Enhancing Refining Yield: Mitigating Losses Through Process Sychronization

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Abtsract

The dissertation "Enhancing Refining Yield: Mitigating Losses Through Process Sychronization" is part of a consulting project in the area of process analysis and improvement. It portrays the reality of a company that operates in the food sector, more specifically in the sugar industry, and that boasts the particularities of a process industry.

Founded in 1962, Company X (pseudonym used for confidentiality purposes) is deeply committed to the production of sugar and has garnered extensive experience in the process industry. Within this industry, substantial investments are required, underscoring the objective of maximizing production output while maintaining optimal performance.

As a result of these circumstances, the Kaizen Institute intervened with two well-defined objectives: The primary aim was to enhance the Overall Equipment Efficiency (OEE) of white sugar production from 69.00% to 80.00%, representing a notable increase of 15.9% and, secondly, an equivalent improvement was targeted for the OEE of brown sugar, with the goal of elevating it from 42.00% to 48.50%. These two operational objectives encompass a larger strategic objective: increasing the company's market competitiveness. This objective begins with the ability to compete in the commodity sector, which requires sustaining low costs per ton in order to offer products at competitive prices and increase market share. Second, responsiveness to the spot market is crucial, necessitating a high degree of production flexibility to promptly meet fluctuating demand. Lastly, there is a strong emphasis on capturing the market for products with added value (PVA), highlighting the importance of offering differentiated, high-quality products that meet the specific requirements of customers.

Company X was aware that a bottleneck had developed in their operational process, impeding their ability to achieve intended results. However, they encountered significant difficulties in precisely assessing the extent of the bottleneck's influence on the process, as well as in conceiving a suitable solution and precisely quantifying the potential benefits. In view of these obstacles, the Kaizen Institute was established with the intention of overcoming them. The primary objective of the institute was to address, design, and quantify the aforementioned problem and its solution. The initial stage encompassed the creation of a traditional simulator that accurately modeled the process, incorporating both the mass and energy aspects. Resorting to this simulator, a comprehensive evaluation was conducted to effectively characterize the existing and desired states. By conducting this analysis, both mass and energy limitations within the process were quantified. These bottlenecks prevented the refinery from expanding its refining rate and reaching the desired OEE. Therefore, an investigation and analysis were conducted to identify the most effective strategies for overcoming these obstacles and facilitating the achievement of the intended OEE targets.

The concluding phase of this study revolves around the evaluation of the proposed solutions in order to determine their adequacy in attaining the predefined objectives. The primary focus is to assess whether these solutions possess the capability to effectively enhance the OEE of the process and to quantify the impact of them, both at production and financial level.

Resumo

A dissertação "Melhoria da Produtividade da Refinação: Mitigação de Perdas Através da Sincronização de Processos" faz parte de um projeto de consultoria na área de análise e melhoria de processos. Retrata a realidade de uma empresa que atua no setor alimentar, mais especificamente na indústria açucareira, e que possui as particularidades de uma indústria de processo.

Fundada em 1962, a Empresa X (pseudônimo utilizado por motivos de confidencialidade) está profundamente comprometida com a produção de açúcar e acumulou ampla experiência na indústria de processos. Dentro desta indústria, são necessários investimentos substanciais, destacando o objetivo de maximizar a produção enquanto mantêm um desempenho ótimo.

Como resultado dessas circunstâncias, o Kaizen Institute interveio com dois objetivos bem definidos: o objetivo principal de melhorar a Eficiência Geral dos Equipamentos (OEE) da produção de açúcar branco de 69,00% para 80,00%, representando um aumento notável de 15,9%. Em segundo lugar, foi definida uma melhoria equivalente para o OEE do açúcar amarelo, com o objetivo de elevá-lo de 42,00% para 48,50%. Esses dois objetivos operacionais englobam um objetivo estratégico maior: aumentar a competitividade da empresa no mercado. Alcançar esse objetivo requer vários fatores e estratégias, começando pela capacidade de competir no setor de commodities, que exige a manutenção de custos baixos por tonelada para oferecer produtos a preços competitivos e aumentar a participação de mercado. Em segundo lugar, a capacidade de resposta ao mercado spot é crucial, exigindo um alto grau de flexibilidade na produção para atender prontamente à procura flutuante. Por fim, há um forte destaque em conquistar o mercado de produtos de valor acrescentado (PVA), realçando a importância de oferecer produtos diferenciados e de alta qualidade que atendam às necessidades específicas dos clientes.

A empresa X estava ciente de que se tinha desenvolvido um bottleneck no seu processo operacional, que impedia a sua capacidade de alcançar os resultados pretendidos. No entanto, a equipa encontrou dificuldades significativas para avaliar com precisão a extensão da influência do gargalo no processo, bem como para conceber uma solução adequada e quantificar com precisão os benefícios potenciais. Perante estes obstáculos, o objetivo principal do instituto Kaizen era abordar, conceber e quantificar o problema acima mencionado e a respetiva solução. A fase inicial englobou a criação de um simulador tradicional representativo do processo, que incorporava tanto os balanços de mássicos como de energéticos. Utilizando este simulador, foi efectuada uma avaliação abrangente para caraterizar eficazmente os estados existentes e desejados. Ao efetuar esta análise, as limitações de massa e energia no processo foram identificadas como obstáculos. Estes bottlenecks impedem a refinaria de expandir a sua taxa de refinação e de atingir o OEE desejado. Por conseguinte, foi realizada uma investigação e uma análise para identificar as estratégias mais eficazes para ultrapassar estes obstáculos e facilitar a obtenção dos objectivos de OEE pretendidos.

A fase conclusiva deste estudo foca-se em torno da avaliação das soluções propostas para determinar a sua viabilidade em atingir os objetivos predefinidos. O principal objetivo é avaliar se as soluções exploradas possuem a capacidade de melhorar efetivamente o OEE do processo e quantificar o seu impacto, tanto em termos de produção como financeiros.

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"Happiness lies in the joy of achievement and the thrill of creative effort"

Franklin D. Roosevelt

Contents

Acronyms and Symbols

KPI	Key Performance Indicator
OEE	Overall Equipment Efficiency
EU	European Union
TPM	Total Productive Maintenance
MTTR	Mean Time to Repair
MTBF	Mean Time Between Failures
TSS	Total Suspended Solids
DWS	Dry White Sugar
RCM	Refined Cooked Mass
AS	Affination Syrup
SC	Sugar Cane
ASC	Affined Sugar Cane
RL	Recovery Liquor
FL	Filtered Liquor
CS	Carbonated Sludge
BL	Brown Liquor
FIL	Final Liquor
RS	Refining Syrup
WWS	Wet White Sugar

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Chapter 1

Introduction

This dissertation presents a process improvement project conducted within a national company operating in the food sector, specifically in the production of white and brown sugar. The company primarily operates in the Iberian Peninsula, with a significant volume of exports to other regions in Europe. The project was executed at the company's facilities in Portugal, where a series of activities were implemented in collaboration with multidisciplinary teams. These actions were initiated using a set of methodologies rooted in the principles of Kaizen philosophy, with the objective of achieving gradual enhancements in the production processes. Furthermore, efforts were made to establish conditions that would facilitate the perpetuation of these improvements over the long term, accomplished through the active involvement of self-directed teams.

1.1 Project Framework and Motivation

The work was performed at Kaizen, a continuous development consulting firm that assists other companies in achieving their goals through a unique methodology. Kaizen is a business philosophy and method that originated in Japan, primarily associated with the Toyota Production System (TPS) and Lean Manufacturing. The term "Kaizen" translates to "continuous improvement" in English. It emphasizes the ongoing pursuit of incremental improvements in all aspects of an organization's operations, processes, and systems.

This research was conducted within a European company that functions as one of the region's few sugarcane-derived sugar producers. The objective of the initiative was to increase the refining capacity for both white and brown sugar by identifying and eliminating manufacturing process bottlenecks. The goal was to achieve an Overall Equipment Effectiveness (OEE) of 80.00% for white sugar, which corresponds to a refining rate of 23.57 ton/h, and an OEE of 48.00% for brown sugar, which corresponds to a refining rate of 0.97 ton/h. To accomplish these goals, a variety of methodologies were developed to address bottlenecks in the sugar refining process in a systematic manner. In addition, the project centered on optimizing and standardizing critical process parameters in order to improve overall efficacy and uniformity. By implementing these strategies, the company intended to achieve a broader strategic goal, which is to enhance its market competitiveness. This entails maintaining a strong position in the commodity sector by consistently achieving low costs per ton. By doing so, the company can offer products at competitive prices, thereby attracting customers and expanding its market share. These operational efforts align with the overall objective of bolstering the company's market competitiveness and ensuring its long-term success in the industry.

