equals the reciprocal of the absolute temperature.

$$\beta = \frac{1}{T} \quad (71)$$

The properties are calculated for the mean temperature, while viscosity and Prandl number are taken at the mean wall temperature (T_m).

$$T_m = \frac{T_w + T_\infty}{2} \quad (72)$$

Combining all single expressions gives following equation:

$$\frac{hL}{k} = b \left[\frac{L^3 \rho^2 g \beta \Delta T}{\mu^2} \left(\frac{c_p \mu}{k} \right) \right]^n \quad (73)$$

b, n		constants
L	[m]	length of vertical surface, as for vertical cylinders the characteristic length is the column length itself
h	[W/m ² ·K]	heat transfer coefficient

$$h = \frac{dq / dA}{T_w - T_c} \quad (74)$$

T_w	[K]	wall temperature
$T_c \equiv T_\infty$	[K]	local average temperature of cold fluid (approximately surrounding temperature)

The product $Gr \cdot Pr$ is also known as Raleigh number

$$Ra = Gr \cdot Pr \quad (75)$$



Equation (66) can therefore be written as

$$Nu = b(Gr Pr)^n \quad (76)$$

The values for b and n for vertical cylinders are found in literature (e.g. [24]):

For $Gr \cdot Pr$ in the range of $10^4 - 10^9$: $b=0.59$ $n=0.25$

For $Gr \cdot Pr$ in the range of $>10^9$: $b=0.13$ $n=0.333$

As air is the fluid medium around the columns, all properties as well as the Prandl-Number can be found in literature tables (e.g. [21]).

10.4.1.3 Radiation

Radiation is a form of heat transfer that can be considered as energy, streaming through space at the speed of light. It can have different sources, like treatment of the surface with electron bombardment, electric discharge, or radiation of definite wavelengths, which allows heat to be transferred without mass being involved. The type of radiation discussed here is thermal radiation, which is a result of temperature and implies that higher temperatures cause higher radiation emission.

In general a body absorbs, reflects or transmits the electromagnetic waves. The fraction of radiation falling on a body that is reflected is called the reflectivity ρ , the fraction that is absorbed is the absorptivity α and the third part known as transmissivity τ . Summing up these single fractions must always equal 1:

$$\alpha + \rho + \tau = 1 \quad (77)$$

An ideal radiator is represented by blackbody radiation. Blackbody radiation is characterised by total absorption of radiation and is the standard to which all other radiators are referred. Radiation depends on geometry, emissivity of both emitting bodies and the arrangement of the bodies.

If the heat is transferred from the surface of body one to body two via radiation, the heat transfer coefficient is defined as follows:

$$h_{12} = \frac{\dot{q}_{12}}{T_{area1} - T_{area2}} \quad (78)$$



The total radiative flux (energy per time) throughout a hemisphere from a black surface of area A and absolute temperature T is given by the Stefan-Boltzmann law:

$$Q = A\sigma T^4 \quad \text{or} \quad q = \sigma T^4 \quad (79)$$

Q	[J]	heat flux
q	[W]	heat flux density or heat flow rate
σ		Stefan Boltzmann constant; $5.672 \times 10^{-8} \text{ W/m}^2\cdot\text{K}^4$
T	[K]	absolute temperature

The Stefan-Boltzmann constant represents the radiation coefficient of the black body.

The ratio of the total radiating power of a real surface to that of a black surface at the same temperature is called emittance ϵ of the surface or emissivity for a perfectly plane surface.

The total radiation per area unit of an opaque body of area A_1 , emissivity ϵ_1 and absolute temperature T_1 therefore is:

$$\frac{q}{A_1} = \sigma \epsilon_1 T_1^4 \quad (80)$$

Usually emitting surfaces however also receive radiation from other surfaces at different temperatures. The absorbed part of the incoming radiation has to be considered in the calculation of the total flux of radiant energy. For the distillation columns this means that walls, ceiling, piping, instruments, as well as the other columns are radiating towards the investigated column. Although the column loses more energy than it absorbs from the environment, the net loss by radiation is less than the result of the previous equation would be.

In general it has to be said that the heat flux transferred between two bodies is difficult to determine, unless it is not a system of two concentric spheres or cylinders, or two parallel planes in big distance to each other.

Usually the total heat loss from a hot body to its surroundings is a simultaneous heat transfer that includes losses by conduction, convection and radiation. As the two types of heat transfer occur in parallel, the total loss, assuming black surroundings, can be given as follows:

$$\frac{q_{total}}{A} = \frac{q_c}{A} + \frac{q_r}{A} = h_c(T_w - T_\infty) + \sigma \epsilon_w(T_w^4 - T_\infty^4) \quad (81)$$

The calculation of the column heat losses is based on the assumption of a grey emitting surface (the column itself) and a black surrounding (not emitting any radiation) and can



therefore be calculated by applying above mentioned equation. The reason for this simplification derives basically out of the rigorous computation effort that would be necessary to model the radiant heat flux in a more accurate way. Based on these presumptions however a kind of worst-case scenario is generated in a sense that the maximal heat loss will be determined, and the real radiant energy loss will be somewhat lower.

For the emissivity of steel [24], the values vary between 0.52 for polished steel casting, up to 0.79 for oxidized steel. As the columns at MAF are covered with dirt, a reliable factor for the emissivity is difficult to estimate and the calculations are based on a chosen emissivity of 0.79, which hides a certain degree of failure that could not be avoided with the available means.

Furthermore the heat losses from the top and bottom surfaces are disregarded and the temperature profile over the columns excluded any side and additional feed streams. The surface temperatures were taken on several spots at the top part of the column, from the lateral area as well as at the bottom part to determine an average wall temperature for each column.

10.4.2 Results column losses

heat loss	convective	radiant	total
Filter column	1 914 W	2 855 W	17 168 kJ / hr
Distillation column	3 890 W	5 777 W	34 800 kJ / hr
Rectification column	3 265 W	4 952 W	29 580 kJ / hr
Oil column	7 500 W	4 511 W	27 001 kJ / hr

The overall calculated energy loss is therefore given with 108,578 kJ/hr representing 2 % of the total steam input in the distillation sector, which is 5.616,953 kJ/hr. Comparing these results with the expected heat losses from the mass balance for the single columns (especially filter and distillation column) illustrates that this estimation of the column heat losses to the environment does not match with the previously calculated losses.

For the filter column the mass balance gives results of 7,490 mole steam that are supposed to condense to fulfil the mass balance, which equals an energy content of 295,891 kJ/hr. An equivalent big difference occurs for the distillation column with 19,985 mole condensed steam or 789,549 kJ/hr lost energy.



10.4.3 Influencing factors

- The temperature measurement at the columns could not be distributed over the whole column surface, only a few accessible spots lead to the calculation of the average column temperature. The correct temperature profile of the column, considering temperature drop at the different feed stages, could not be determined with the available means.
- For the temperature measurement of the column surface a common thermometer was used and the small contact area between thermometer ball and column surface was additionally influenced by dirt and oxide layers at the column surface.
- The taken heat transfer coefficient is only an estimated value based on literature data
- The radiative losses are based on the assumption of a grey emitting surface and a black surrounding. This model is highly simplified, but was chosen as the influence of walls, columns and piping system would have been a rigorous task, which would not have been relevant at this stage of the plant assessment.
- For the calculated reboiler duties for each column a cross-check was not possible at all. There was neither information from the steam distribution system available (no measurement devices), nor from the reboilers, as no specification sheets were found in the documentary of the company. The temporarily dismantling of the reboilers during maintenance was not supported from the management to avoid unnecessary longer standstill-times.
- Considering the overall mass balance a deviation between in- and output streams has been stated already. The mass balancing for the single columns had to be accomplished by pure calculation, as comparative measurements were not feasible with the available means. It therefore however cannot be assumed, that the balances for the single columns are more accurate than the overall mass balance, although almost exact figures might appear. For this reason a sensitivity analysis has been generated. Flowrates and concentrations have been varied in different ways, to see how sensitive the whole system is responding to changes in mass or composition of the streams.

10.4.4 Sensitivity analysis

The underlying massbalance has already been presented earlier. This massbalance is now subject to a sensitivity analysis to see how flexible the system is responding to changes in either flowrates or composition of streams.

A deviation factor from 0.01 to 0.1 (1 - 10 %) was chosen for the different scenarios, as the overall massbalance has deviation rates in this range. The investigated scenarios are:



- Varying only the reflux flowrates up to + / - 10%
- Varying only the bottom concentration up to + / - 10%
- Varying all measured flowrates (mass) up to + / - 10%
- Varying all concentrations (mass) up to + / - 10%

As especially the calculated “subcooled / losses” and “reboiler duties” shall be investigated here, the presented charts and data only display the varying outcome for these specific molar flowrates.

10.4.5 Results of sensitivity analysis

table 10: results of sensitivity analysis

		Filter	Distillation	Rectification	Oil
		subcooled & losses	subcooled & losses	reboiler duty	reboiler duty
original balance	[mole / hr]	7,490	19,985	10,704	12,982
only reflux flowrate +0.3 %	[mole / hr]	7,510	10,639	1,494	13,049
only reflux flowrate +0.29 %	[mole / hr]	7,510	10,951	1,801	13,047
all concentrations +10 %	[mole / hr]	8,420	21,002	8,068	12,076
all concentrations -10 %	[mole / hr]	6,559	18,968	13,340	13,888
only bottom concentration +10 %	[mole / hr]	2,911	3,486	- 10,374	12,982
only bottom concentration -10 %	[mole / hr]	13,086	40,150	36,465	12,982
all measured flowrates +10 %	[mole / hr]	8,239	21,984	34,668	- 2,119
all measured flowrates -10 %	[mole / hr]	6,741	17,987	- 13,260	28,084
only reflux flowrate +10%	[mole / hr]	8,180	- 291,550	- 296,280	15,204
only reflux flowrate -10%	[mole / hr]	6,799	331,520	317,688	10,761
all concentrations (mass) +10 %	[%]	- 12	- 5	25	7
all concentrations (mass) -10 %	[%]	12	5	- 25	- 7
only bottom concentration +10 %	[%]	61	83	197	0
only bottom concentration -10 %	[%]	- 75	- 101	- 241	0
all measured flowrates +10 %	[%]	- 10	- 10	- 224	116
all measured flowrates -10 %	[%]	10	10	224	- 116
only reflux flowrate +10 %	[%]	- 9	1,559	2,868	- 17
only reflux flowrate -10 %	[%]	9	- 1,559	- 2,868	17

The different scenarios are summarized in table 10: results of sensitivity analysis, showing the absolute impact on the different variations as well as the deviation expressed as percentage compared to the flowrates out of the original massbalance. The percentage deviation is calculated in following way:



$$\text{deviation} = 100 / \text{sc}_0 * (\text{sc}_x - \text{sc}_0) \quad (82)$$

sc_0 ... original subcooled molar flow

sc_x ... new subcooled molar flow

$$\text{deviation} = 100 / \text{rd}_0 * (\text{rd}_x - \text{rd}_0) \quad (83)$$

rd_0 ... original molar reboiler duty

rd_x ... new molar reboiler duty

The graphs for all four columns with the different scenarios are attached in the appendix.

10.4.6 Interpretation

Very obvious is that varying the reflux flowrates has a big influence on distillation and rectification column. These two columns react extremely sensitive within the balancing calculation: a variation of only 0.3 % on the reflux flowrates causes a 50% reduction of the calculated subcooled/losses molar flowrate. The balancing of the filter column on the other hand does not respond to deviations of flowrates or concentrations in a big extent. Only the variation of the bottom stream concentration of 10 % leads to a 50 % deviation on the internal condensed vapour flow. Similar results have been found for the oil column, where only the variation of all measured flowrates has bigger impacts on the molar reboiler duty.

Obviously not only the molar flow balance is changing, also the massbalance for the single columns shows different values for steam and drainage flowrates. As these streams could only be measured in total, it is impossible to make a valuable statement which scenario might be the correct one. Of course the balance could be modified to get a more accurate result, but for a satisfactory result and reliable crosscheck of the calculated data it will be necessary to install additional flowmeters and make frequent analysis. Especially for the steam lines to the single columns data verification through measured values would be useful, also to have the advantage of improved process controllability and independence on manual adjustments based on the operator's experience and skills.

It was therefore decided to work with the actual measured data, instead of artificial adjusted values, as due to process instabilities the assumption of perfect conditions would not be reasonable at all. The consequence for the Heat Exchanger Network development is that it has to be restricted to the overhead vapour streams, the product streams, feed molasses, drainage and boiler feed water preheating. The reboilers had to be excluded completely as this is one of the most uncertain parts within the balancing.



11 Heat Exchanger Network

11.1 Existing Heat Exchanger Network

The existing HEN was modelled and includes all the condensers but excludes the reboilers due to reasons, which have already been explained. The cooling water streams are divided in order to carry the according heat load for each condenser.

Hence hot and cold stream composition curve superimpose perfectly, apart from a slight deviation between S10 (feed molasses) and S21_3 (the distillation overhead vapour used to preheat the feed molasses), as based on the actual measured flowrates and temperatures a perfect match could not be generated. As these two streams however are supposed to reject and absorb the same amount of heat, the lower enthalpy is shown here in the HEN.

table 11: existing Heat Exchanger Network streams

#	Description	Type	Heat type	T ₁ (K)	T ₂ (K)	H (kW)	m·cp (kW/K)
1	10 (feed preheater)	Cold	Sensible	298.2	343.4	161.5	3.5723
2	40 (cond. F1)	Hot	Latent	347.0	347.0	-68.2	-
3	41 (cond. F2)	Hot	Latent	347.0	347.0	-10.6	-
4	21-3 (cond. preheater)	Hot	Latent	344.0	344.0	-154.2	-
5	24 (cond D1)	Hot	Latent	343.9	343.9	-130.4	-
6	23 (cond. O2)	Hot	Latent	345.7	345.7	-4.4	-
7	45 (cond. O1)	Hot	Latent	345.7	345.7	-109.4	-
8	46 (cond. R1)	Hot	Latent	349.2	349.2	-187.5	-
9	26 (cond. R2)	Hot	Latent	349.2	349.2	-22.9	-
10	product cooling	Hot	Sensible	343.2	308.2	-3.7	0.1049
11	cooling water S31	Cold	Sensible	304.0	310.7	130.4	19.5472
12	cooling water S29	Cold	Sensible	304.0	323.9	68.2	3.4258
13	cooling water S30	Cold	Sensible	304.0	307.9	10.6	2.7151
14	cooling water S34	Cold	Sensible	304.0	315.4	187.5	16.4187
15	cooling water S35	Cold	Sensible	304.0	305.9	22.9	11.9172
16	cooling water S36	Cold	Sensible	304.0	313.4	109.4	11.6173
17	cooling water S37	Cold	Sensible	304.0	309.9	4.4	0.7476
18	cooling water S47	Cold	Sensible	304.0	305.7	3.7	2.1952



These streams are illustrating the prevailing situation, whereas latent (condensation) and sensible (subcooling) heat have been summarized into one total energy stream to be removed, due to software limitations. Hence:

$$\text{Total DQ} \equiv \text{DQ}_{\text{latent}} + \text{DQ}_{\text{sensible}} \text{ [kW]} \quad (84)$$

The temperature is always taken as condensation temperature without considering subcooling or slight superheating, which means that a temperature range is only given for pure sensible heat streams. Additionally only one overall heat transfer coefficient U was defined without formulating a profile over the condenser.