1.2 Kaizen Institute

The Kaizen Institute is a management consulting company with a focus on efficiency and process improvement. During World War II, Japan created the idea of kaizen, or "continuous improvement," in an effort to revive the economy of the nation. The Kaizen Institute was established in 1985 by Japanese consultant and author Masaaki Imai with the goal of popularizing Kaizen and helping businesses use its ideas (kai (2015)).

The institution initially concentrated on advising and training for businesses in Japan. Yet, when Japanese businesses started to grow abroad in the 1990s, it began to expand internationally. The institution today has operations in over 30 nations and has collaborated with organizations from a variety of industries, as can be seen in Figure 1.1.

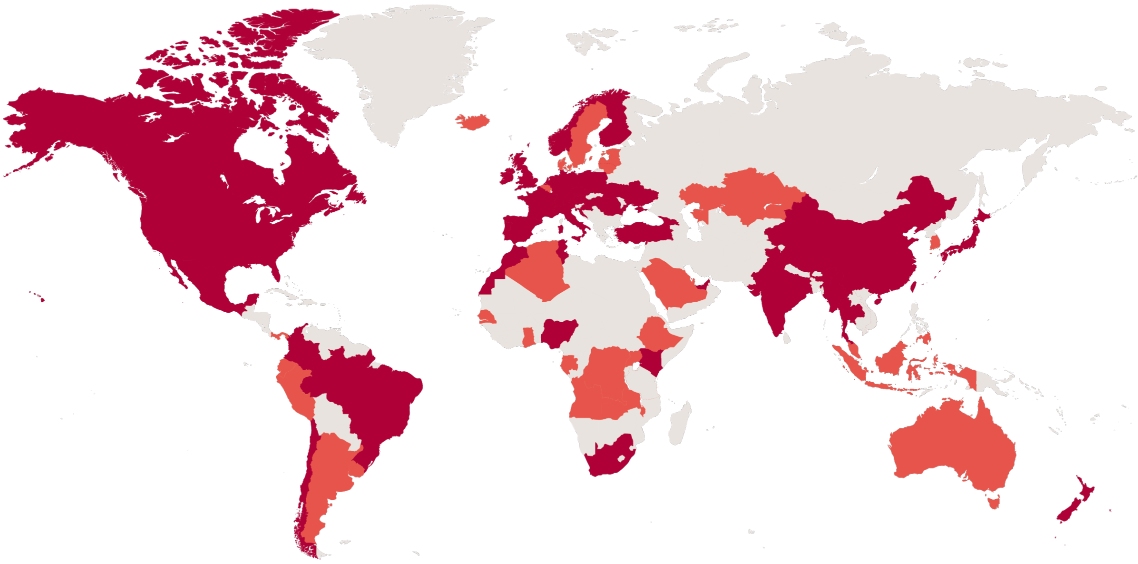


Figure 1.1: Global presence of Kaizen (Adapt from kai (2015))

To put it in perspective, Kaizen is a Japanese phrase made up of two terms with the same roots: Kai, which means change, and Zen, which means better in Japanese. Hence, Kaizen refers to ongoing improvement (Imai (1986)), or more specifically, the ongoing improvement of every individual, every day, in every way. The institute's method of fostering small, gradual improvements in systems and procedures places an emphasis on employee engagement and empowerment. In addition, the institute has created a variety of tools and processes to aid in continuous improvement, such as the Kaizen Blitz, a targeted improvement activity that brings together cross-functional teams.

The Kaizen Institute has made a substantial contribution to spreading the idea of Kaizen and helping businesses all around the world improve their systems and processes. The success of businesses across several industries has been aided by its emphasis on continual development and employee empowerment.

1.2.1 Kaizen Foundations

The application of kaizen methodologies necessitates a challenging attitude toward existing paradigms, requiring individuals and organizations to alter the manner in which they act and solve problems. In this way, kaizen entails the adoption of a well-defined culture with a set of guiding principles. A set of guiding principles for behavior:

- Create customer value: Value is what the customer is willing to pay for. Thus, the first Kaizen Principal holds that any product or service that is developed by an organization should aim to match market needs and meet customer expectations.
- Create flow efficiency: Flow efficiency can be obtained through the elimination of three Ms: Muda (waste), Mura (unevennes) and Muri (overburden). Eliminating these stages reduces production times, optimizes the equipment, and boosts productivity (Imai (2012)).
- Be Gemba oriented: Gemba is the Japanese word for workplace, so it is the place where value is added. This principle emphasizes the significance of traveling to the actual site of an organization's problems in order to identify their fundamental causes and devise solutions.
- Empower people: For a kaizen strategy to be successful, it is crucial to comprehend the significance of the role of the organization's people, and the strategy should center on them. Consider the complaints, problems, and opinions of every level of the organization, from administrators to employees, and give them responsibility and trust.
- Be scientific and transparent: Always support facts and reasoning with data. Utilize visual management to convey messages and ideas intuitively.

1.3 Company X

The Portuguese sugar refining industry was made up of a small number of units up to the 1960s, the majority of which were small-scale, had extremely basic equipment, and couldn't produce high-quality sugar. According to the then-in-effect industrial conditioning strategy, the Sugar Refining Industrial Sector had to change in order to accommodate the construction of modern, bigger, and technologically advanced sugar refining plants. Due to the clustering of nine small sugar refineries in the country's north, Company X was established in 1962. Up to the start-up of the refinery built to replace these units, the corporation started marketing the current production of these tiny units.

1.3.1 Company X Market Characterization

The European Union (EU) is a major player on the sugar market, especially in terms of sugar beet production. The majority of the EU's sugar beet cultivation occurs in the most competitive producing regions of the north, including France, Germany, the Netherlands, Belgium, and Poland. Sugar beet requires a three to four year crop rotation to prevent pests and diseases, which results in seasonal production (Fikry et al. (2021)). Consequently, fluctuations in sugar beet production have an effect on the availability of beet sugar, especially white sugar. The EU produced 13.5 million tonnes of sugar beets in 2019 and 12.80 million tonnes in 2020, based on Commission (2023) report. The sugar market has experienced significant price increases in recent years. In April 2023, white sugar reached a contract high of 720.10€ per tonne, the highest price since October 2011. In addition, the price of raw sugar peaked in April 2023 at 26.99 cents, the highest since November 2011. These price hikes are part of a larger trend of rising food prices, with sugar experiencing a significant price increase. Since August 2021, monthly data reveals an increase in the annual rates of change, which reached 61.00% in February and March 2023 compared to the same months in 2022 (eur (2023)).

According to sta (2023), in 2022/23, global sugar production was approximately 177 million metric tons, with Asia being the largest sugar-producing region, led by India, Thailand, and China. Brazil, the world's leading producer of sugar cane in 2021, is expected to increase its output to approximately 42 million metric tons in 2023/24. The EU ranks third in global production of beet sugar, behind India and Brazil. However, European farming practices have changed as farmers switch from sugar beet to more lucrative crops, resulting in a slight decline in EU sugar production.

In conclusion, the EU is a significant player on the sugar market, particularly in the production of beet sugar. The availability of beet sugar is affected by seasonal fluctuations in sugar beet production, while the sugar market experiences price increases. The EU continues to be a major player in global sugar production, but changes in agricultural practices are causing slight decreases in beet sugar production.

1.4 Project's Main Objectives and Method followed in the project

This research project's primary objective is to achieve specific operational efficiency goals for the refining of white and brown sugar. The objective is to achieve an OEE of 80.00% for white sugar, which corresponds to a refining rate of 23.57 ton/h, and an OEE of 48.00% for brown sugar, which corresponds to a rate of 0.97 ton/h.

To accomplish these goals, the research will employ a variety of strategic approaches. Initially, a comprehensive analysis will be conducted to evaluate the current status of the sugar refining process for both white and brown sugar, with an emphasis on identifying improvement areas. This analysis will assess production capacity, equipment efficacy, and overall performance, among other variables.

Additionally, the research will investigate the significant categories of losses that affect the efficiency of equipment, including aspects of availability, quality, and speed. By understanding these losses, targeted improvement measures can be implemented to increase the equipment's overall efficacy. Consequently, this assessment is going to help identifying bottlenecks associated with mass flow and energy consumption in the sugar refining procedure. After identifying these bottlenecks, the research will suggest solutions to eliminate or mitigate their effects, thereby optimizing the overall efficacy of the process. The research will also focus on the standardization and optimization of process parameters such as brix and flow rates. The research aims to increase the efficacy and consistency of the sugar refining process by establishing optimized and standardized parameters.

Through the implementation of these strategies, this dissertation intends to improve the operational efficiency of white and brown sugar refining, ultimately leading to the achievement of the desired OEE levels and refining rates for each sugar type.

1.5 Thesis Structure

A dissertation's structure is crucial for organizing and presenting research findings in a clear and logical manner. A well-structured dissertation improves the research's legibility and comprehension while conveying its significance and coherence. Consequently, this dissertation is structured in the following order:

- Literature Review: Existing research and academic writings pertinent to the topic of the study are evaluated critically in the literature review. It summarizes prior research, identifies gaps in the literature, and provides a theoretical framework for the current investigation.

- Characterization of the Current State: This section describes the current process of refining, including production days, idle days, and maintenance days. It provides an exhaustive description of the process's parameters, with a particular emphasis on critical stages. By identifying the most important aspects of the process, this chapter highlights potential areas for improvement.

- Methodology: The methodology section describes the strategy employed to identify process constraints. Utilizing mass and energy balances to determine the overall process efficacy. The dissertation aims to identify process bottlenecks that hinder optimal performance through an analysis of the calculations derived from the conducted balances, both mass and energy related. In this section, you will also design solutions to eradicate these bottlenecks and optimize process parameters.

- Conclusion: The conclusion summarizes the key findings of the study, restates the research objectives, and offers a final evaluation of the study. It discusses the findings' implications and suggests their practical applications.

- Future Works: The "Future Work" section discusses prospective areas for future research or implementation based on the dissertation's findings.

Chapter 2

Literature Review

This chapter provides a succinct summary of the theoretical underpinnings that served as the project's fundamental foundation. Its purpose is to elucidate the most pertinent knowledge regarding the investigated subjects by providing a theoretical exposition of the dissertation's overall trajectory.

2.1 Total Productive Maintenance

Total productive maintenance (TPM) was created and first applied by the Japanese in 1971 as a solution to the maintenance and support issues that arise in manufacturing environments (Chan et al. (2005)). TPM is a maintenance approach that covers every aspect of equipment life, including planning, manufacturing, and maintenance. It explains how all organizational functions work together, but especially how production and maintenance do (Nakajima (1988)). It seeks to maximize the effectiveness of the equipment through a system that ensures its maintenance and all that is related to it, incorporating in this process all employees, through motivational management and satisfaction of them (Tsuchiya (1992)).

2.1.1 OEE- Overall Equipment Efficiency

As was previously stated, TPM aims to eliminate defects, waste, equipment breakdowns, and accidents. In this manner, Nakajima (1988) introduced the key performance indicator (KPI) overall equipment effectiveness (OEE); this metric was devised as part of total productive maintenance (TPM) to measure the equipment productivity in a manufacturing system. OEE is a ratio between actual manufacturing and what could be manufactured optimally. This indicator is widely acknowledged by many businesses, for instance, when implementing lean manufacturing or maintenance programs in order to monitor actual performance of an equipment (Sharma and Bhardwaj (2012)).

The OEE denotes the proportion of added value produced by a given piece of equipment. As such, it is a tool used to identify operational losses in order to enhance the overall performance and reliability of the equipment. It provides a basis for establishing improvement priorities and identifying root causes by categorizing significant losses or reasons for poor performance. It is the product of three factors: availability, throughput, and quality, and it indicates the extent to which the output meets specifications. As depicted in Figure 2.1, Nakajima and Chan et al. (2005) designated the six most significant losses in equipment efficiency as efficiency losses:

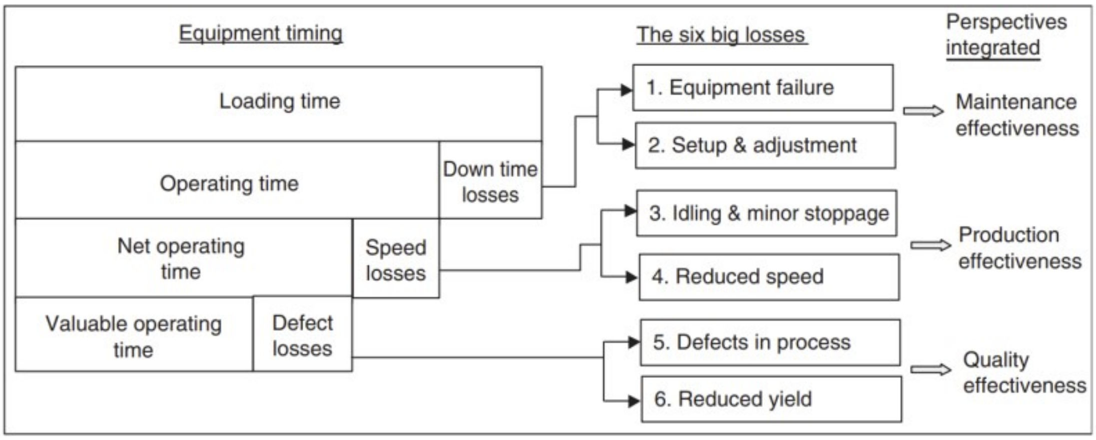


Figure 2.1: OEE measurement tool and the performance perspectives integrated into the tool (Adapt from Cheh (2014))

1. Breakdown losses/ Equipment failure: Due to the sporadic/chronic nature of these losses, they are accompanied by time losses (output decline) and volume losses (defect occurrence). The occurrence of sporadic failures in response to changes in certain conditions represents a restoration issue, necessitating the implementation of corrective measures to restore the condition to its original state. When machinery equipment contains concealed flaws, there will be recurring failures.

2. Set-up and adjustment losses: This refers to the amount of time lost between the conclusion of the production of the previous item and the point at which the production of the new item is wholly satisfactory. Set-up refers to the process of preparing a machine or equipment for a specific production task. It involves the installation of tools, configuration of settings, and adjustment of components, to accommodate a new process or product.

3. Minor/idling stoppage losses: Minor and idling stoppages occur when a transient malfunction interrupts production or when a machine is idling.

4. Reduced speed losses: These are losses that occur because the equipment speed is slow. They can be defined as losses due to a difference between the design speed and the actual speed or losses caused when the design speed is lower than present technological standards or the desirable condition.

5. Defect/rework losses: Defined as volume losses resulting from defects and rework, as well as the time required to rectify defective products and transform them into excellent ones.

6. Reduced yield: Defined as time and volume losses from the following: Start-up after periodic repair; Start-up after suspension; start-up after holidays; start-up after lunch breaks.

Equipment availability, performance efficiency, and quality rate can be expressed as follows based on the above loss measurements:

- Availability Ratio: the proportion between the actual and anticipated production times.

$$OEE = \frac{Loading\ Time - Downtime}{Loading\ Time} \tag{2.1}$$

- Performance Ratio: production losses caused by underutilized apparatus. In other words, losses occur when the apparatus is not operated at maximum speed.

$$OEE = \frac{Processed\ Amount}{Operating\ Time/Theoretical\ cycle\ time} \tag{2.2}$$

- Quality Ratio: the quantity of output that must be discarded or eliminated.

$$OEE = \frac{Processed\ Amount - Defect\ Amount}{Processed\ Amount} \tag{2.3}$$

When combining Equation 2.1, 2.2 and 2.3, the OEE for a given company operation is determined from:

$$OEE = Availability\ Ratio \cdot Performance\ Ratio \cdot Quality\ Ratio \tag{2.4}$$

$$OEE = \frac{(Processed\ Amount - Defect\ Amount) \cdot Theoretical\ Cycle\ Time}{Loading\ Time} \tag{2.5}$$

In conclusion, OEE is a fundamental measurement technique for performance measurement systems. For a deeper comprehension of machine utilization, a broader classification of losses is required. In addition, the levels of OEE measurement and the factors that influence them vary across industries and business sectors. Consequently, a customized OEE in various industries or business sectors is required. This KPI is a crucial metric for the analysis of a facility, especially in the process industry where productivity is highly dependent on the performance of the equipment.

2.2 Process Industry

It was deemed appropriate to conduct a framework of the characteristics that define the continuous production industry and the applicability of Lean tools. Beginning with the classification of the various forms of manufacturing, two categories stand out: the process industry and discrete manufacturing. Pérez-Muñoz et al. (2015) defines discrete manufacturing, as a type of production process that involves the assembly of individual components or elements into a final product, whereas in process manufacturing, the production process involves combining, blending, or chemical reactions to create a final product.

According to the article Pérez-Muñoz et al. (2015), the primary distinction between discrete production and process industry is the method of production. In discrete manufacturing, the product is generated in distinct phases, each of which produces a distinct component or part. In process manufacturing, basic materials are transformed into finished products through a series of chemical or physical processes, either in a continuous or discrete process. Within the scope of this thesis, the project concentrates on sugar refining; consequently, the production referred to in the text is a process industry.