For the product cooler, the software cannot handle 3 different streams to be cooled in one condenser. The three product streams are therefore described with one single stream, whereas 35 °C was chosen as end temperature for all streams, which leads to a theoretical total input stream temperature of:

$$T_{\text{start}} = \frac{\text{total DQ}_{\text{S14}} + \text{total DQ}_{\text{S16}} + \text{total DQ}_{\text{S17}}}{\text{CP}_{\text{S14}} + \text{CP}_{\text{S16}} + \text{CP}_{\text{S17}}} + T_{\text{end}} \quad (85)$$

$$T_{\text{start}} = 70 \text{ }^\circ\text{C} = 343.16 \text{ K}$$

In order to fulfil the overall cooling water balance, the temperature of the back-flowing cooling water stream (containing all the streams after the single condensers) has to be corrected. The measured temperature point is direct at the cooling tower, which means that along the piping system, from the distillation hall to the cooling tower, heat losses are occurring. The temperature at the cooling tower is 310.9 K. To fulfil the cooling water balance, the temperature right after the merging of all cooling water streams must be 311.9 K.

Totally excluded from the cooling water balance are those condensers that are not working at the moment (acidity condenser and condenser D2). Water is actually also flowing through these two units, but there is no basis to determine the amount of the water in there.



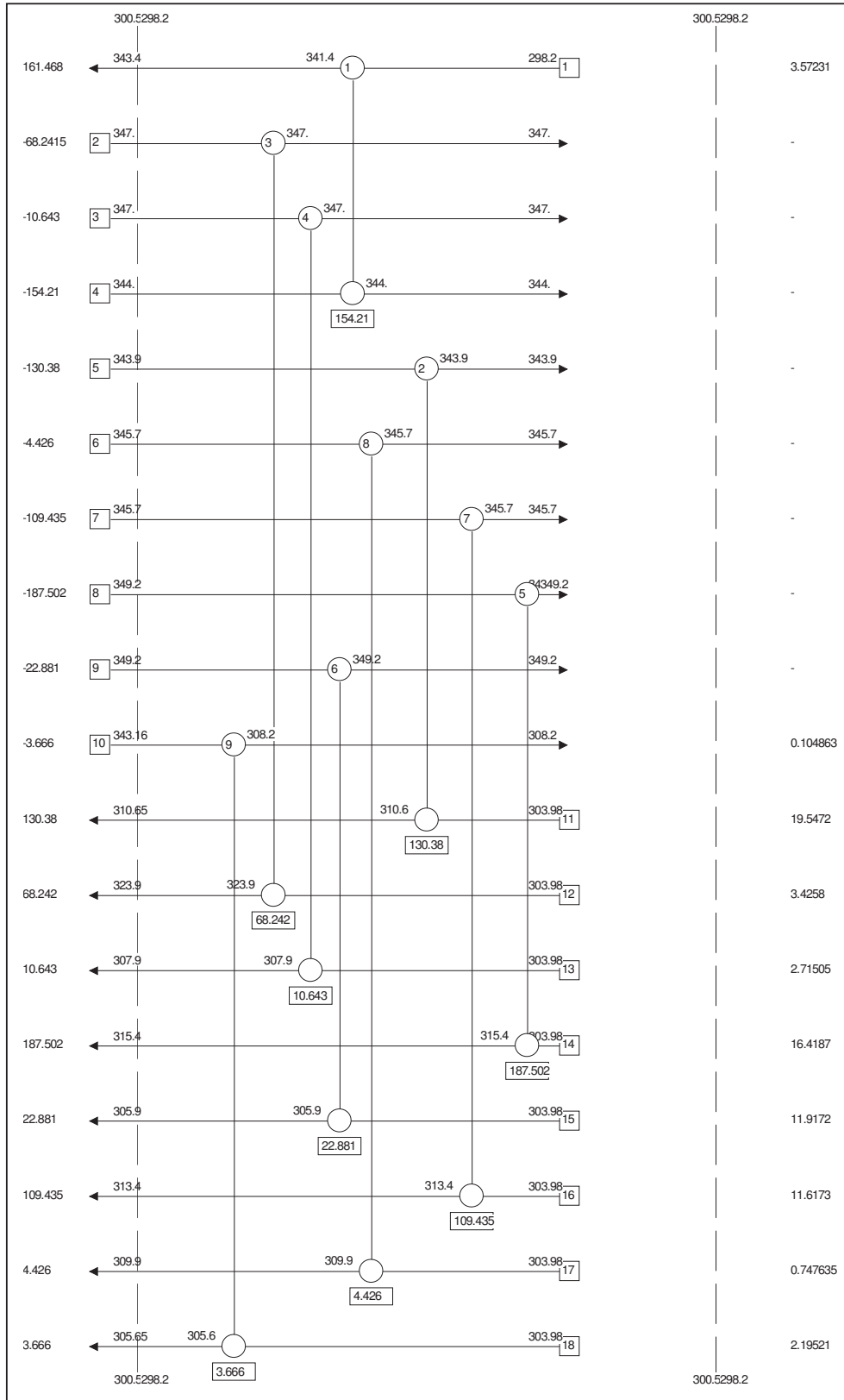


figure 36: existing Heat Exchanger Network



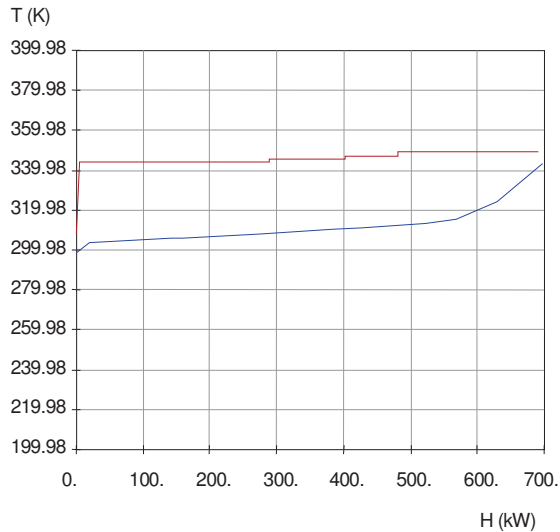


figure 37: composite curves for existing HEN

Analysing the existing Heat Exchanger Network for all condensers (figure 36) shows that hot and cold composite curves are not pinched (figure 37) but a so called threshold problem. This means that up to a certain ΔT_{\min} no Pinch occurs and heating and cooling duties therefore remain independent from the chosen ΔT_{\min} , which would not be the case for a pinched problem. These threshold problems therefore do not have a trade-off between energy and capital, as the utility usages are invariant. Hence no thermodynamic constraints are imposed by the Pinch and the problem can not be designed with the Pinch Design Method.

However B. Linnhoff [15] points out that, if there is no trade-off between energy and capital for $\Delta T_{\min} < \Delta T_{\text{threshold}}$, some threshold problems might still show a trade-off by increasing the temperature difference beyond the threshold value and create a pinched problem. He proposes to test, whether there might be a trade off or not, by plotting ΔT_{\min} against external heating and cooling duties and test the system with “experience values” of ΔT_{\min} . B. Linnhoff recommends:

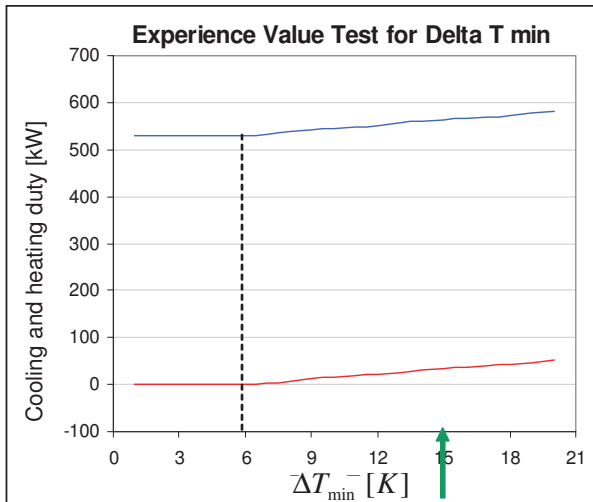
“If it lies well in the “pinched” region then there will almost certainly be an energy/capital trade-off, to be explored as already described. A Maximum Energy Recovery design will be produced by the Pinch Design Method. If the ΔT_{\min} experience value lies near the threshold value, then there is likely to be a trade-off and the problem should be treated as pinched problem. If the experience value falls well into the insensitive region, then there is no trade-off and the problem cannot be designed by the Pinch Design Method.”



11.2 Retrofit Heat Exchanger Network

As the previous presented network already included the cold utility, the same network was generated again without cooling water, to determine the threshold temperature difference. The different scenarios revealed that up to a ΔT_{min} of 5.8 °C no external heating duties are required and 529.9 kW have to be carried away by cooling water. Further increasing of ΔT_{min} would create additional demand for external heating and consequence in an additional heat exchanger.

For this Heat Exchanger Network however the “experience ΔT_{min} ” test shows (figure 38), that it is likely that a capital/energy trade-off exists and it should be tried to solve the problem with the Pinch Analysis.



The dotted line indicates $\Delta T_{threshold}$ and the green arrow $\Delta T_{min, experience}$

The existing network however does not include all potential streams within the distillation sector. So far the boiler feed water, that has to be preheated and the hot drainage stream are not implemented at all and should be considered when proposing a solution. Table 12 summarizes the streams to be considered for the new Heat Exchanger Network.

figure 38: experience value test 1

table 12: streams for retrofit HEN

#	Description	Type	Heat type	T ₁ (K)	T ₂ (K)	H (kW)	m·cp (kW/K)
1-10	see table 11						
11	boiler feed water preheat	Cold	Sensible	298.2	342.2	112.4	2.5538
12	drainage stream total	Hot	Sensible	363.2	298.2	-463.5	7.1313

The cooling water is not yet implemented as this will be considered as utility stream and not as process stream. The composite curves (figure 39) indicate again a threshold problem and not a pinched problem:



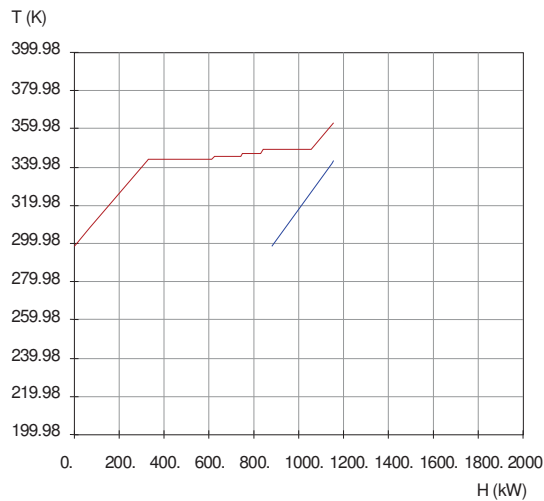


figure 39: composite curves for retrofit HEN

Making the “experience ΔT_{\min} ” test for this problem (figure 40), the experience based ΔT_{\min} lies well in the region, where the Pinch Method cannot be applied.

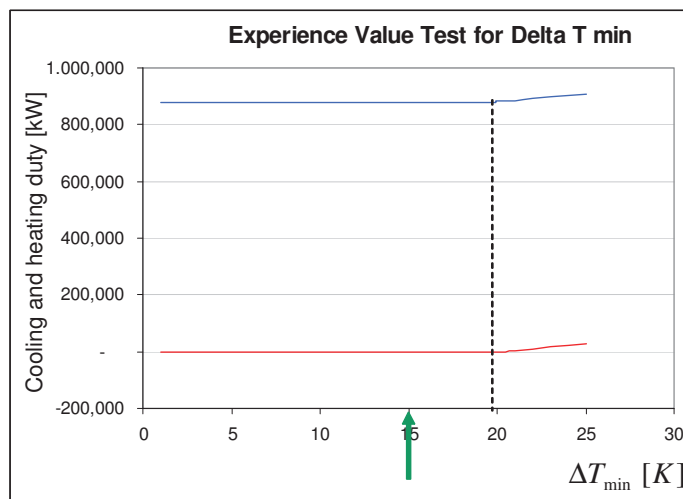


figure 40: experience value test 2

Obviously a similar graph results from a scenario where the overhead vapours are only taken as one stream, without splitting into condenser 1 and condenser 2 streams, as the total energy to be transferred remains the same. ΔT_{resh} would, like in the previous scenario, again be 19.9 °C, hence the Pinch Method can not be applied.

B. Linnhoff suggests for these kind of problems:

“...usually a great many designs are possible. The design will generally be determined by placing heaters or coolers for good control, applying the ticking-off-rule, and by identifying essential matches at the “non-utility” end. ... In placing matches, the smallest value of ΔT is kept as large as possible, but allowed to reach some practical minimum if necessary. This value of course has nothing to do with the trade-off between energy and capital.”

Based on these guidelines and the restriction from the management to keep additional investment as low as possible, it was decided to influence the existing network in a way that no big modifications should be done until the process optimization project itself is implemented. Still the two new streams should be integrated, to lower energy losses and external utility usage.

11.2.1 Scenarios

As the Pinch Analysis could not be applied anymore, the problem was reduced to following scenarios:

1. Everything remains as it is, only one additional heat exchanger should be implemented to preheat the boiler feed water by using the drainage stream.
2. The drainage stream is used to preheat feed molasses and boiler feed water.
3. The drainage stream is only used to preheat the molasses stream, and the distillation column overhead vapour is used for the boiler feed water preheating.
4. The drainage stream is only used to preheat the molasses stream while the distillation column overhead vapour is cooled by cooling water.