2.3 Lean in Process Industry

2.3.1 Working standards

Standard work is a collection of work procedures that establish the optimal methods and sequences for each process and worker (Pereira et al. (2016)). As stated by Mr. Taiichi Ohno, "where there is no standard, there is no room for improvement," standard work seeks to minimize waste while optimizing the performance of each worker's duties and operations.

Variability in work processes increases the likelihood of disruptions (any deviation from the expected outcome), errors, and negative iteration, resulting in schedule and budget overruns. Standardizing work methods reduces the likelihood of breakdowns, which improves work flow, provides a foundation for learning from breakdowns that do occur, and enables the testing of alternative work method designs Feng and Ballard (2008).

As stated in Pereira et al. (2016), this is a device very common in cellular manufacturing and draw production to maintain the production rate aligned with the flow of customer orders and to make it easy for operators to switch positions within the process. Standard work consists of three essential elements:

1. Takt Time- Time available divided by customer demand represents the utmost time allowed per unit to meet customer demands.

2. Standard work sequence- Sequence of tasks or steps that must be followed to obtain consistent and efficient results in a specific process.

3. Standard work-in-progress inventory- Work in progress that is optimal for a particular procedure, minimizing waste and ensuring a smooth workflow.

In the process industry, however, the topics associated with this instrument significantly change. As stated previously, the process industry is significantly less reliant on human action and much more reliant on machinery and parameters. Therefore, in this domain, standardization work concentrates more on the optimization and normalization aspects that are directly influencing parameters. In addition, set-up and idleness are crucial factors for the standardization of work in this industry.

2.3.2 Line Blancing and Yamazumi

In modern manufacturing processes, it is essential for organizations to maximize production efficiency in order to remain competitive in a constantly changing market. A common method for achieving high productivity and output is an assembly line consisting of multiple workstations where duties are performed sequentially. To ensure maximum performance, companies must manage the production line effectively and distribute the workload evenly across multiple workstations. This is where line balancing, a systematic technique that eliminates bottlenecks and improves assembly line efficiency, plays a crucial role (Talapatra et al. (2018)).

Based on Coimbra et al. (2009), line balancing is a crucial instrument for decreasing production time, maximizing output, and minimizing expenses. It entails arranging the processing and assembly duties at each workstation so that the total time required at each workstation is roughly equivalent. This aids in the elimination or reduction of non-value-added activities, which are tasks that do not contribute to the value of the ultimate product. To effectively implement line balancing, it is necessary to have detailed information regarding the process flow and cycle time at each workstation. The objective of the traditional line balancing problem is to designate each operation to a workstation while minimizing the number of workstations and adhering to precedence constraints. Each workstation's cycle time should not exceed the specified takt time, which is the time required to manufacture one unit of the product. Idle time is the difference between the takt time and the cycle time. The objective is to reduce the balance delay time as well as the number of workstations.

According to Rathod et al. (2016), a Yamazumi chart is a visual tool used to depict the balance of cycle time workloads among operators in an assembly line or work cell (Figure 2.2). It is a stacked bar chart that shows the distribution of work content. The Yamazumi chart can be created for a single product or a multi-product assembly line. It helps visualize the work content of tasks, facilitate work balancing, and identify and eliminate non-value-added work content.

Figure 2.2: Example of a Yamazumi chart (Taken from mud (2015))

In conclusion, the advantages of line balancing in production and the utilization of the Yamazumi chart methodology contribute significantly to the enhancement of the efficiency and efficacy of assembly line operations. The bulk flow of products along the production line is optimized by achieving a balanced workload distribution, resulting in a smoother and more consistent workflow. This ensures that each workstation operates at its optimum capacity, minimizing downtime and maximizing productivity overall. In addition, the identification and removal of obstructions through line balancing contribute to the streamlining of the production process, resulting in decreased delays and increased throughput. With its graphical representation of workload balance, the Yamazumi chart facilitates the identification of imbalanced areas and enables targeted enhancements, leading to enhanced overall performance and optimal resource utilization. Consequently, the implementation of line balancing techniques and the incorporation of the Yamazumi chart provide organizations with valuable tools for optimizing production lines and achieving operational excellence. To explore the complete potential of these methodologies in a variety of industrial contexts, additional research and practical implementations are suggested.

2.3.3 Line Blancing and Yamazumi in Process Industry

Line balancing and Yamazumi, which are utilized extensively in discrete production, may not be directly pertinent to the process industry due to its inherent differences. In contrast to discrete production, where duties can be distributed and balanced to eliminate bottlenecks, this approach is impractical in the process industry. However, by drawing analogies and adapting methodologies, analogous principles can be applied to this context. Nevertheless, despite the fundamental distinctions between discrete production and the process industry, it is possible to draw analogies and employ a similar method of analysis. Organizations in the process industry can perform an identical analysis that integrates a mass balance by applying the same principles and concepts used in discrete production time analysis(Veverka and Madron (1997)).

Mass balance analysis is a fundamental principle and accounting technique employed in the process industry to track and manage the material flow within a system or process (Veverka and Madron (1997)). Quantifying and balancing the mass of inputs, outputs, and intermediate streams within a defined system boundary constitutes mass balance. It is mentioned that a mass balance is used to ensure that the mass of all materials entering a process is equal to the mass of all materials exiting the process, taking into consideration any transformations or conversions that may occur. A mass balance takes into account both mass and material properties conservation, such as chemical composition and physical state. In this article it is mentioned that engineers and operators can gain insight into the efficacy, performance, and reliability of a process by conducting a mass balance analysis. It enables them to identify any discrepancies, losses, or gains of materials, and to then take corrective actions to optimize operations, minimize waste, and ensure regulatory compliance. In industries where material flows must be closely monitored and controlled, such as chemical manufacturing, food processing, oil refining, and wastewater treatment, mass balances are extensively used.

In compliance with Veverka and Madron (1997), the mass balance plays a vital role in determining the distribution of workload across various phases. Similar to line balancing analysis and yamazumi creation in discrete production, this method seeks to identify and eradicate bottlenecks, resulting in an increase in production rates. This dissertation concentrates on performing a mass balancing analysis within the refining process in an attempt to quantify the flow rate and properties at each stage. It is possible to identify process bottlenecks and determine optimal parameters by comparing these values to the nominal maximum capacity of the equipment designated to each stage. This analysis will provide valuable insights for optimizing the refining process, maximizing production output, and enhancing efficiency.

2.4 Sugar Cane Refining Process

To ensure the effective implementation of mass balance, a comprehensive understanding of the stages under analysis is essential. This section provides a detailed examination of these stages, encompassing a thorough analysis of their components, input and output elements, as well as the pertinent formulas. The information presented in this sub-chapter will serve as a foundation for conducting the mass balance in Chapter 4. By delving into the intricacies of each stage and elucidating the relevant variables, this chapter sets the stage for a robust and accurate mass balance assessment. The insights gained here will pave the way for a comprehensive analysis in the subsequent chapter, enabling a meticulous evaluation of the system's performance and identification of areas for improvement.

2.4.1 Sugar Cane

Raw sugar consists of sugar crystals encased in a thin layer of liquid, which plays a crucial role in facilitating the bulk transport of sugarcane across diverse climatic zones without compromising its physical and chemical properties. This protective film functions as a shield, preserving the sugar crystals during transport. It prevents changes such as moisture loss, chemical reactions, and contamination, preserving the unprocessed sugar's integrity until it reaches its final destination for further processing.

According to Rein (2007), several important characteristics are evaluated to determine the quality and refinement suitability of raw sugar. Moisture content is an important parameter because it indicates the amount of water present in the unprocessed sugar. By subtracting the moisture percentage from 100, the brix value can be determined. Brix is a crucial indicator of the sugar concentration in a solution, allowing for precise determinations of the sugar content, and as stated by Chou (2000) it can be obtained from Equation 2.6:

$$Brix = \frac{Dry\,Matter}{Total\,Mass} \tag{2.6}$$

Color is another essential characteristic of unrefined sugar that is evaluated. It functions as a visual indicator of the quality, purity, and potential impurities of the sugar. The color can vary depending on the refining process, storage conditions, and presence of non-sugar components. Polarization is an essential metric for determining the sucrose content of unprocessed sugar. Typically ranging between 0.97 and 0.99, the pol/sucrose ratio reveals the amount of sucrose present in the solution. This parameter aides in the evaluation of sugar content and serves as a guide for further processing and refining. Dextran, a non sugar component, is crucial in the extraction and refining of cane juice. While other non sugar components are typically accounted for when measuring the purity of juice, Rein (2007) defends that dextran must be evaluated separately due to its significant influence on sugar loss in molasses and the processing of cane juice. High concentrations of dextran can increase the viscosity of massecuites, which hinders the sugar recovery process during pan simmering. In extreme cases, sugar recovery is impossible, whereas in less severe cases it considerably slows down the pan-boiling process and contributes to sugar loss in molasses. Additionally, the unprocessed sugar's starch and total soluble solids (TSS) content play a significant role, particularly in the carbonation phase of the refining process.