Another option be to take the distillation column overhead stream as one stream, without the distinction of preheater and second condenser, as a partial condensation is not necessary as such. This would mean that one bigger condenser would replace the two smaller ones. The two others might have found usage at another place in the network. However the constraint that the feed molasses is already at the upper temperature and capacity limit (the massflowrate is determined by the pump capacity) does not allow this option. Further preheating of the molasses is not desired, as the molasses temperature is already close to the ethanol boiling point and the impact of partial vaporization has to be modelled first in terms of process optimization before the energetic impact can be analysed.



table 13: scenario overview

scenario	general advantage	general disadvantage	Add ext. duties	# add. HE	# add. pumps
1	No intervention in existing network		-	1	1
2			CW ¹⁾ for S21_3	2	1 ²⁾
3		Unwieldy additional piping system, from distillation hall to boiler house	maybe extra CW	1(2) ³⁾	2
4		The boiler feed water remains without preheating	CW* for S21_3	1	1

The advantage of using the drainage stream for the boiler feed water preheating is, that the water can be brought to higher temperatures, as it would be possible with the vapour stream. This would not only lead to steam savings, but also to direct oil savings, as the heat to bring the water to the boiling point does not have to be covered by fuel oil as it is the case now, but can be provided by the molasses stream.

$$m_{H_2O} \cdot c_{p,H_2O} \cdot \Delta T = m_{oil} \cdot \text{calorific value} \quad (86)$$

$$\Delta T = T_{boiling} - T_{feed}$$

The remaining streams are therefore reduced and treated as the system to be investigated (table 14: considered streams for retrofit HEN). Drawing another composite curve diagram confirms that it is still not pinched (figure 41).

table 14: considered streams for retrofit HEN

#	Description	Type	Heat type	T ₁ (K)	T ₂ (K)	H (kW)	m·cp (kW/K)
1	10 (feed preheater)	Cold	Sensible	298.2	343.4	161.5	3.5723
2	21-3 (cond. preheater)	Hot	Latent	344.0	344.0	-154.2	-
3	boiler feed water preheat	Cold	Sensible	298.2	342.2	112.4	2.5538
4	drainage stream total	Hot	Sensible	363.2	298.2	-463.5	7.1314

¹ CW...cooling water

² The molasses stream has to be splitted, but one pump can be used for the total system (like cooling water system)

³ if the vapour stream cannot be completely condensed in the boiler feed water heat exchanger, an additional cooling water heat exchanger has to be installed



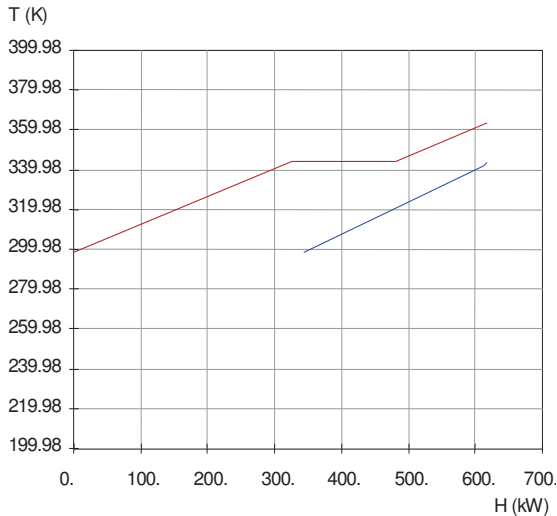


figure 41: composite curves for remaining retrofit streams

11.2.2 Results

Scenario 1: The drainage stream will cool down to 374.4 K after heating up the boiler feed water (112.4 kW). A minimum temperature difference of 21 °C can be maintained in the heat exchanger for this scenario.

Scenario 2: The drainage stream has to be splitted 50:50. This measure allows keeping a temperature driving force of 19 °C for both exchangers. 154.21 kW have to be taken away by cooling water from stream 21_3.

Scenario 3: The drainage stream leaves the system with 67.6 °C after heating up the feed molasses to 70 °C. The overhead vapour will not be fully condensed after heating up the boiler feed water to 69 °C, and therefore needs additional cooling water, that carries away the remaining 41.8 kW – this requires the implementation of an additional heat exchanger.

Scenario 4: Same situation as in scenario 3 for drainage and molasses stream. An additional heat exchanger operated with cooling water to condense the complete stream 21_3 (154.21 kW) has to be installed and steam to preheat the boiler feed water (112.4 kW) is still necessary. Assuming that the boiler feed water temperature is not raised, the results presented in table 15 emerge for the different scenarios¹⁾:

¹ NOTE: the prices are actual prices for MAF in Birr (local currency) converted to EUR with an exchange rate of 10.5; 1 m³ water = 3 Birr, 1 l fuel oil light = 2.2 Birr; 245 working days; 24 operating hours



table 15: results for different retrofit scenarios 1

Scenario	Steam (water) savings / yr	Oil savings / yr	Add CW	Heat Exchanger	Pumps
	[EUR]	[EUR]	[kW]	[units]	[units]
1	294.00	16,256.00	-	1	1
2	294.00	16,256.00	154.21	2	1
3	294.00	16,256.00	41.80	2	2
4	-	-	154.21	1	1

Assuming that the molasses stream could preheat the water up to 82 °C (which equals a ΔT_{\min} of 10 °C) the different scenarios¹⁾ can be compared as follows (summary in table 16):

Scenario 1: 145.4 kW are transferred from the drainage stream to the boiler feed water. The drainage stream leaves the system with 69.5 °C.

Scenario 2: The molasses stream again has to be splitted, but still carries enough energy to provide both cold streams. The extra cooling water demand remains of course the same with 154.21 kW.

Scenario 4: The drainage stream has an end temperature of 67.5 °C, cooling water and steam demand remain the same.

table 16: results for different retrofit scenarios 2

Scenario	Steam (water) savings / yr	Oil savings / yr	Add CW	Heat Exchanger	Pumps
	[EUR]	[EUR]	[kW]	[units]	[units]
1	294.00	19,701.00	-	1	1
2	294.00	19,701.00	154.21	2	1
4	-	-	154.21	1	1

¹ NOTE: Scenario 3 had to be excluded as distillation column overhead vapour has only $T = 71$ °C and can therefore not be used to preheat the boiler feed water up to 82 °C.



One more option that might have relevance for the future is that more boiler feed water might have to be preheated, once the new boiler will be installed and the second line will operate as well. Assuming the new boiler will operate at maximum capacity and produce 6000 kg steam per hour, the required energy to preheat 6000 kg water from 25 °C to 82 °C is 397 kW, which can be provided by the drainage stream (total 463 kW). The drainage stream will be cooled down to 34 °C and a ΔT_{\min} of 9 °C is still available. So far however it remains unknown how much additional steam will really be required for the second line, and an analysis can only be done once this information is available.

11.2.3 Cost estimation

It is very clear that scenario one offers the highest savings, without additional cooling water demand, requiring only one additional heat exchanger, one pipeline and one pump for transferring the drainage stream from the distillation hall to the boiler house.

For cost estimations of plants a pre-calculation according Chilton [30] can be used.

1. Heat exchanger * factor 1.15 = installed heat exchanger

2. From the installed heat exchanger price, additional surcharge for

	Recommended range	Chosen factor
2.1 Piping for liquid media	30-60 %	45 %
2.2 Instrumentation (medium automation)	5-10 %	7.5 %
2.3 Housing (already existing)	5-20 %	5 %
2.4 Supply and service equipment	0 %	0 %
2.5 Connection piping	0-5 %	2.5 %

3. Subtotal

From the subtotal additional surcharge for

	Recommended range	Chosen factor
3.1 Engineering and construction	20-35 %	0 % ¹⁾
3.2 Unexpected difficulties	10-20 %	0 % ¹⁾
3.3 Size factor of the plant (small)	5-15 %	0 % ¹⁾

¹ As the additional equipment to purchase and install is limited to pumps, heat exchanger and piping system, it does not seem reasonable to consider these three factors.



4. Based on the sum a final correction with the Chemical Engineering Plant Cost Index or Process Engineering Plant Cost Index has to be considered, unless not an actual price offer for the basic equipment is available.

As an index adaptation for Ethiopia was not available, the price calculation has to be based on recent offers for heat exchangers and pumps that are specifically addressed to the National Alcohol and Liquor factory.

To know which heat exchanger size is necessary to fulfil the desired heat exchange between drainage stream and boiler feed water, a heat transfer coefficient has to be determined in order to get the required HE area and to choose the appropriate heat exchanger.

Table 17 gives an overview about the actual calculated heat transfer coefficients U for the existing heat exchangers and corresponding recommended values from literature sources. It can be seen that the U values are varying a lot at times, which indicates fouling influence in existing heat exchangers and different assumptions for fouling considerations.¹⁾

table 17: heat transfer coefficients

condenser Nr.	surface (outer)	corres- ponding U	theoretical U from literature [15]	theoretical Q	Efficiency	theoretical U from literature [21]	theoretical Q	Efficiency
	[m ²]	[W/m ² K]	[W/m ² K]	[W]	[%]	[W/m ² K]	[W]	[%]
Preheater	22.65	696	524	118,918	132.80	675	153,187	103.10
D Cond 1	17.76	277	764	359,013	36.32	300	140,974	92.49
F Cond 1	11.26	188	764	276,996	24.64	300	108,768	62.74
F Cond 2	2.51	113	764	71,691	14.85	300	28,151	37.81
R Cond 1	17.76	269	764	531,589	35.27	300	208,739	89.83
R Cond 2	2.51	382	764	45,777	49.98	300	17,975	127.29
FO Cond 1	11.26	264	764	316,320	34.60	300	124,210	88.11
FO Cond 2	2.51	49	714	64,141	6.90	300	26,950	16.42

Different assumptions for the overall heat transfer coefficients lead also to a very diverse area demand for the additional heat exchanger in both scenarios:

¹ The formation of deposit on heating surfaces is known as fouling and causes an additional fouling-resistance in the heat transfer process. In general the fouling resistance is an unknown parameter, but can be estimated for similar thermal and hydrodynamic conditions.



- 1A boiler feed water is heated up to 69 °C
 1B boiler feed water is heated up to 82 °C

How varying the area requirements are, is shown in following overview (table 18):

table 18: area demand for different scenarios and U estimations:

		scenario	
		1A	1B
Q	[kW]	112,64	145,44
T_{lm}	[K]	33,1	21,3
U (B. Linnhoff) for low viscosity organic fluid vs. treated cooling water [15]	[W/m ² K]	720	720
A	[m ²]	4,72	9,46
U (Perry's) for organic sol. and water [24]	[W/m ² K]	283,92	283,92
lower limit of range (283.9 – 851.7 W/m ² K) including dirt factor 0.0005283 [m ² / s K J]			
A	[m ²]	11,98	24,00
VDI Wärmeatlas liquid-liquid [21]	[W/m ² K]	150	150
150-1200 [W / m ² K]			
lower limit for non-ideal conditions			
A	[m ²]	22,67	45,42

As the lower limit of the recommended VDI [21] value might be somehow too conservative and the Linnhoff [15] value a bit too optimistic, it was decided to calculate the heat exchanger area with the lower recommendation of U from Perry's Chemical Engineers' Handbook [24].

For these area demands, following recent price offer build the basis for further calculations:

	Type I	Type II
Shell diameter	600 mm	400 mm
Shell length	2400 mm	2500 mm
# tubes	103	52
Outer tube diameter	35 mm	35 mm
Tube length	2000 mm	2000 mm
HE area	22.65 m ²	11.44 m ²
Price	13,250.00 EUR	7,850.00 EUR



11.2.4 Return on Investment and Payback ratio [31]

The payback period is the number of years from plant start-up required to recover all expenses involved in a project, if all the profits were used for this purpose. A payback period of less than five years is usually required for a project to proceed (preferably closer to 3 years).

$$\text{Payback} = \text{investment} / \text{average annual savings or profits} \quad (87)$$

The Return On Investment (ROI) is the expected profit divided by the total capital invested, expressed as a percentage return. The ROI is compared with the earnings that could be achieved by an alternative investment, e.g. capital bonds. An ROI of at least 15-20% is usually expected assuming that the project is not particularly risky.

$$\text{ROI} = (\text{cost saving} - \text{calc. interest rate}) / \text{investment} \quad (88)$$

11.2.5 Result

table 19: result – retrofit scenarios

	scenario 1A	scenario 1B
	actual offer	actual offer
	[EUR]	[EUR]
heat exchanger	7,850	13,250
instalment	9,028	15,238
subtotal	9,028	15,238
pipng	4,062	6,857
instrumentation	677	1,143
housing	451	762
connection piping	226	381
subtotal [EUR]	14,444	24,380
Pump [EUR] (5 m ³)	13,125	13,125
total [EUR]	27,569	37,505
savings [EUR]	16,550	19,995
payback time [years]	1.67	1.88
ROI [%]	59.99	53.29

Analysing this result in terms of fuel oil savings per litre ethanol, the initial 1.11 l fuel oil per litre produced ethanol (valid for the investigation period) will be reduced to 1.01 l oil/l ethanol for scenario 1B or 1.02 l oil/l ethanol for scenario 1A.



11.2.5.1 Uncertainties

As the plant is located in Addis Ababa it is indispensable for the company to get an actual offer that includes the whole range of surcharges, as the here calculated surcharges are based on western standards and do not display Ethiopian pricing levels.

11.2.6 Interpretation

Both scenarios show very good payback times and ROI values. Scenario 1A seems to be slightly better, but when taking the annual savings into account it, scenario 1B will cause higher savings of 3,445 EUR per year compared to scenario 1A, once the break even point is reached. An additional factor is that the results here are based on the assumption of constant oil prices. If the oil prices increase in the coming years, the relative savings will be higher compared to the calculated savings here. A final decision on which alternative is more preferable can only be made by the management of MAF itself, based on internal cost accounting. The only clear recommendation that can be made here is, that one of the two heat exchangers definitely should be installed as the investment is justified in energetic and economical terms.

The final comparison for the actual reduction of fuel oil input per litre produced ethanol shows that the achievements in terms of fuel savings are about 10 %, but still minor when comparing it with competitors that claim to have results in the range of 0.33 litre fuel per litre ethanol instead of about 1 litre fuel per litre ethanol as it is the case at MAF.

Further energy savings might come up in the course of the column process optimization project, as feed location, reflux ratio and bottom and top stream specification have significant impact on the reboiler (steam) and condenser (cooling water) duty.



12 Results and discussion

The result shows that under current conditions for two different scenarios

- 1A boiler feed water is heated up by the molasses drainage stream to 69 °C
 1B boiler feed water is heated up by the molasses drainage stream to 82 °C

following investment leads to the presented annual savings:

scenario	1A	1B
total investment [EUR]	27,569	37,505
Annual savings [EUR]	16,550	19,995

This equals payback times of less than 2 years and is therefore an attractive solution for fuel oil savings.

scenario	1A	1B
payback time [years]	1.67	1.88
ROI [%]	59.99	53.29

However the desired reduction of fuel oil input from 1.11 l fuel oil per liter produced ethanol to about 0.33 could by far not be achieved. Implementing the proposed heat exchanger will lead to a 10 % reduction of the index number, but it seems higher savings have to be achieved by optimizing the distillation process first, under parallel consideration of reboiler and condenser duties. Then it will be necessary to establish a reliable data collection system for a profound base to generate mass- and energy balances. Once these balances are in an acceptable accuracy range, another investigation of the Heat Exchanger Network might reveal some more energy recovery possibilities. This implies also requirements in terms of stable process control as well as permanent installed measurement devices and proper maintenance procedures. With the investment proposed here, annual savings between 16,550 and 19,995 EUR can be achieved, but good household keeping and long-term competitiveness will definitely demand further process and procedure improvements.