2.4.2 Centrifugation

As stated in Chou (2000), the purpose of the affination station is to separate the molasses film encircling the raw sugar crystal while minimizing crystal dissolution. The quality of the sugar has a significant effect on the affination procedure, with large, uniform crystals being easier to cleanse and purge. In contrast, microscopic crystals and a high percentage of fines can present difficulties during affination. Consequently, the proportion of fines in raw sugar is frequently specified or penalized in raw sugar supply contracts. The centrifugation efficacy at this stage is approximately 0.95, indicating that approximately 5% of the raw sugar cane's dry matter, or molasses, is extracted(Rein (2007)).

2.4.3 Carbonatation and Filtration

Carbonatation is a sugar refining procedure utilized to purify and refine sugar liquor. It involves the precipitation of calcium carbonate by adding lime and gassing with a carbon dioxide-containing vapor. By incorporating impurities into the crystals, the resulting crystalline mass serves as a filter assist in the pressure filtration process. According to Rein (2007), the optimal lime dose ranges from 0.40% CaO on solids in some liquors to over 1.20% CaO in others.

Based on Chou (2000), the filtration process of raw sugar is influenced by two primary factors: insoluble suspended matter (referred to as TSS) and starch. The presence of suspended solid materials mechanically impacts the filterability of the sugar solution, necessitating their removal through the addition of filter aid or the carbonatation process involving precipitated calcium carbonate. Higher levels of suspended solids require a larger quantity of filter aid or an increased lime dosage rate during carbonatation to improve filtration.

Starch exerts a significant influence on the filterability of raw sugar. Specifically, the amylose fraction of starch acts as a protective colloid, forming a coating on the surface of growing sugar crystals. This coating impedes the formation of agglomerates and reduces interparticle repulsion charge, resulting in a tightly packed cake with low filterability. Consequently, the presence of starch in the sugar solution negatively affects the filtration efficiency.

2.4.4 Ion-Exchange Decolorization

Ion-exchange resins are granules or beads made of polymers containing functional groups that can effectively bind ions possessing charges of opposite polarity (Hassan and Carr (2018)). These resins are classified into two types: cation-exchange and anion-exchange resins. However, for the purpose of removing reactive dyes from effluent, cation-exchange resins are not suitable due to their similar charge properties with the dyes. Pure sucrose is colorless, but may appear colored because of the inclusion of small amounts of colored material in the sugar. Color is the generic term used to cover a wide range of components which contribute to the color of sugar. Most of these compounds are complex and not easy to quantify and so color is measured as the total effect of all colorants on light absorbance (Rein (2007)). Depending on the color loading, the resin has a cycle life of between 200 and 400. The system's color load, according to Rein (2007) can be calculated as tonnes of dissolved solids in the feed liquor multiplied by color and then divided by the resin volume.

2.4.5 Double Effect Evaporator

A double-effect evaporator is a multi-effect evaporator configuration used to concentrate liquid solutions by removing water in numerous industrial sectors, including food processing, chemical production, and pharmaceutical manufacturing. Its energy efficacy is superior to that of single-effect evaporators, making it a popular option for many applications. According to Rein (2007), the operating principle of a double-effect evaporator is to use the vapor produced by one effect to fuel the evaporation process of the other effect. This configuration consists of two interconnected evaporator vessels, also known as effects. The first effect operates at a higher temperature and pressure than the second. Despite its importance in numerous industrial processes, the evaporator is notorious for its high energy consumption. When this equipment becomes an energy bottleneck, it impacts not only energy efficiency but also the system's mass equilibrium. Therefore, Rein (2007) recommends for a efficient analysis to conduct both a mass balance and an energy balance to assure a comprehensive evaluation of its performance. This requires the application of specific formulas designed to quantify the energy transfer within the system, allowing for a comprehensive analysis of the evaporator's energy consumption and its effect on the overall process efficiency. Figure 2.3 shows a schematic of a multi-effect evaporator:

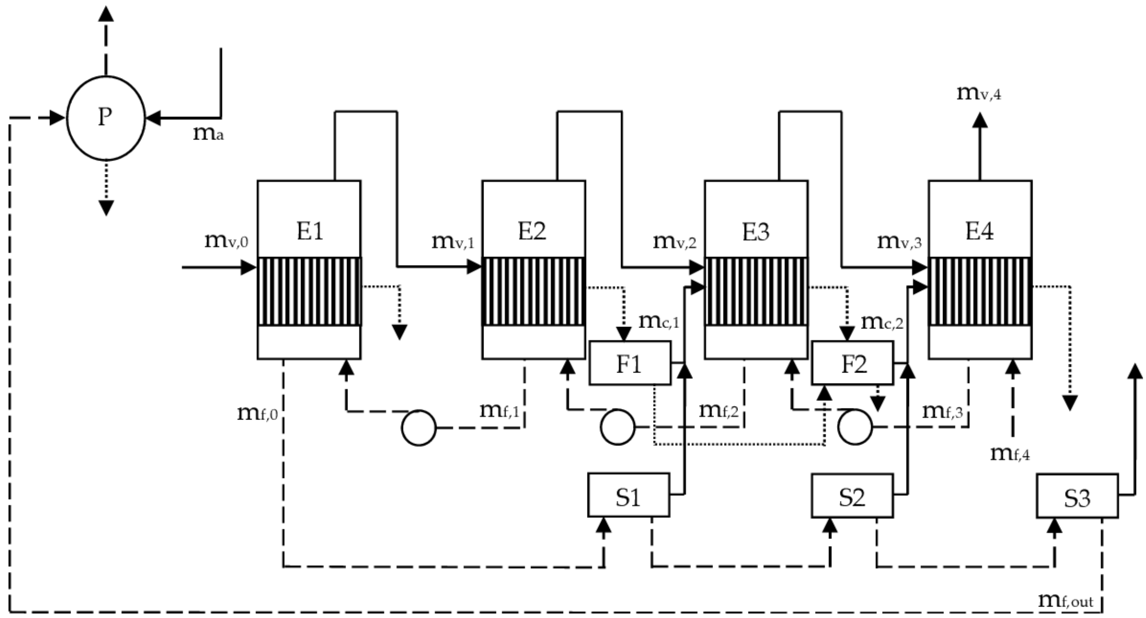


Figure 2.3: Mass diagram of a multieffect evaporator (Adapt from Chantasiriwan (2020))

The mass balance of the evaporator can be determined using Formula 2.7. According to Formula 2.8, the formula is marginally modified for the energy balance (designation of each label in the annex D).

$$\dot{m}_{v,0} + \dot{m}_{f,0} = \dot{m}_{v,n} + \dot{m}_{f,n} + \sum \dot{m}_{c,i} + \sum \dot{m}_{vb,i} \tag{2.7}$$

$\dot{m}_{v,0}$= Steam Input Mass in the Evaporator
$\dot{m}_{f,0}$= Liquor Input Mass in the Evaporator
$\dot{m}_{v,n}$= Steam Input Mass in the Effect n
$\dot{m}_{f,n}$= Liquor Outlet Mass in the n effect of the Evaporator
$\dot{m}_{c,i}$= Condensate Outlet Mass in the Effect n
$\dot{m}_{vb,i}$= Vapour Outlet Mass in the i effect of the Evaporator

$$\dot{m}_{v,0} \cdot h_{v,0} + \dot{m}_{f,0} \cdot h_{f,0} = \dot{m}_{v,n} \cdot h_{v,n} + \dot{m}_{f,n} \cdot h_{f,n} + \sum \dot{m}_{c,i} \cdot h_{c,i} + \sum \dot{m}_{vb,i} \cdot h_{v,i} \tag{2.8}$$

$h_{v,0}$= Enthalpy of the Steam Input in the Evaporator
$h_{f,0}$= Enthalpy of the Liquor Input in the Evaporator
$h_{v,n}$= Enthalpy of the Steam Input in the Effect n
$h_{f,n}$= Enthalpy of the Liquor Outlet in the n effect of the Evaporator
$h_{c,i}$= Enthalpy of the Steam Input in the Effect n
$h_{vb,i}$= Enthalpy of the Vapour Outlet in the i effect of the Evaporator

By applying Formula 2.9 and having an understanding of the input and output brix and the input mass flow rate, it is possible to calculate the amount of water evaporated:

$$\dot{m}_e = \dot{m}_{l,0} \cdot (1 - \frac{Brix\ Inlet}{Brix\ Outlet}) \tag{2.9}$$

The Formula 2.10 is used to calculate sensible heat, which is the heat responsible for the temperature rise to saturation and the Formula 2.11 is used to calculate latent heat, which is the heat required to evaporate a solution.

$$\dot{Q} = \dot{m} \cdot c_p \cdot \Delta T \tag{2.10}$$

c_p= Specific Heat of the Solution

$$\dot{Q} = \dot{m}_{e,1} \cdot \Delta h_{lv} \tag{2.11}$$

$\dot{m}_{e,1}$= Total Mass of Evaporated Water in 1 effect
Δh_{lv}= Enthalpy of Vaporisation

According to LYLE (1957), in a double effect evaporation there are three stages of evaporation:

First Effect-heat is supplied to the liquid, causing a portion of it to evaporate and form vapor. This vapor carries away the latent heat of vaporization and is directed to the second effect. Meanwhile, the remaining liquid, now more concentrated, flows to the second effect. The first effect's evaporation rate is given by Equation 2.10.

Flash-refers to the sudden release or liberation of vapor when a liquid is subjected to a lower pressure environment. In this particular context, the flash occurs when the concentrated liquid from the first effect of the evaporator enters the second effect, where the pressure is lower. As the liquid passes through the second effect, it experiences a drop in pressure, causing some of the liquid to rapidly vaporize or "flash off." This vaporization helps in further concentrating the liquid while generating vapor that can be used to provide heat for subsequent stages of the evaporation process or for other purposes within the industrial operation. It is feasible to compute flashing effect's rate of evaporation with Equation 2.12 (LYLE (1957)):

$$\dot{m}_{e,f} = \dot{m}_{v,2} \cdot (T1 - T2) \cdot Self\ Evaporation\ Rate \tag{2.12}$$

The self evaporation rate is given by Table D.2 in Appendix D.

Second Effect-the concentrated liquid from the first effect is further heated and evaporated using the vapor generated in the previous stage. This generates a higher concentration of the desired product in the liquid, as more water or solvent is removed. The vapor produced in the second effect may be sent to subsequent effects in a multiple-effect system or utilized for other purposes. The second effect's evaporation rate is given by:

$$\dot{m}_{e,2} = \dot{m}_e - \dot{m}_{e,1} - \dot{m}_{e,f} \tag{2.13}$$

2.4.6 Crystallisation

Crystallization is initiated by the introduction of finely dispersed seed particles conveyed as a slurry. These seeds serve as nuclei, establishing the basis for crystallization to begin. To guarantee successful crystallization, the concentration of the mother liquor is carefully regulated to prevent the dissolution of existing crystals and the formation of false grain, also known as new nuclei. This necessitates the establishment of a sufficient crystal surface area and precise control over the input to the crystallization vessel, which enables effective manipulation of the mother liquor concentration.

Traditionally, this procedure is performed in vacuum-sealed batch containers. Using a vacuum to sustain a lower temperature minimizes the formation of color and prevents the inversion or decomposition of sucrose during the process. The combination of vacuum, mass transfer, and evaporation processes allows for the desired crystallization effect.

The cycle of vacuum pans is a seven-stage process. It begins with liquor feeding and pan loading, followed by concentrating the solution and adding seed material to increase the liquor brix. The third stage involves another liquor feeding phase. The fourth stage focuses on concentration up to the syrup inlet, resulting in an increased liquor brix. The fifth stage is the syrup inlet phase, while the sixth stage involves crystal squeezing. The final stage includes discharge and pan washing. These stages ensure efficient sugar production with proper crystallization and quality control (LYLE (1957)).

2.5 Chapter Overview

The literature review presented is able to reach conclusions regarding the implementation of a Kaizen strategy in the process industry based on the literature review presented. To achieve operational excellence, the fundamental principles of waste elimination and the improvement of material and information flow must be consistently considered. However, due to the distinct characteristics of the process industry, applying these concepts presents a unique challenge. While the objective of obtaining benefits remains constant across contexts, the approach to processes and use of performance indicators vary. Traditional methods such as line balancing and creation cannot be applied directly in this context, necessitating a translation of these concepts into the process industry's reality. This requires a concentration on mass equilibrium and, if necessary, energy balance. As a result, it is essential to focus on equipment performance, ensuring that operational parameters are maintained within the desired ranges to facilitate continuous process improvement. The following chapter provides a detailed description of the problem that inspired this thesis, which serves as the focal point for the methodologies discussed thus far.

Chapter 3

Characterisation of the Present State

To enhance comprehension of the proposed challenge, it is essential to depict and analyze the organization's initial state at the onset of the project. This chapter conducts a comprehensive analysis of the current state, focusing on the identification of areas ripe for improvement. It focuses on the most vulnerable process areas within the organization and seeks to identify potential bottlenecks that could impede progress. By examining these facets, a comprehensive picture of the organization's current state is created, allowing for the exploration of opportunities to increase the project's efficiency and efficacy in pursuit of its objectives.

According to the aforementioned methodology, the process is first described from the time the raw material is introduced into the transport section until it reaches the desired final state, white or brown sugar, and is ready for shipment. Problems and obstacles encountered in the current situation that will hinder the achievement of the objectives are then highlighted

3.1 Evolution of the refining rate and working days in the last years

In terms of data extraction and processing, Company X faces a multitude of obstacles. The absence of dependable measuring instruments and sensors able to provide accurate values for process parameters has been a major setback. Moreover, the data extraction process has been laborious due to a lack of automation to retrieve data autonomously and a reliance on manual labor. In order to obtain reliable data that could support a comprehensive analysis of extant issues, it was essential to overcome this obstacle. Collaboration between the project team and the IT department of Company X was crucial to achieving this goal. They engaged the relevant database-holding companies in a concerted effort to extract all obtainable data. In order to consolidate and organize the acquired data, a computerized record was subsequently created. This exhaustive data accumulation served as the foundation for the conducted analysis.

The term "stationary" working rhythm signifies a state of continuous operation within a factory setting, characterized by an absence of interruptions or breaks. This particular working rhythm indicates a consistent and unbroken workflow, wherein the production process maintains a constant pace without frequent instances of pauses or downtime. There was a notable 11% rise in free time

Table 3.1: Refining Rate

	2020		2021		2022	
White Sugar ($\cdot 10^3$ ton)	134.00	(21.80ton/h)	134.00	(21.50ton/h)	138.00	(20.30ton/h)
Brown Sugar ($\cdot 10^3$ ton)	7.30	(1.17ton/h)	6.70	(0.97ton/h)	6.20	(0.84ton/h)
Shutdown or Startup (Days)	27.30	(8.00%)	25.30	(7.00%)	24.00	(7.00%)
Stationary Work (Days)	217.00	(60.00%)	223.00	(61.00%)	247.70	(68.00%)
Cleaning (Days)	13.70	(4.00%)	12.30	(3.00%)	12.00	(3.00%)
Refinery Stopped (Days)	108.00	(30.00%)	105.30	(29.00%)	81.30	(22.00%)

between 2021 and 2022, resulting in an additional 25 days of stationary production. This increase is expected to contribute to a refined white sugar production increase of 13,000 tons (+9.60%). However, despite the extended duration, the absolute volume of sugar only witnessed a modest growth of 4,200 tons (+3.10%) due to a slower production rate of -1.20 ton/h. In addition to examining refining patterns and the working period, an analysis of cleansing and maintenance days was conducted. Notably, there was a significant 21% reduction in maintenance time observed between 2018 and 2022, signifying a notable improvement within the refining period.

As seen in the preceding chapter, the OEE is an indicator that measures the efficiency of a production operation, and it is calculated as represented:

$$OEE = \frac{Annual\ Production}{Factory\ Maximum\ Output} \tag{3.1}$$

In order to attain maximum output, it is imperative to consider the optimal cycle time of vacuum pans. With a prescribed cycle time of 110 minutes, equating to 13 boilings periods per day, and operating under ideal conditions encompassing optimal brix levels, steam pressure, and a lack of breakdowns, the pinnacle of output can be reached. Each boiling cycle generates 18 tons of production. Consequently, the calculation for attaining maximum output is obtained by multiplying 18 tons per boiling cycle by 13 boiling cycles per day, and further multiplying this by 3 (representing the number of vacuum pan sets). The resulting figure amounts to 702 ton/day, or approximately 29.25 ton/h. This approach guarantees a seamless workflow and efficient utilization of the vacuum pans, thereby facilitating optimal levels of production. So, this KPI was calculated for white sugar performance in order to identify the fundamental causes of the decrease in production rhythm. This key performance indicator is subdivided into availability losses, speed losses, and quality losses, allowing for the identification of the most detrimental losses to the manufacturing process's efficacy. For the calculation of this indicator, resorted to the nominal maximum of the section designated as the bottleneck of the process, the filtration.