13 Summary

The whole work was very much influenced by unstable process conditions and lack of reliable data, due to non-available specifications, measurement devices and manual process control. All this had major impact on the procedures and progress during process description, data collection and verification. The data collection process itself was a time-consuming work that finally did not pay off very well, as high deviation rates had to be accepted. It would have been advantageous to have long-term databases that show also seasonal changes. As this necessity however was not yet available, the gained data had to be taken with all the consequences in later balancing steps.

To achieve reliable mass- and energy balances it was tried to compare the measured data with theoretical simulations of the single columns. As the plate efficiency of the columns could not be determined however, this remained a non-useful approach. The only conclusion that could be made out of the modelling part is that the feed stage locations of some streams are quite misplaced. It is crucial to initiate a proper column investigation and optimization project, not only to gain reliable data, also to improve the distillation process itself.

Hence as the mass balances could not be verified, the energy balances are hiding the same level of uncertainties. Again it was tried to calculate the column heat losses in two different ways: once out of the molar flow balance, that determines how many moles of vapour are supposed to be condensed or evaporated in order to fulfil the mass balance, and once via calculation of convective and radiative energy losses from the columns. The results were not matching but could have been modified easily by considering the possible deviation ranges. Nevertheless an artificial whitewashing was deliberately avoided, as it was an important aspect of the whole work also to point out where the weak points are, and to stress how important it is to invest in proper process control devices.

The developed Heat Exchanger Network (HEN) is therefore based on the measured results, in complete awareness of the accuracy degree of the balances. When starting to prepare literature and get familiar with the Pinch Analysis software, it was not yet clear that the problem can not be solved with the proposed methodology, as the streams that were supposed to be implemented in the retrofit HEN did not form a pinched problem anymore. This situation led to four remaining scenarios that were feasible without big interference in the existing HEN. Out of these possibilities a saving and cost analysis, considering steam (fuel oil) savings, additional cooling water demand and required equipment (heat exchangers, pumps, piping and instrumentation) singled out one solution for two different ways of application. The drained hot molasses stream should heat up the boiler feed water, where either the present temperature level can be kept or more energy can be extracted from the drainage stream that will reduce the actual fuel oil consumption additional.

Both solutions however do not lead to the desired reduction of fuel oil input from 1.11 l fuel oil per litre produced ethanol to about 0.33. Only a 10 % reduction from 1.11 to 1.00 l fuel oil per



litre produced ethanol could be achieved. A final economical analysis is comparing the two scenarios as follows:

scenario	1A	1B
total investment [EUR]	27,569	37,505
Annual savings [EUR]	16,550	19,995
payback time [years]	1.67	1.88
ROI [%]	59.99	53.29

It is obvious that the investment in one of the recommended solutions is paying off very well and should be implemented for reasons of cleaner production as well as cost savings. What has to be kept in mind however is that only good household keeping with proper maintenance procedures and better control and measurement devices will lead to long term competitiveness.



14 Indexes

14.1 Literature

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14.2 Abbreviations

\dot{m}	Mass flowrate
°C	Degree Celsius
a	Year
A	area
AAU	Addis Ababa University
AI	Artificial Inteligence
CP	the mass specific heat of the streams
DAP	Diammoniumphospat
dGH	Deutsche Grad Härte (German Degree Hardness)
etc.	et cetera
FG	Flue gas
Gr	Grashof Number
HE	Heat Exchanger
HEN	Heat Exchanger Network
IEA	International Energy Agency
K	Kelvin
L	Liter
MAF	Mekanissa Alcohol Factory
MER	Minimum Energy Requirement
MUL	Montanuniversität Leoben
Nu	Nusselt Number
OD	optical density
p	pressure
PFD	Process flow diagram
pH	Degree of acidity
PID	Piping and instrumentation diagram
Pr	Prandl Number
Ra	Raleigh Number
RC-losses	Radiation and Convection Losses
ROI	Return On Investment
T	Temperature
UA analysis	Heat transfer coefficient – area - analysis
UNEP	United Nations Environment Programme
UNIDO	United Nations Industrial Development Organization
VLE	Vapour Liquid Equilibrium



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15 Appendix

15.1 Column description

Column	Material	Dimension	Sections	Accessory parts
Distillation column	Stainless steel; upper four sections: copper	Ø 90.7 cm H: 1010.5 cm	12 sections stainless steel, add. the 13 th and 14 th section (on top) are made of copper. 1x11 bubble cup plates Two additional copper sections on top (Ø ~ 40 cm) with each four bubble cup plates the copper sections serve as acidity segment	Preheater Cond. Type 1 D Cond. 1 Type 4 D Cond. 2 Type 5 Acidity Cond. Type 2
Filter (or deputation) column	Stainless steel	Ø 80 cm H: 636cm	five sections : 1. 0 bubble cup plates 2. 6 bubble cup plates 3. 6 bubble cup plates 4. 5 bubble cup plates 5. 6 bubble cup plates sect. 1 = bottom section sect. 5 = top section ∑ 23 bubble cup plates	F Cond. 1 Type 3 F Cond. 2 Type 5
Rectification column	Stainless steel	Ø 80.9 cm H: 1214cm	8 sections : 1. 0 bubble cup plates 2. 7 bubble cup plates 3. 6 bubble cup plates 4. 9 bubble cup plates 5. 9 bubble cup plates 6. 9 bubble cup plates 7. 9 bubble cup plates 8. 10 bubble cup plates sect. 1 = bottom section sect. 8 = top section ∑ 59 bubble cup plates	R Cond. 1 Type 4 R Cond. 2 Type 5 R Reboiler No data available
Fusel oil column	Stainless steel	Ø 55.7 cm H: 1002 cm	10 sections : 1. 0 bubble cup pl. 2-10 6 bubble cup pl. sect. 1 = bottom section sect. 10 = top section ∑ 54 bubble cup plates	FO Cond. 1 Type 3 FO Cond. 2 Type 5 FO Reboiler No data available
Demethylizing column	copper	-	10 sections : 1. 0 bubble cup pl. 2-10 5 bubble cup pl. sect. 1 = bottom section sect. 10 = top section ∑ 45 bubble cup plates	-



15.2 Condenser description

All condensers are vertical one pass shell and single tube heat exchangers with fixed tubeplates. The cooling water in the tubes flows countercurrent to the condensing vapour at the shell side.

Type 1

Tube # : 103
 Tube: ($\varnothing_{\text{outer}} \times L_{\text{tube}}$) 35 x 2000 mm
 Tube thickness: 1.5 mm
 Shell: ($\varnothing_{\text{inner}} \times L$) 600 x 2400 mm
 Channel length (L_{ch}): 200 + 200 mm

Type 2

Tube # : 13
 Tube: ($\varnothing_{\text{outer}} \times L_{\text{tube}}$) 40 x 1420 mm
 Tube thickness: 3 mm
 Shell: ($\varnothing_{\text{inner}} \times L$) 195 x 1900 mm
 Channel length (L_{ch}): 200 + 280 mm

Type 3

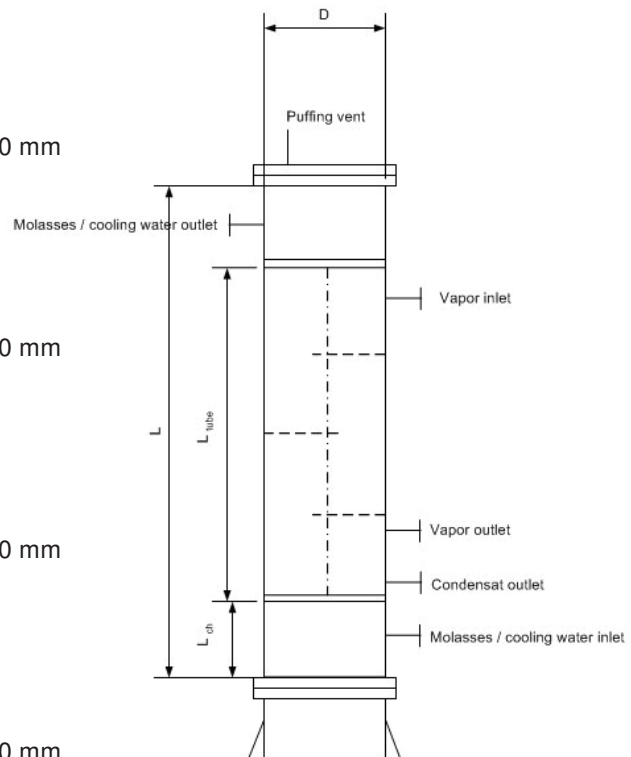
Tube # : 52
 Tube: ($\varnothing_{\text{outer}} \times L_{\text{tube}}$) 35 x 1970 mm
 Tube thickness: 1.5 mm
 Shell: ($\varnothing_{\text{inner}} \times L$) 400 x 2400 mm
 Channel length (L_{ch}): 215 + 215 mm

Type 4

Tube # : 103
 Tube: ($\varnothing_{\text{outer}} \times L_{\text{tube}}$) 28 x 1960 mm
 Tube thickness: 1.5 mm
 Shell: ($\varnothing_{\text{inner}} \times L$) 600 x 2400 mm
 Channel length (L_{ch}): 220 + 220 mm

Type 5

Tube # : 19
 Tube: ($\varnothing_{\text{outer}} \times L_{\text{tube}}$) 28 x 1500 mm
 Tube thickness: 1.5 mm
 Shell: ($\varnothing_{\text{inner}} \times L$) 250 x 1930 mm
 Channel length (L_{ch}): 215 + 215 mm



15.3 Specification of Fuel Oil Light (inland code: 98 181 FC)

Test	Specification	Test method
Specific Gravity @ 60 Deg. F	0.890 – 0.950	ASTM D 1298
Flash Point (P.M.) Deg. F	min 150	ASTM D 93
Explosiveness %weight	max 50	FS 1151
Sulphur %weight	max 0.3	ASTM D 1552
Calorific Value, Calc BTU / LB	min 18500	calculated
Viscosity Kinematic @ 122 D.F., CST	45 – 80	ASTM D 445
Viscosity Redwood No. 1 @ 100 D.F. Sec.	300 -600	10 - 70
Pour Point Deg. F	max 55	ASTM D 97
Carbon Residue %weight	max 10	ASTM D 189
Asphaltene %weight	max 3	IP - 143
Ash %weight	max 0.05	ASTM D 482
Sediment by Extraction %weight	max 0.1	ASTM D 473
water content %Vol	max 0.5	ASTM D 95
water & sediment %Vol	max 0.5	ASTM D 1796
Thermal stability NBTL Heater	max No.1	FS 3461

15.4 Vestas Aircoil Cooling Tower nominal specifications

Cooling capacity	1257.0 kW
Wet bulb temp.	17.0 °C
Water inlet temp.	40.0 °C
Water outlet temp.	20.0 °C
Water flow	54.0 m ³ /hr
Water consumption	1.6 m ³ /hr



15.5 Column heat losses – calculation tables

T	ρ	c_p	β	λ	η	ν	Pr
°C	kg/m ³	kJ/kg K	10 ⁻³ K ⁻¹	10 ⁻³ Wm ⁻¹ K ⁻¹	10 ⁻⁶ Pas	10 ⁻⁷ m ² s ⁻¹	-
0	1.2750	1.006	3.674	24.18	17.24	135.20	0.7179
10	1.2300	1.007	3.543	24.94	17.74	144.20	0.7163
20	1.1880	1.007	3.421	25.69	18.24	153.50	0.7148
30	1.1490	1.007	3.307	26.43	18.72	163.00	0.7134
40	1.1120	1.007	3.200	27.16	19.20	172.60	0.7122
60	1.0450	1.009	3.007	28.60	20.14	192.70	0.7100
80	0.9859	1.010	2.836	30.01	21.05	213.50	0.7083
100	0.9329	1.012	2.683	31.39	21.94	235.10	0.7070

Column	surface area	T_{average}	$T_{\text{surrounding}}$	emissivity	Boltzmann constant	heat loss radiation
	[m ²]	[K]	[K]	-	[W/m ² · K ⁴]	[W]
filter	15.98	326.55	293.15	0.79	5.67E-08	2854.83
distillation	28.79	330.03	293.15	0.79	5.67E-08	5777.06
rectification	30.85	323.61	293.15	0.79	5.67E-08	4951.76
oil	17.53	338.48	293.15	0.79	5.67E-08	4510.77

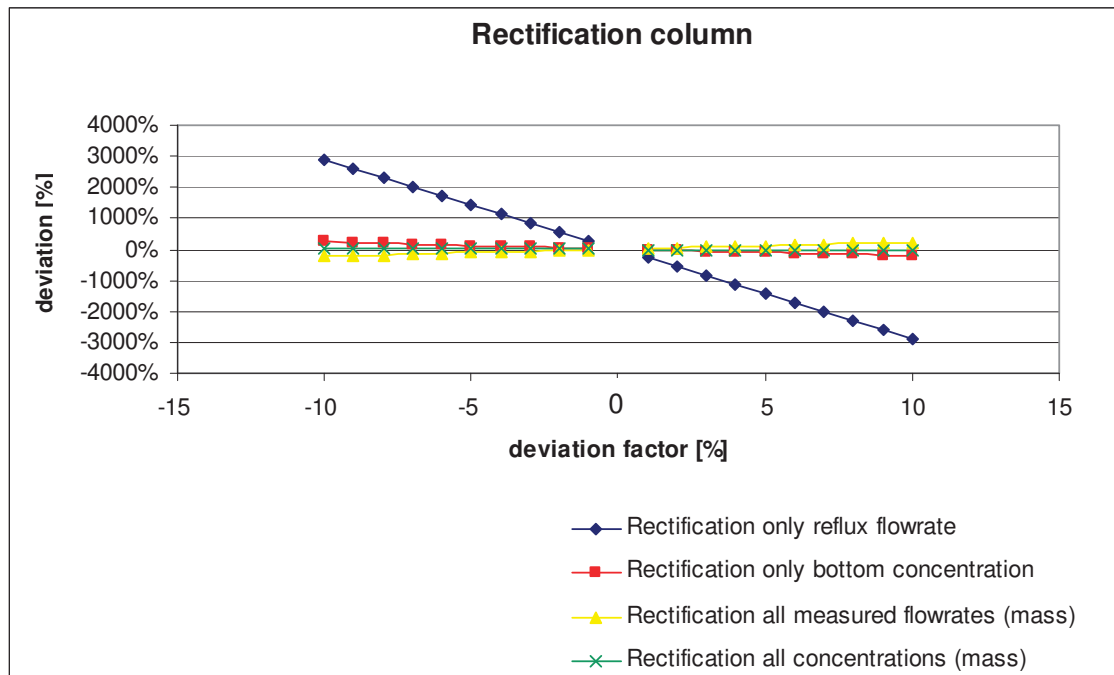
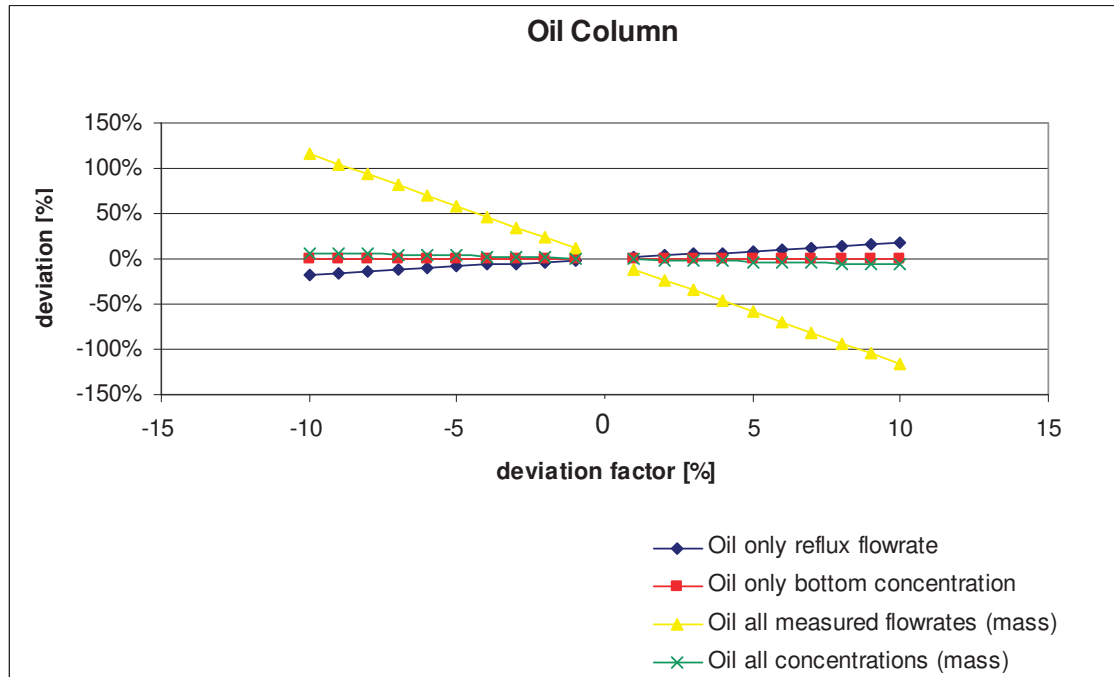
Column	height	$T_{\text{wall, average}}$	T_{mean}	gravitational acceleration	density air	density air	density air
	[mm]	[°C]	[°C]	[m/s ²]	[kg/m ³]	[kg/m ³]	[kg/m ³]
for T					$T_{\text{surrounding}}$ (20 °C)	T_{wall}	T_{mean}
filter	6,360	53.40	36.70	9.81	0.90	0.81	0.85
distillation	10,105	56.88	38.44	9.81	0.90	0.80	0.85
rectification	12,140	50.46	35.23	9.81	0.90	0.82	0.86
oil	10,020	65.33	42.67	9.81	0.90	0.78	0.84

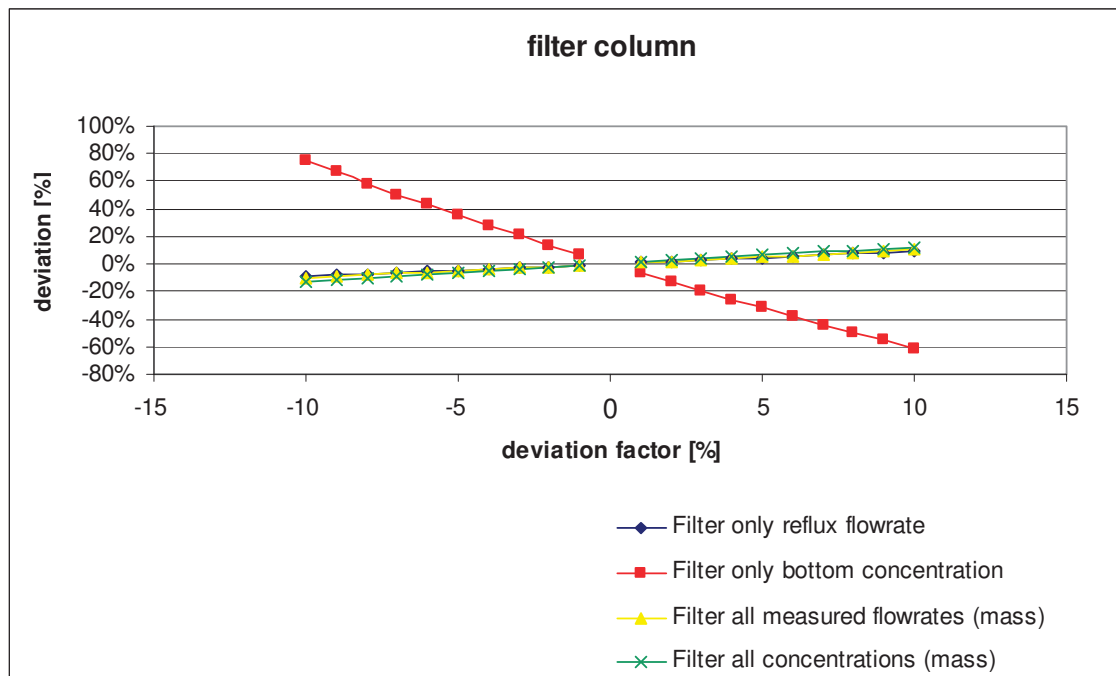
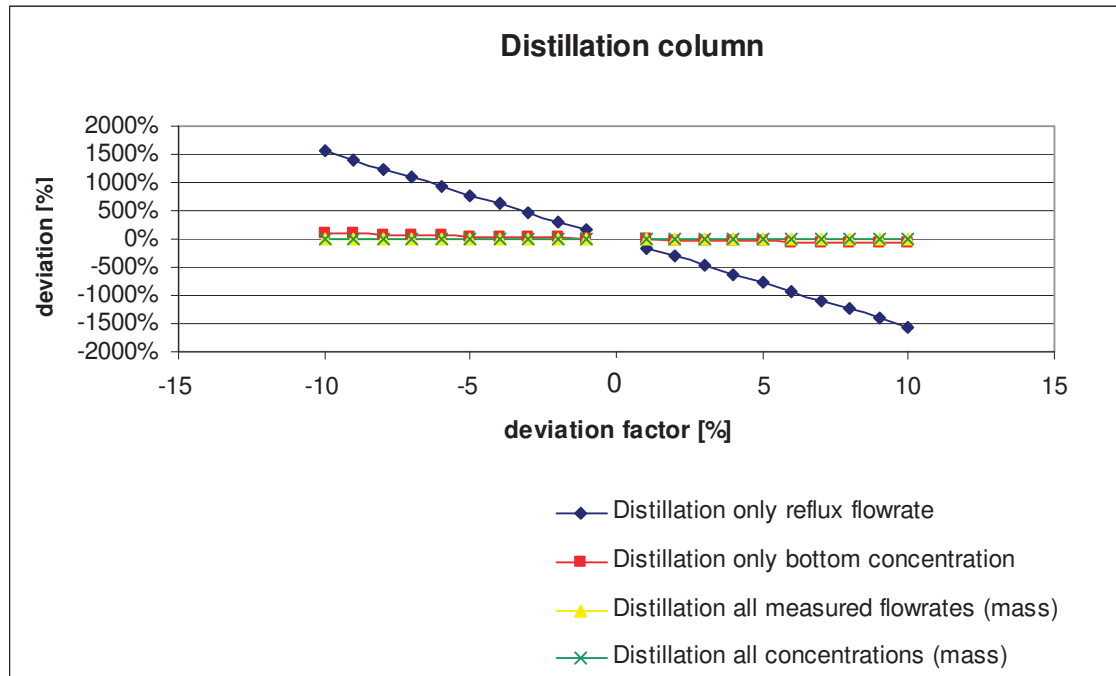
Column	beta* delta T	kinematic viscosity, air ny	absolute viscosity, air eta	specific heat capacity, air c _p	thermal conductivity, air λ or k	thermal conductivity, air λ or k
		[m ² /s]	[Pas]	[kJ/kg K]	[W/m °C]	[W/m °C]
for T	T_{mean}	T_{wall}	T_{wall}	T_{wall}	T_{wall}	T_{mean}
filter	0.11	0.000018607	0.00001983	1.00834	0.028125	0.026919
distillation	0.12	0.000018956	0.00001999	1.00869	0.028375	0.027046
rectification	0.10	0.000018311	0.00001969	1.00805	0.027913	0.026812
oil	0.14	0.000019825	0.00002038	1.00927	0.028976	0.026625

Column	Gr	Pr	Gr*Pr	b	n	Conv. heat transfer coeff.	heat loss convection	total heat loss
						[W/m ² °C]	[W]	[W]
for T		T_{mean}						
filter	5.06E+11	0.712596	3.61.E+11	0.13	0.33	3.584968	1,913.94	4,768.78
distillation	2.17E+12	0.712388	1.55.E+12	0.13	0.33	3.663352	3,889.59	9,666.65
rectification	3.30E+12	0.712773	2.36E+12	0.13	0.33	3.473998	3,264.77	8,216.53
oil	2.40E+12	0.711907	1.71E+12	0.13	0.33	3.760934	2,989.41	7,500.18