In Figure 3.1, this indicator is represented graphically:

The diagram of Figure 3.1 illustrates the temporal losses attributable to various components within the framework of Overall Equipment Effectiveness (OEE). The proportions of time lost in each specific OEE category are represented by the red, orange and gray graph segments. The red segment represents availability loss, which is calculated by subtracting the availability loss

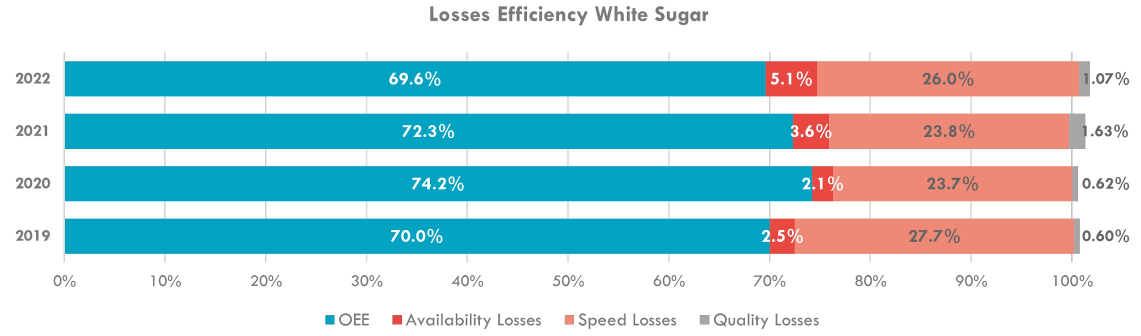


Figure 3.1: Graphical Representation of Efficiency Losses in the Production of White Sugar

percentage from 100%. In a similar fashion, the orange segment represents speed losses, which are calculated by subtracting the percentage of speed losses from 100%. The grey segment represents quality losses, which are calculated by subtracting the quality loss percentage from 100%. This graphical representation provides a concise overview of the impact of availability, speed, and quality factors on the overall effectiveness of the equipment, thereby providing valuable insights for development initiatives in these areas.

The indicator displayed, OEE, endures a -3.90% decline from 2021 to 2022, which explains the -1.20k ton/h decrease in output observed in 2022 despite the increase in opening time. The decline described is predominantly attributable to two factors: declines in speed and availability. Due to a reduction in the refining process's throughput, speed losses increased by 9%. Availability losses, on the other hand, increased by 42% in 2022, primarily as a result of frequent shutdowns of specific refinery equipment. These two factors have contributed significantly to the observed decline.

The emphasized production cadence can be divided into two regimes: stationary and stop-start, with the latter lasting one shift (eight hours) for white sugar and nearly a day (24 hours) for brown sugar. Thus, 87% of the refinery's working time in 2022 was spent in stationary mode, with the remaining time spent on start-ups, shutdowns, and cleansing. During the period under review, the steady-state refining rate was 21,75 ton/h, which was less than 26% of the utmost rate. During startup, shutdown, and cleansing, the production rate was only 8.21 ton/h. In comparison to the previous year, steady-state production decreased by nearly 6.60%, while stoppages, startup, and cleansing decreased by 8.90%.

Taking into consideration the two aforementioned production regimes, the total production rate (20,30 ton/h) is only 69% of the maximum possible rate (29,30 ton/h). As depicted in Figure 3.2, it is feasible to identify the causes associated with this 8.95 ton/h difference. The losses of rhythm are caused predominantly by speed losses that occur during "normal" refining periods, resulting in a decrease of 5.80 ton/h. In addition, interruptions, restarts, and cleansing operations at the refinery account for 20% of the total rhythm losses, or 1.72 ton/h. These factors have a significant impact on the refinement process's overall cadence and efficacy.

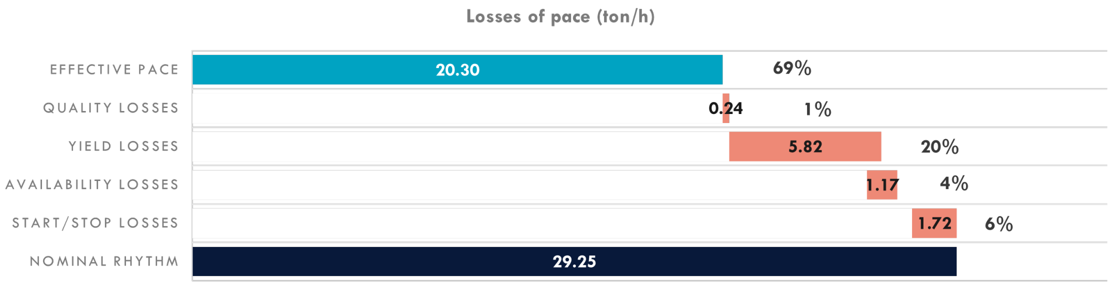


Figure 3.2: Graphical Representation of Pace Losses

Due to the absence of additional information and in order to facilitate this analysis, the following assumptions were made:

- Organic, extra-white and extra-coarse production shifts have been purged
- The availability losses are accounted for by the downtime of the refining centrifuges
- All micro-stops that do not cause the refining centrifuges to stop are counted as loss of speed

By making these assumptions, it can be concluded that the majority of pace losses, approximately 20% of the total pace losses, are attributable to speed losses caused by steadily increasing equipment malfunction. Starting and halting results in a loss of 1,72 ton/h of total speed, which is consider a significant factor.

3.1.1 Conclusions

In conclusion, the analysis reveals that despite an 11% increase in open time, the white sugar refining volume only experienced a modest 3.10% growth. This indicates a disparity between the extended operational period and the actual output achieved. The primary cause of this issue can be attributed to the worsening refining rhythm during working periods, which suffered from a notable increase in refining speed losses (-3.90 percentage points OEE) in 2022. Furthermore, a significant portion of the overall refining efficiency losses (8.95 ton/h) can be directly attributed to rhythm losses resulting from start-ups and shut-downs, accounting for 20% (-1.72 ton/h) of the total losses. It is worth noting that the number of start-ups and shut-downs remained consistent in recent years, implying a recurring problem. Addressing these issues and implementing measures to improve refining speed losses, reduce rhythm losses, and optimize start-up and shut-down procedures will be crucial for enhancing the overall refining efficiency and achieving substantial growth in white sugar refining volume.

3.2 Flow mapping

Before addressing the current characterization of the machinery and process parameters, a very brief material flow map was developed to contextualize and facilitate the subsequent analysis.

Sugar Cane consists of sugar crystals surrounded by a layer of nectar. The first step in the sugar refining process is the separation of the fluid film from the sugar crystal, which is known as affination. Here, the sugar cane and syrup are combined to create the affination magma. This magma is centrifuged in order to separate the affination sugar from the fluid, which is then recirculated into the magma. To this syrup is added pre-recovery syrup. The excess affination syrup is sent to the recovery, where it is recrystallized and the sugar it contains is recovered, leaving behind molasses.

The affination liquor undergoes a preliminary purification during the carbonation procedure. Carbonation involves the addition of calcium hydroxide to the liquor, followed by the precipitation of calcium carbonate through the bubbling of carbon dioxide. With this precipitation, insoluble substances and high molecular weight compounds that enter the precipitate and are separated in the subsequent filtration are separated simultaneously, after which it is circulated for decolorization. Decolorization is the process of eliminating pigmented compounds from the liquid. Resins accomplish decolorization by exchanging the colored ion for a decolored ion (chloride ion).

After decolorization, the liquor is partially evaporated in order to concentrate it. This concentration is achieved using evaporators with a dual effect. In vacuum pans, the liquor that emerges from the evaporators is crystallized. In crystallization, the water present in the liquor is evaporated and the sugar is concentrated to the point where sucrose crystals can grow. To aid this crystal formation, a small amount of sugar powder, the "seed", is added to the concentrated liquor in the Vacuum Pan. In the final part of the crystallization operation (boiling) syrup from previous crystallizations is added. At the end of the boiling process a mass is obtained, the boiled mass.

The crystals are separated from the surrounding syrup, the Refining Syrup, through centrifugation. In the drying sector the crystals are then dried in the sugar dryer, cooled and separated in vibrating classifiers, obtaining the different types of sugar. The classified sugar is then sent to the packaging sector.

The process of production for yellow sanded sugar is identical to that of white sugar until the product enters the vacuum pans. This form of sugar is refined using a mixture of liquors and syrups, which is first crystallized at a high temperature into a mass containing small crystals. This material is discharged into low-pressure stirrers called areadores, where evaporation and a second crystallization occur. Here, the cooked mass is transformed into sanded sugar, which is subsequently chilled, sieved, and packaged (representative scheme of the process in Figure A.1 of Appendix A).