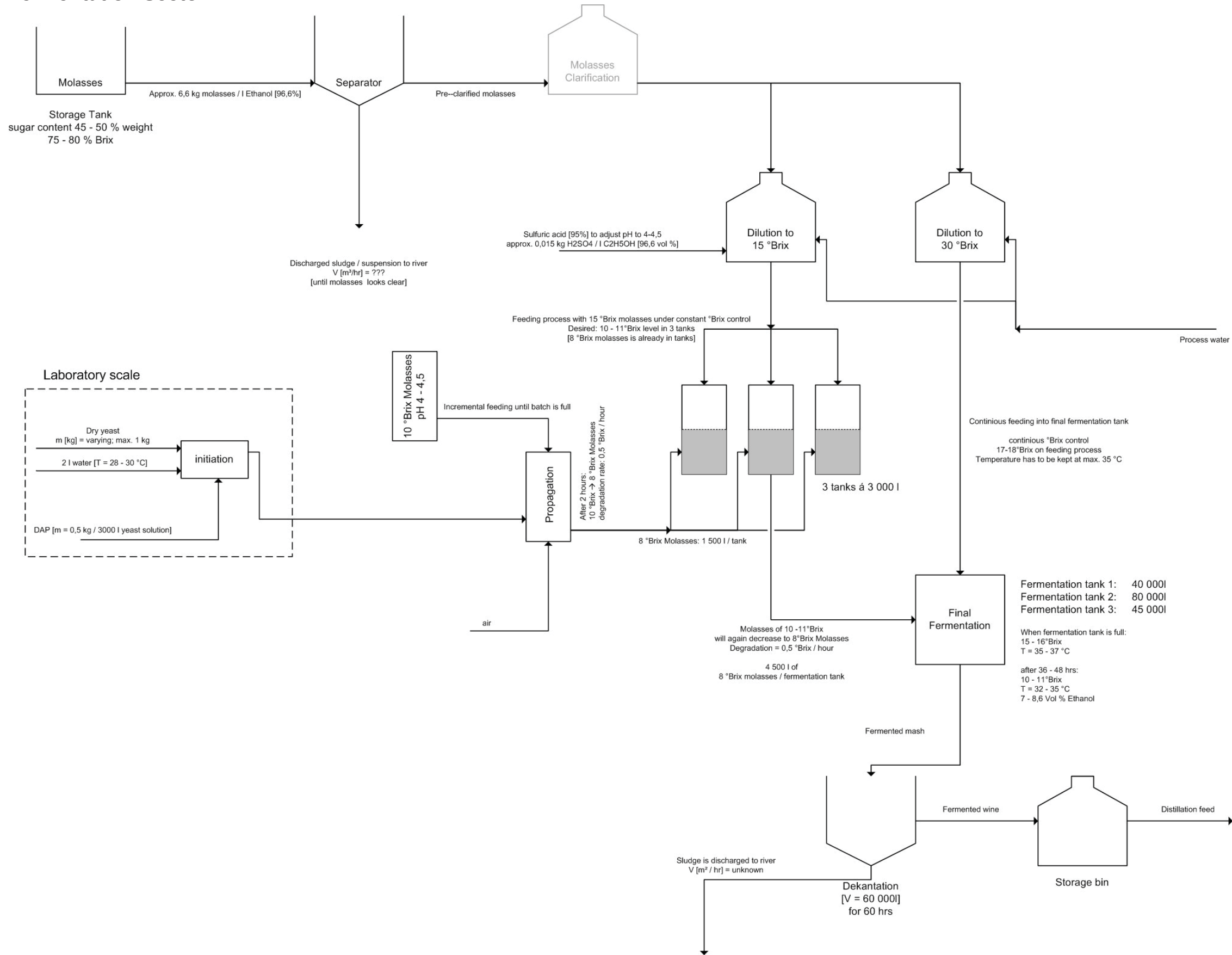


15.6 Molar flow balance – sensitivity analysis

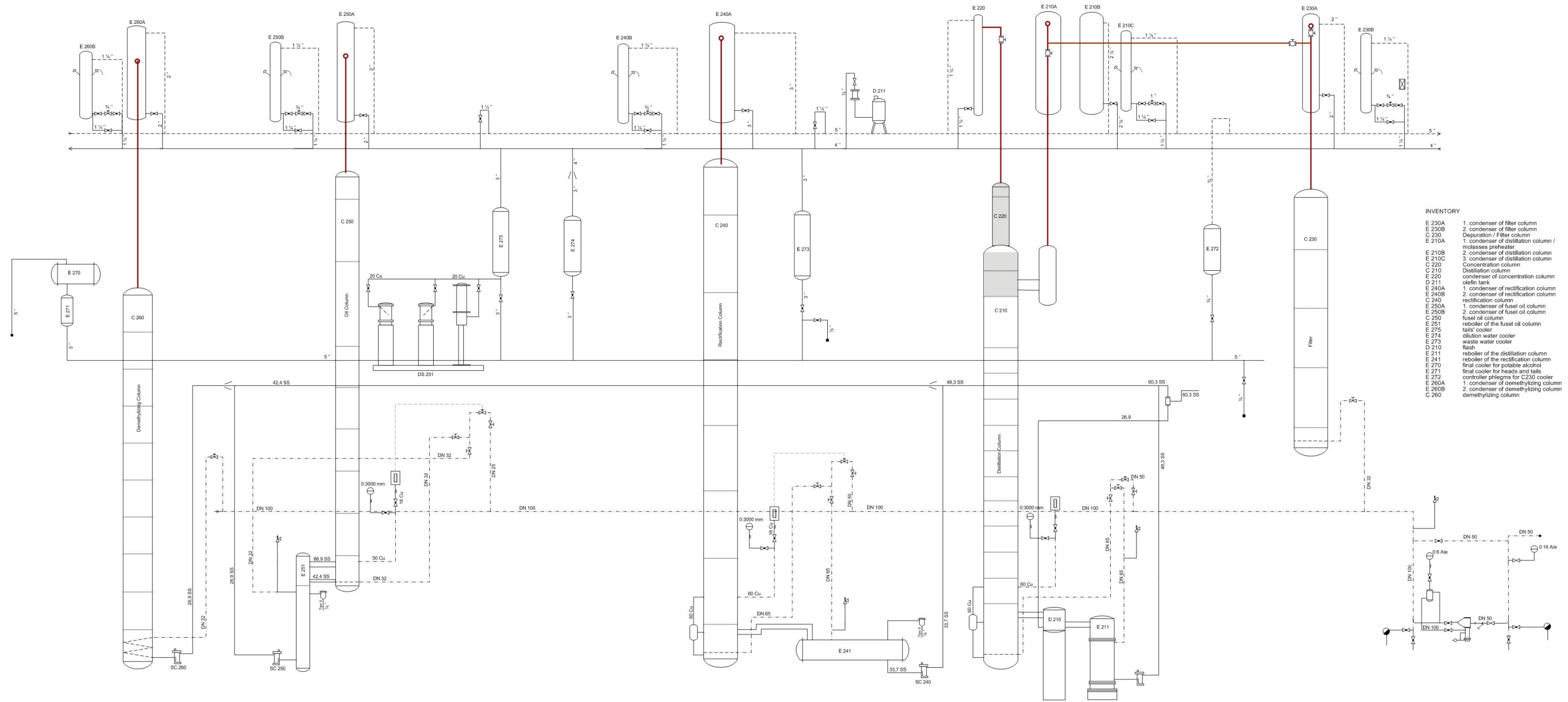




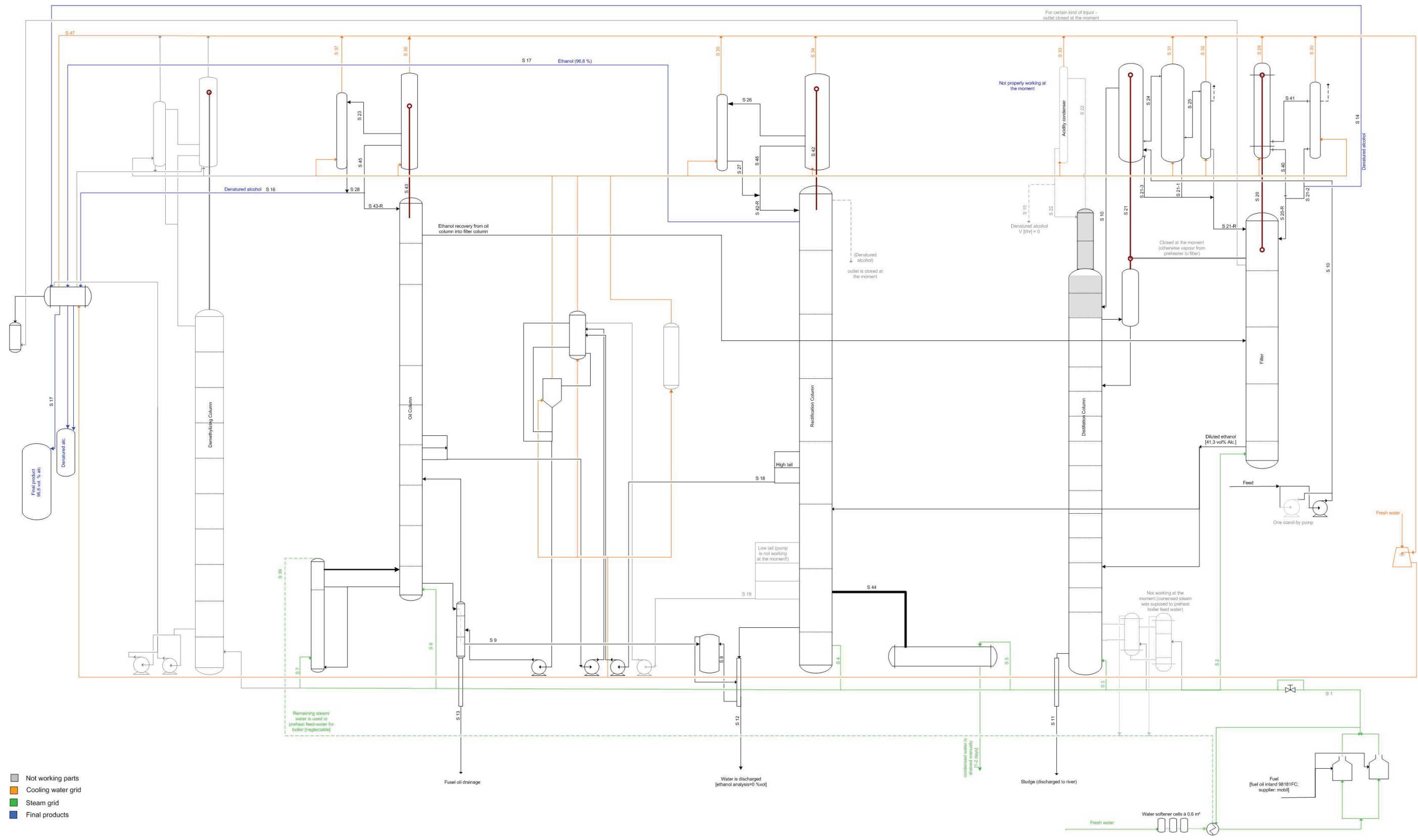
15.7 Flowsheet - Fermentation Sector



15.8 PID – Distillation



15.9 Flow sheet - Distillation Sector



15.10 Overview: streams total

Stream Nr.	Massflowrate	T _{start, average}	T _{start, average}	T _{end, average}	T _{end, average}	Heat capacity	Heat capacity	Latent Heat	CP	CP
		[°C]	[K]	[°C]	[K]	[liquid]	[vapour]		m [*] c _p	m [*] c _p
	[kg/hr]	[°C]	[K]	[°C]	[K]	[kJ/kg K]	[kJ/kg K]	[kJ/kg]	sensible	sensible
									[kJ / hr K]	[W / K]
10	3,093	25.0	298.2	70.3	343.4	4.1583			12,860	3,572.31
14	14	66.5	339.7	35.0	308.2	2.6884	1.4795	994.2141	36	10.11
16	5	66.8	339.9	35.0	308.2	2.7335	1.4914	1038.2229	15	4.10
17	119	74.5	347.7	39.0	312.2	2.7332	1.4913	980.6993	326	90.62
20	298	74.0	347.2	-	-	3.1980	1.6137	949.1849	954	264.87
21	622	70.8	343.9	-	-	3.6543	1.7390	1616.2220	2,272	631.13
21-3	342	70.8	343.9	-	-	3.6597	1.7405	1623.4536		
23	16	72.5	345.7	66.8	339.9	3.1955	1.6132	964.4553	52	14.39
24	280	70.8	343.9	51.5	324.7	3.6477	1.7372	1607.3834	1,021	283.50
26	82	76.0	349.2	42.3	315.4	3.1855	1.6099	893.1286	262	72.84
21-1	280	70.8	343.9	51.5	324.7	3.6477	1.7372	1607.3834	1,021	283.50
21-2	27	74.0	347.2	66.5	339.7	3.1765	1.6078	919.4843	86	23.89
27	82	76.0	349.2	42.3	315.4	3.1855	1.6099	893.1286	262	72.84
28	11	72.5	345.7	66.8	339.9	3.1955	1.6132	964.4553	35	9.59
S20-R	285									
S21-R	622									
S42-R	823									
S43-R	342									
29		30.8	304.0	50.8	323.9	4.1790				
30		30.8	304.0	34.8	307.9	4.1790				
31		30.8	304.0	37.5	310.7	4.1790				
34		30.8	304.0	42.3	315.4	4.1790				
35		30.8	304.0	32.8	305.9	4.1790				
36		30.8	304.0	40.3	313.4	4.1790				
37		30.8	304.0	36.8	309.9	4.1790				
47		30.8	304.0	32.5	305.7	4.1790				
40	258	74.0	347.2	-	-	3.2014	1.6146	953.8689		
41	41	74.0	347.2	66.5	339.7	3.1765	1.6078	919.4843	129	35.84
42	823	76.0	349.2	-	-	3.1981	1.6137	909.6048		
43	347	72.5	345.7	-	-	3.3523	1.6562	1179.2207		
45	331	72.5	345.7	-	-	3.3600	1.6583	1189.7352		
46	741	76.0	349.2	-	-	3.1982	1.6134	911.0791		
cooling water total	59,123	31.0	304.2	38.8	312.0	4.1790			247,073	68,631.38
feed water preheating	2,200	25.0	298.2	69.0	342.2	4.1790			9,194	2,553.83
drainage stream total	4,450	90.0	363.2	25.0	298.2	5.7692			25,673	7,131.39

15.11 Overview: heat exchanger streams

Stream Nr.	CP	CP	ΔQ known	ΔQ known	ΔQ known	ΔQ known	total ΔQ	total ΔQ	respective CW	respective CW	respective CW
	m*cp	m*cp	condensation	subcooling	product cooler	heating / cooling	for HEN	for HEN		CP	CW stream
	sensible	latent									
	[kJ / hr K]	[kJ / hr]	[kJ/hr]	[kJ/hr]	[kJ/hr]	[kJ/hr]	[kJ/hr]	[W]	[kg]	[kg]	
10	12,860					581,929	581,929	161,647			
14	36	13,461	12,449	323	1,147		12,772	3,548	1,895	7,918	S47
16	15	5,610	5,212	99	469		5,311	1,475			
17	326				11,581		13,197	3,666			
20	954							-			
21	2,272							-			
21-3		555,156	555,156				555,156	154,210			
23	52	15,636	15,636	298			15,933	4,426			
24	1,021	449,722	449,722	19,646			469,369	130,380			
26	262	73,523	73,523	8,850			82,373	22,881			
21-1	1,021	449,722	449,722	19,646			469,369	130,380	16,847	70,405	S31
21-2	86	24,898	24,898	645			25,543	7,095	2,341	9,783	S30
27	262	73,523	73,523	8,850			82,373	22,881	10,284	42,977	S35
28	35	10,424	10,424	199			10,622	2,951	644	2,693	S37
40		245,669	245,669				245,669	68,242	2,952	12,335	S29
41	129	37,347	37,347	968			38,315	10,643			
42		748,793									
43		409,602									
45		393,967	393,967				393,967	109,435	10,011	41,837	S36
46		675,006	675,006				675,006	187,502	14,148	59,125	S34
cooling water total	247,073					1,933,840	1,933,840	537,178			
feed water preheating	9,194					404,527	404,527	112,369			
drainage stream total	25,673										
											S30 = incl CW for S14
											S37 = incl CW for S16

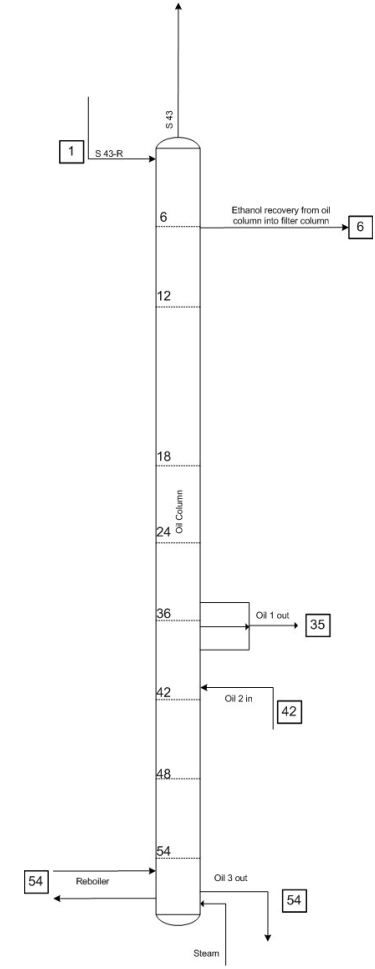
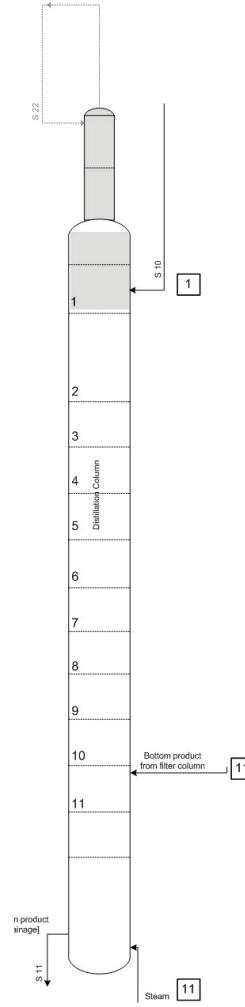
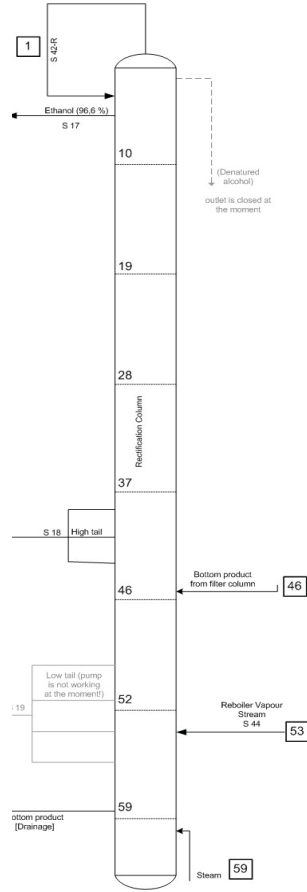
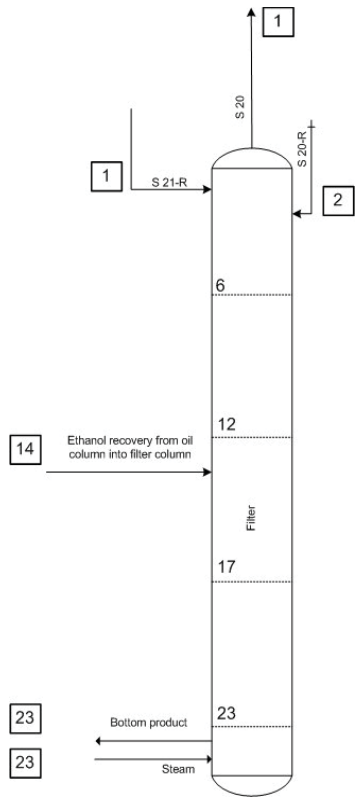
CW ... cooling water

15.12 Existing heat exchanger analysis

condenser	surface (outer)	2 CS in [tube- side]	2 T CS in	2 CS out [tubeside]	2 T CS out	1 HS in [shellside]	1 T HS in	vapour / liquid	1 HS out [shellside]	1 T HS out	vapour / liquid	T _{im}	Q 2	Q 1	actual Q	corres- ponding U
	[m ²]		[K]		[K]		[K]			[K]		[K]	[kJ/hr]	[kJ/hr]	[W]	[W / m ² K]
Preheater	22.65	S 10	298.15	S 10	343.40	S 21	343.90	vapour	S 21-3	343.90	liquid	10.02	581,929	555,156	157,928	696
									S 24	343.90	vapour			-		
D Cond 1	17.76	S31	303.98	S31	310.65	S 24	343.90	vapour	S 21-1	324.65	liquid	26.46	469,369	469,369	130,380	277
F Cond 1	11.26	S29	303.98	S29	323.90	S 20	347.15	vapour	S 40	347.15	liquid	32.19	245,669	245,669	68,242	188
									S 41	347.15	vapour			-		
F Cond 2	2.51	S30	303.98	S30	307.90	S 41	347.15	vapour	S 21-2	339.65	liquid	37.43	38,315	25,543	10,643	113
									S 14	339.65	liquid			12,772		
R Cond 1	17.76	S34	303.98	S34	315.40	S 42	349.15	vapour	S 46	349.15	liquid	39.18	675,006	675,006	187,502	269
									S 26	349.15	vapour			-		
R Cond 2	2.51	S35	303.98	S35	305.90	S 26	349.15	vapour	S 27	315.40	liquid	23.90	82,373	82,373	22,881	382
FO Cond 1	11.26	S36	303.98	S36	313.40	S 43	345.65	vapour	S 45	345.65	liquid	36.76	393,967	393,967	109,435	264
									S 23	345.65	vapour			-		
FO Cond 2	2.51	S37	303.98	S37	309.90	S 23	345.65	vapour	S 28	339.90	liquid	35.83	15,933	10,622	4,426	49
									S 16	339.90	liquid			5,311		
Product cooler	no data av,	S47	303.98	S47	305.65	S 14	339.65	liquid	S 14	308.15	liquid	14.21	13,197	1,147	3,666	---
Product cooler	no data av,	S47	303.98	S47	305.65	S 16	339.65	liquid	S 16	308.15	liquid			469		
Product cooler	no data av,	S47	303.98	S47	305.65	S 17	347.65	liquid	S 17	312.15	liquid			11,581		

CS...cold side
HS...hot side

15.13 Column modelling



Following notation and algorithm have been applied to model the prevailing columns:

L	molar liquid stream
V	molar vapour stream
x	molar liquid composition
y	molar vapour composition

Every column algorithm starts from the top stage, where vapour and reflux composition as well as the molar flowrates are known. Assuming constant molar flow, the only two unknown parameter are $x(i+1)$ and $y(i+1)$, which are incoming vapour and outgoing liquid composition. As the stage equilibrium model is based on the assumption that all streams leaving a stage are in equilibrium with each other, $x(i+1)$ can be calculated as function from $y(i)$. As an overall efficiency factor was implemented, the predicted liquid stream composition is denoted as $x_{peff}(i)$. The remaining unknown composition of the entering vapour stream from the stage below is calculated out of the stage component balance. Sidestreams and additional feed streams are considered at the respective stage, by correcting the molar flowrates accordingly.

DISTILLATION column (11 stages)

1		$L2 = L1A + V1B$ $V2 = V1 + V1B$ $x_{L2} = f(y1) \rightarrow x_{peff2}$ $y_{V2} = (V1 \cdot y1 + L2 \cdot x_{peff2} - L1A \cdot x_{L1A}) / V2$
2-10		<p>For $i = 2 : 10$</p> $x_{L(i+1)} = f(y_i) \rightarrow x_{peff(i+1)}$ $y_{V(i+1)} = (V2 \cdot y(i) + L2 \cdot x_{peff(i+1)} - L2 \cdot x_{peff(i)}) / V2$
11		<p>As result Bottom composition x_e and steam composition y_e after the bottom stage will be calculated by the program, to see which specifications can be achieved with varying input (composition, flowrate) and efficiency values. B_F = Bottom Stream from Filter column</p>

The approximation of the steam condensation in the first stage involves an uncertainty factor, as it remains unknown, whether the subcooled liquid will really be heated up in the first stage or will be distributed over more stages. In that case the vapour-liquid ratio would be slightly different.



FILTER column (23 stages)

1		$L2 = L1A + V1B$ $V2 = V1 + V1B$ $x_{L2} = f(y1) \rightarrow x_{peff2}$ $y_{V2} = (V1 \cdot y1 + L2 \cdot x_{peff2} - L1A \cdot x_{L1A}) / V2$
2		$L3 = L2 + S20_R$ $V3 = V2$ $x_{L3} = f(y2) \rightarrow x_{peff3}$ $y_{V3} = (V2 \cdot y2 + L3 \cdot x_{peff3} - L2 \cdot x_{peff2} - S20_R \cdot x_{S20R}) / V3$
3-13		<p>For $i = 3 : 13$</p> $x_{L(i+1)} = f(yi) \rightarrow x_{peff(i+1)}$ $y_{V(i+1)} = (V3 \cdot y(i) + L3 \cdot x_{peff(i+1)} - L3 \cdot x_{peff(i)}) / V3$
14		$L15 = L3 + OIL_REC$ $V15 = V3$ $x_{L15} = f(y14) \rightarrow x_{peff15}$ $y_{V15} = (V3 \cdot y14 + L4 \cdot x_{peff15} - L3 \cdot x_{peff14} - OIL_REC \cdot x_{OIL_REC}) / V15$
15-22		<p>For $i = 15 : 22$</p> $x_{L(i+1)} = f(yi) \rightarrow x_{peff(i+1)}$ $y_{V(i+1)} = (V15 \cdot y(i) + L15 \cdot x_{peff(i+1)} - L15 \cdot x_{peff(i)}) / V15$
23		<p>As result the bottom composition x_e and steam composition y_e after the bottom stage will be calculated by the program, to see which specifications can be achieved with varying input (composition, flowrate) and efficiency values.</p>

FILTER column (23 stages) corrected modell

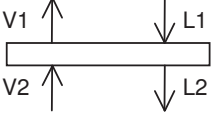
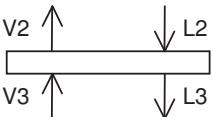
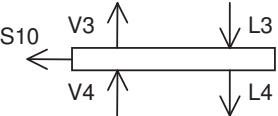
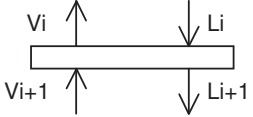
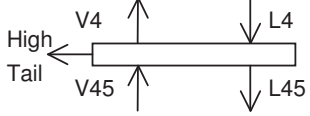
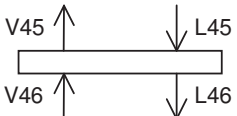
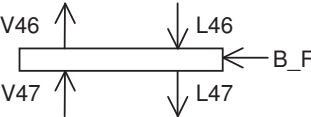
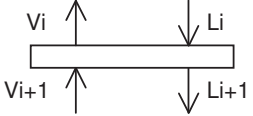
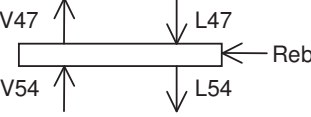
As the filter column shows misplaced feed stage locations, the modelling of the column with the stage equilibrium model was not feasible and the result displays high deviations from the actual behaviour inside the column. Especially the leaving streams from first and second stage can not be in equilibrium, considering the feed streams that are entering the column in the upper two stages with completely different compositions. As a solution the two upper stages were combined in one and it was tried to bring the further stage compositions back to the 45° line, as this is the maximum achievable driving force. Everything beyond this line would be impossible. Applying this correction algorithm gives a somehow reasonable picture of the column, but it has to be mentioned, that only further investigations and feed stage correction mechanism will give valuable information about the internal column processes.

For the time being the problem was identified and the overhead streams, which have to be condensed, independently of the feed stage, have been taken as they are for the Heat Exchanger Network development.

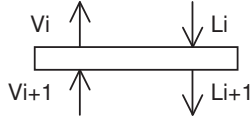
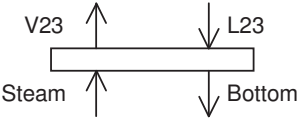
Feed stage corrections however will influence composition and flowrates of the single streams, so a final valuable Heat Exchanger Network can only be designed based on a prior process optimization if an overall good performance is supposed to be achieved.



RECTIFICATION column (59 stages)

1		$L1 = V1 = L2 = V2$ $x_{L2} = f(y1) \rightarrow x_{peff2}$ $y_{V2} = (V1 \cdot y1 + L2 \cdot x_{peff2} - L1 \cdot x_{L1}) / V2$
2		$L3 = L1$ $V3 = V1$ $x_{L3} = f(y2) \rightarrow x_{peff3}$ $y_{V3} = (V2 \cdot y2 + L3 \cdot x_{peff3} - L2 \cdot x_{peff2}) / V3$
3		$L4 = L3 - S10$ $V4 = V3$ $x_{L4} = f(y3) \rightarrow x_{peff4}$ $y_{V4} = (V3 \cdot y3 + L4 \cdot x_{peff4} + S10 \cdot x_{S10} - L3 \cdot x_{peff3}) / V4$
4-43		For $i = 4 : 43$ $x_{L(i+1)} = f(yi) \rightarrow x_{peff(i+1)}$ $y_{V(i+1)} = (V4 \cdot y(i) + L4 \cdot x_{peff(i+1)} - L4 \cdot x_{peff(i)}) / V4$
44		$L45 = L4 - \text{High_Tail}$ $V45 = V4$ $x_{L45} = f(y44) \rightarrow x_{peff45}$ $y_{V45} = (V4 \cdot y44 + L45 \cdot x_{peff45} + \text{High_Tail} \cdot x_{\text{High_Tail}} - L4 \cdot x_{peff44}) / V45$
45		$L46 = L45$ $V46 = V45$ $x_{L46} = f(y45) \rightarrow x_{peff46}$ $y_{V46} = (V45 \cdot y45 + L46 \cdot x_{peff46} - L45 \cdot x_{peff45}) / V46$
46		$L47 = L46 + \text{Bottom_Filter}$ $V47 = V46$ $x_{L47} = f(y46) \rightarrow x_{peff47}$ $y_{V47} = (V46 \cdot y46 + L47 \cdot x_{peff47} - L46 \cdot x_{peff46} - B_F \cdot x_{B_F}) / V47$ $B_F = \text{Bottom Stream from Filter column}$
47-52		For $i = 47 : 52$ $x_{L(i+1)} = f(yi) \rightarrow x_{peff(i+1)}$ $y_{V(i+1)} = (V47 \cdot y(i) + L47 \cdot x_{peff(i+1)} - L47 \cdot x_{peff(i)}) / V47$
53		$L54 = L47$ $V54 = V47 - \text{Reb}$ $x_{L54} = f(y53) \rightarrow x_{peff54}$ $y_{V54} = (V47 \cdot y53 + L54 \cdot x_{peff54} - L47 \cdot x_{peff53} - \text{Reb} \cdot x_{\text{Reb}}) / V54$ $\text{Reb} = \text{Reboiler stream}$



54-58		<p>For $i = 54 : 58$ $x_{L(i+1)} = f(y_i) \rightarrow x_{peff(i+1)}$ $y_{V(i+1)} = (V_{54} \cdot y(i) + L_{54} \cdot x_{peff(i+1)} - L_{54} \cdot x_{peff(i)}) / V_{54}$</p>
59		<p>As result the bottom composition x_e and steam composition y_e after the bottom stage will be calculated by the program, to see which specifications can be achieved with varying input (composition, flowrate) and efficiency values.</p>



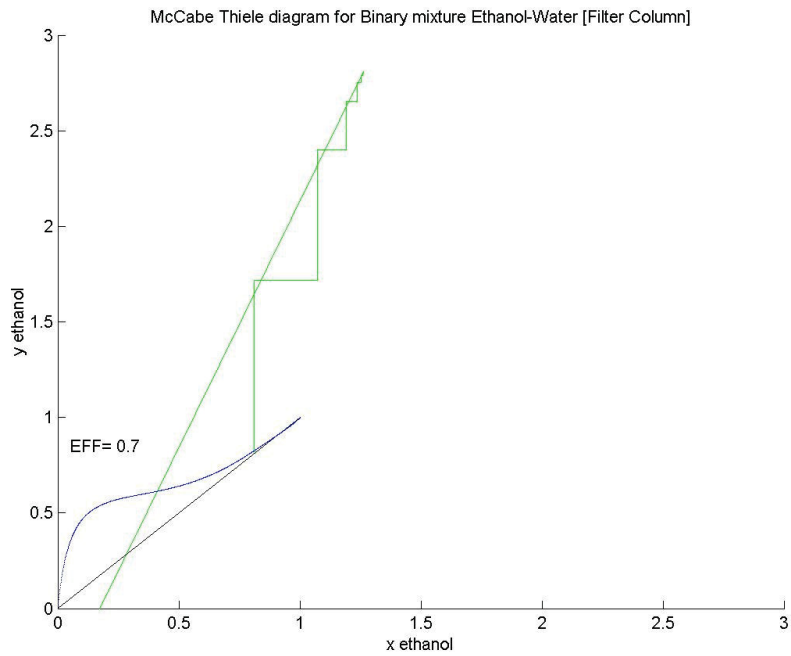
OIL column (54 stages)