3.3 Filtration

Given its significance in the sugar production process as a whole, filtration is the primary focus of this investigation. As previously discussed in Chapter 2, the efficacy and efficiency of filtration are highly dependent on the specific characteristics of the consumed raw sugar. Notably, this study recognizes the possibility of production limitations arising during the filtration stage. In this chapter, an initial step is taken to characterize the current state of filtration capacity, followed by a thorough statistical examination of the influence of unprocessed sugar properties on filtration performance and its subsequent effects. This method seeks to promote a comprehensive understanding of the complex relationship between filtration efficiency, raw sugar characteristics, and their effects on production outcomes. By obtaining knowledge from this study, informed decisions and potential improvements to the filtration process can be pursued to optimize sugar production operations as a whole.

To conduct the exercise effectively, the characteristics of the sugar cane consumed in the preceding two years were carefully examined, resulting in the derivation of an average value (characteristic of sugar cane consumed Table F.1 Appendix F). For comparative purposes, the characteristics of the sugar cane that closely resembled the calculated average, specifically the Peligiani sugar cane, were considered (characteristics shown in Table 3.2).

Table 3.2: Peligiani Sugar Cane

Sugar Cane	**Starch**	**TSS**	**Dextrane**
Peligiani	234.75	12380	93

The second phase involved examining the filtration records for the time period during which this raw sugar was consumed to determine the filtration rate and cycle duration's. It was determined the parameters shown in Table 3.3.

Table 3.3: Filtration Cycle Parameters

Flow at Variable Pressure	15.60 m^3/h
Time at Variable Pressure	6.00 h
Flow Loss at Constant Pressure	2.50 $(m^3/h)/h$
Filtration Stopping Flow	12.00 m^3/h
Setup Time	0.67 h

A filtration cycle consists of two stages namely variable pressure filtered flow and constant pressure filtered flow. The flow filtered at a variable pressure is calculated using the following Formula 3.2 and the flow at a constant pressure with Formula 3.3.

$$Total\ Flow = Flow\ Variable\ Pressure * Time\ Variable\ Pressure = 15.6 * 6 = 93.6\mathrm{m}^3 \quad (3.2)$$

$$Total\ Flow = \frac{F.\ Stopping\ Flow * Flow\ Variable\ Pressure}{2} * \frac{Flow\ Variable\ Pressure - F.\ Stopping\ Flow}{Flow\ Loss\ Constante\ Pressure}$$
$$Total\ Flow = \frac{12 * 15.6}{2} * \frac{15.6 - 12}{2.5} = 19.9\mathrm{m}^3 \quad (3.3)$$

The filtration capacity for a sugar cane, with characteristics based on the average of the last two years, consumption level is established at 113.50 m3 per filtration cycle. Each cycle entails a duration of 8.11 hours (Time at variable pressure+Setup Time+Time at constant pressure= 6 + 0.67 + $\frac{15.60-12}{2.50}$ = 8.11 h), resulting in an average filtration rate of approximately 14 m^3/h. The facility is equipped with two Putsch filters, each contributing with the filtration capacity calculated, 14m^3/h. Aditionally, there is also the presence of other filtration equipment, a rotating filter, being an antiquated apparatus, poses challenges in accurately determining its filtration capacity. Due to the absence of original components and the unavailability of a corresponding manual, conventional calculations cannot be employed. Moreover, the lack of instrumentation further hampers the characterization of filtration cycles. Consequently, historical data, limited to total flow per hour, was utilized as an alternative approach. Through examination, it was ascertained that the filter typically operates at a filtration rhythm equivalent to half of the filtered flow observed in the Putsch cycles. This reliance on historical data offers an imperfect yet practical means of estimating the filtration capacity of the rotating filter under consideration. Therefore, the rotating filter has capacity of around 7 m^3/h for this specific sugar cane. Consequently, the combined filtration capacity of the two Putsch filters totals 28 m^3/h, while the rotating filter adds an additional 7 m^3/h. By summing the respective capacities, the overall filtration capacity of the plant is determined to be 14 m^3/h (Putsch filter) + 14 m^3/h (Putsch filter) + 7 m^3/h (rotating filter), resulting in a total filtration capacity of 35 m^3/h.

In conclusion, the preceding discussion relates to the consumption of sugar cane with average characteristics observed in prior years. However, it is essential to recognize that an increase in the raw material's starch content or Total Suspended Solids (TSS) has a deleterious impact on its filtration capacity. The graphic presented in Figure 3.3 indicates a upward trend in starch concentrations over the past few years.

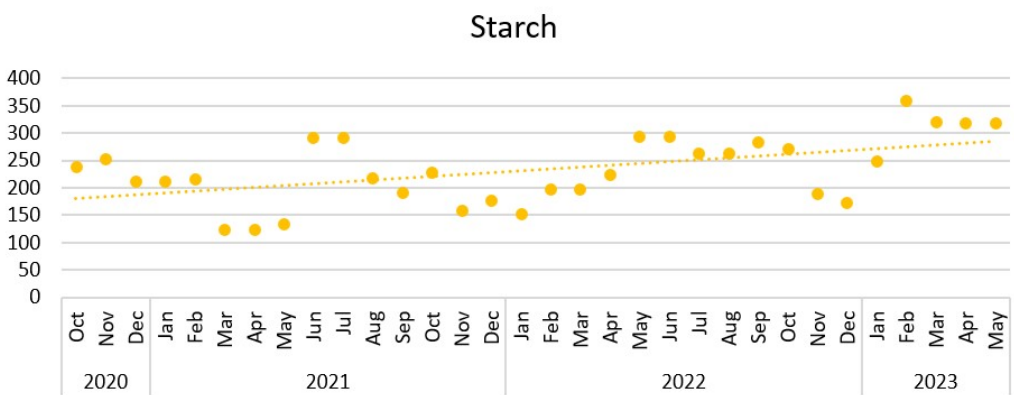


Figure 3.3: Graphic Representation of Starch Concentration in Consumed Sugarcane

The exercise was conducted for a starch percentage of approximately 234 ppm, whereas the current value (currently 300-350 ppm) is much higher. This will result in an even lower filtration capacity, as the graph demonstrates. Due to the use of increasingly inferior sugarcane quality, the filtration capacity is compromised, rendering it inadequate to satisfy the desired refining rates. This event generates a bottleneck in the filtration process, reducing its effectiveness (theme addressed in chapter 4).

In addition to the previously mentioned problem of deteriorating sugar cane quality, a reduction in maintenance time has emerged as a factor in the compromised filtration capacity. Due to the limited amount of time allocated for equipment maintenance, equipment functionality has degraded over time, resulting in a decline in filtration efficacy. The inadequacy of maintenance interventions has hindered the timely detection and resolution of equipment problems, thereby exacerbating the detrimental effect on filtration performance. As a result, the filtration capacity has suffered as a result of both the deteriorating quality of the sugar cane feedstock and the lack of attention paid to essential equipment maintenance. Analyzing the filtration rate in the last three years, a reduction of the rate is observed, as represented in Figure 3.4:

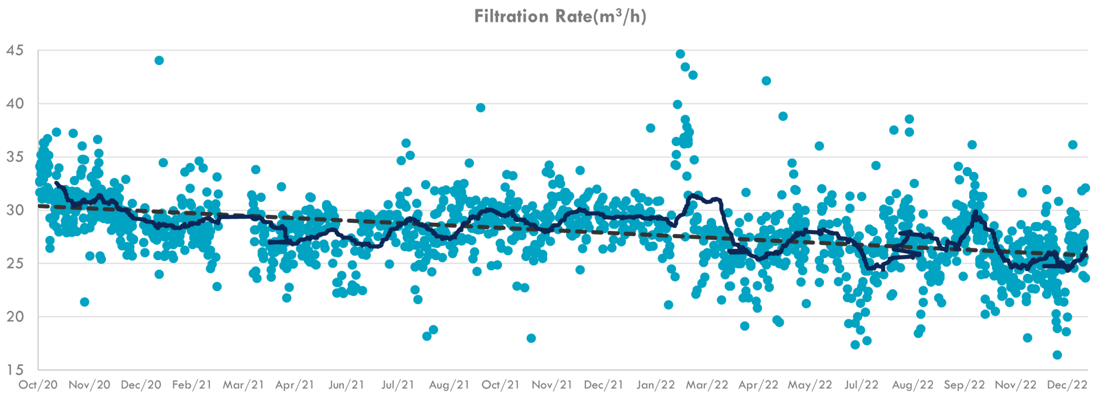


Figure 3.4: Filtration Rate Representation Graphically

In the past three years, the filtration rate has decreased by 13%, from 30.70 m^3/h to 26.90 m^3/h. This decrease is attributable to the approximately 48% increase in the number of malfunctions in this section. In addition, the filtration time of Putsch filters has fluctuated significantly over the past few years, indicating that a change in the method of