1		$L2 = L1$ $V2 = V1$ $x_{L2} = f(y_1) \rightarrow x_{peff2}$ $y_{V2} = (V1 \cdot y_1 + L2 \cdot x_{peff2} - L1 \cdot x_{L1}) / V2$
2-5		For $i = 2 : 5$ $x_{L(i+1)} = f(y_i) \rightarrow x_{peff(i+1)}$ $y_{V(i+1)} = (V2 \cdot y(i) + L2 \cdot x_{peff(i+1)} - L2 \cdot x_{peff(i)}) / V2$
6		$L7 = L2 - OIL_REC$ $V7 = V2$ $x_{L7} = f(y_6) \rightarrow x_{peff7}$ $y_{V7} = (V2 \cdot y_6 + L7 \cdot x_{peff7} - L2 \cdot x_{peff6} + OIL_REC \cdot x_{OIL_REC}) / V7$
7-34		For $i = 7 : 34$ $x_{L(i+1)} = f(y_i) \rightarrow x_{peff(i+1)}$ $y_{V(i+1)} = (V7 \cdot y(i) + L7 \cdot x_{peff(i+1)} - L7 \cdot x_{peff(i)}) / V7$
35		$L36 = L7 - OIL_1_out$ $V36 = V7$ $x_{L36} = f(y_{35}) \rightarrow x_{peff36}$ $y_{V36} = (V7 \cdot y_{35} + L36 \cdot x_{peff36} - L7 \cdot x_{peff35} + OIL_1_out \cdot x_{OIL_1_out}) / V36$
36-41		For $i = 36 : 41$ $x_{L(i+1)} = f(y_i) \rightarrow x_{peff(i+1)}$ $y_{V(i+1)} = (V36 \cdot y(i) + L36 \cdot x_{peff(i+1)} - L36 \cdot x_{peff(i)}) / V36$
42		$L43 = L36$ $V43 = V36 + OIL_2_in$ $x_{L43} = f(y_{42}) \rightarrow x_{peff43}$ $y_{V43} = (V36 \cdot y_{42} + L43 \cdot x_{peff43} - L36 \cdot x_{peff42} - OIL_2_in \cdot x_{OIL_2_in}) / V43$
43-53		For $i = 43 : 53$ $x_{L(i+1)} = f(y_i) \rightarrow x_{peff(i+1)}$ $y_{V(i+1)} = (V43 \cdot y(i) + L43 \cdot x_{peff(i+1)} - L43 \cdot x_{peff(i)}) / V43$
54		As result the bottom composition x_e and steam composition y_e after the bottom stage will be calculated by the program, to see which specifications can be achieved with varying input (composition, flowrate) and efficiency values.

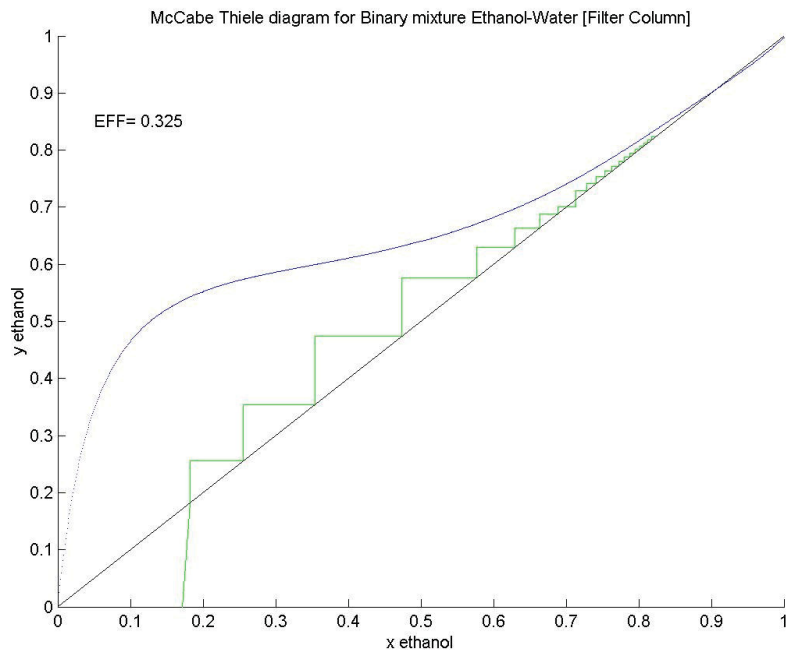


Following diagrams will be generated, when running the mentioned *matlab* file:

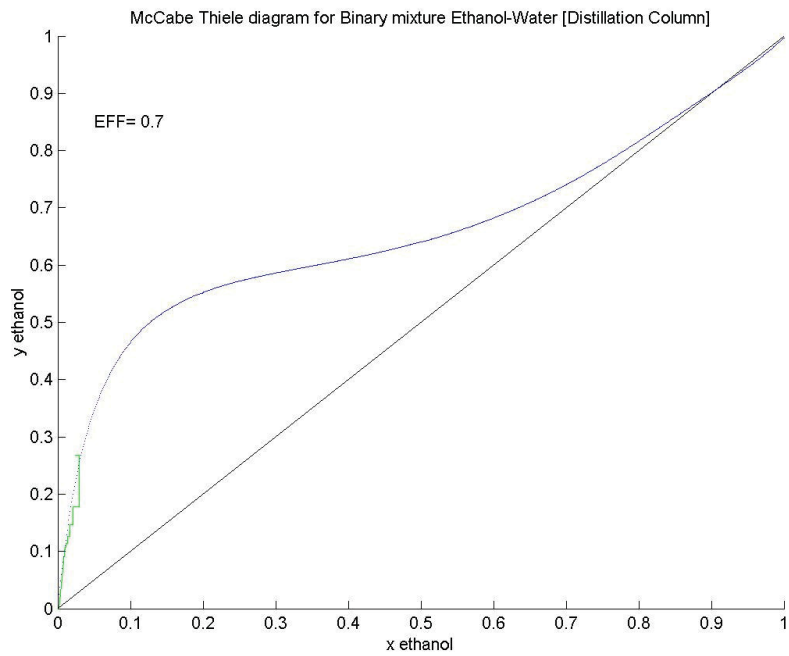
Filter_real:



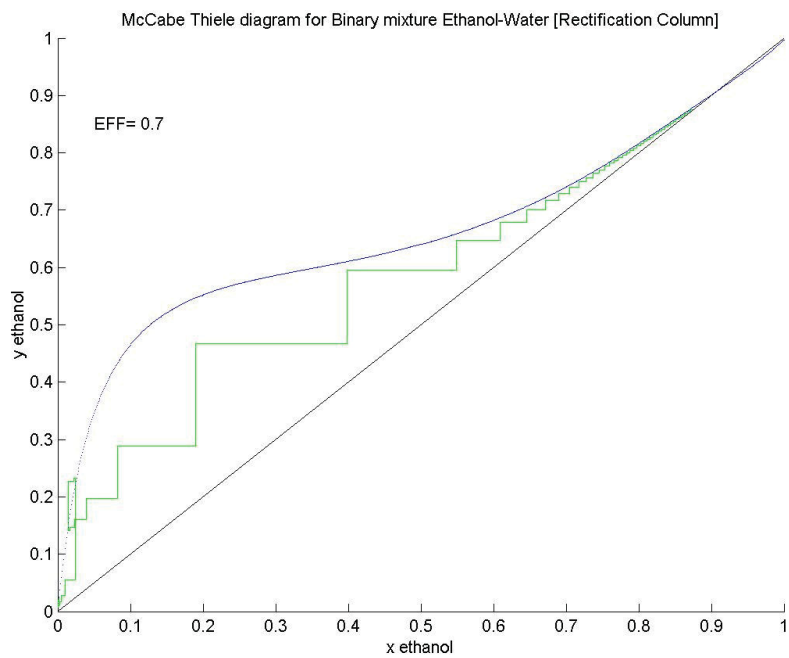
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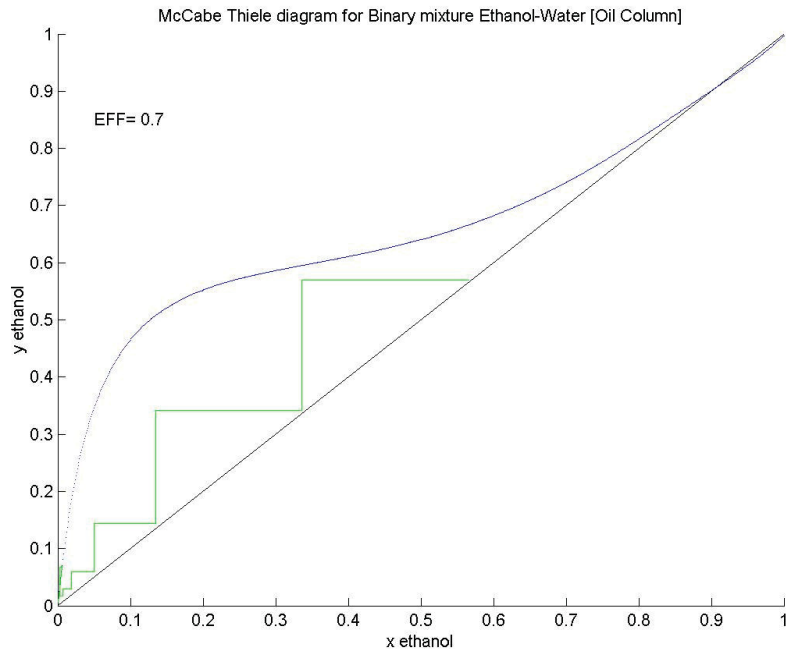
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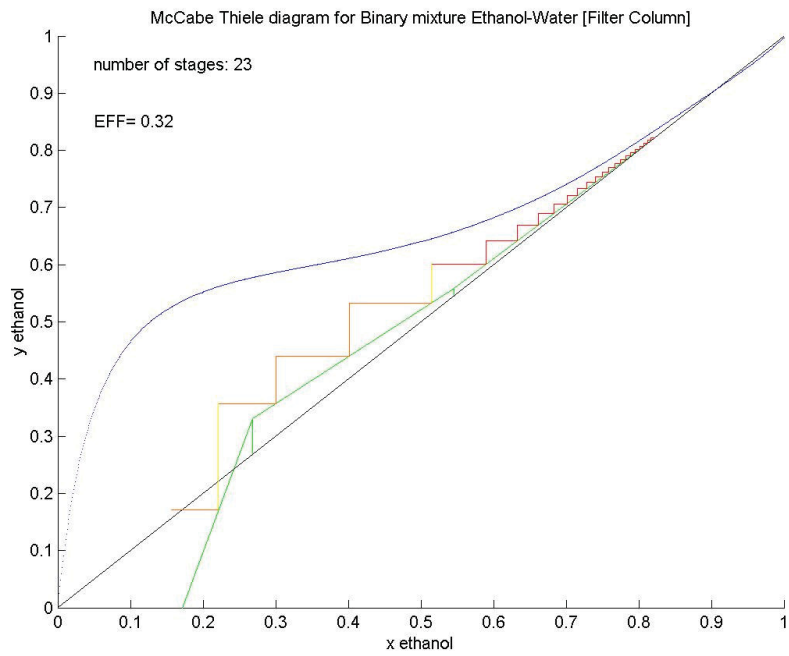
Rectification_real:



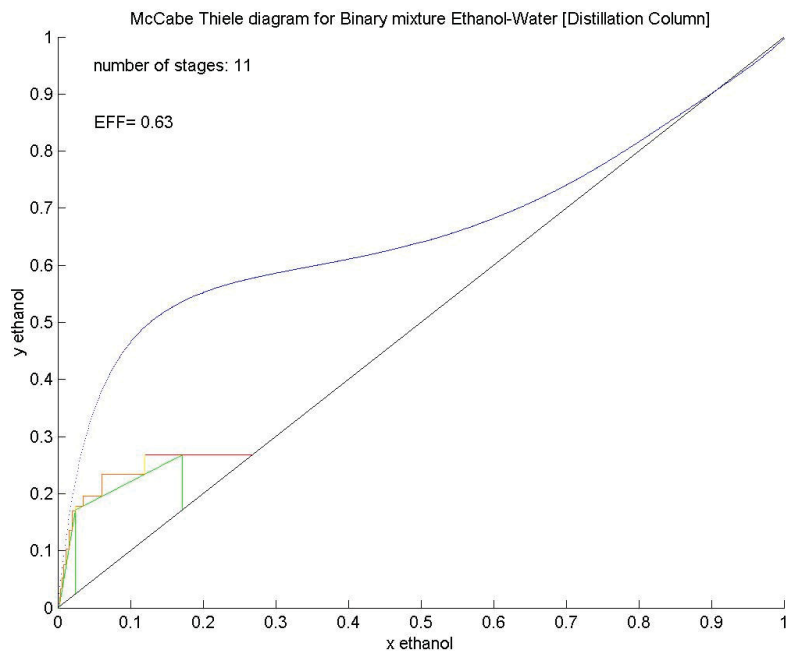
Oil_real:



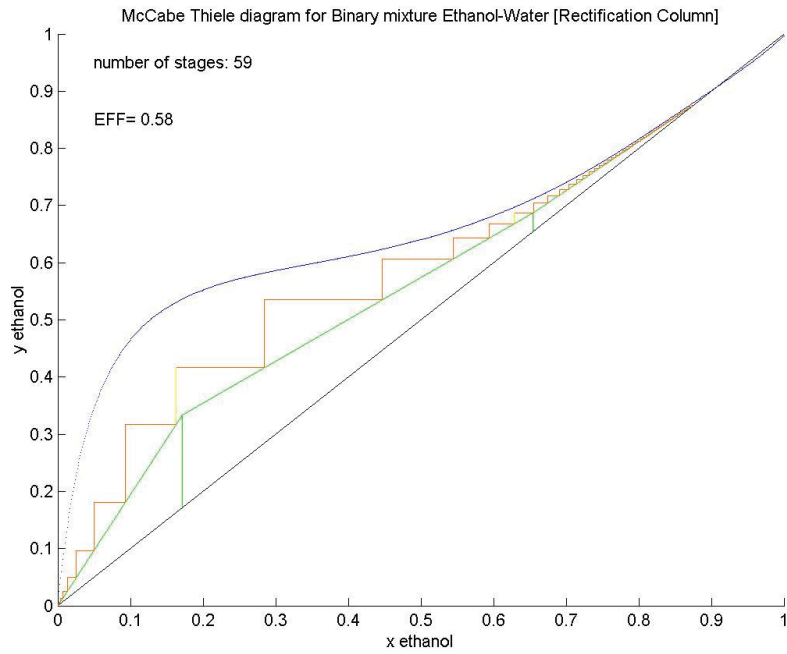
McCabeThieleFilter:



McCabeThieleDistillation:



McCabeThieleRectification:



McCabeThieleOil:

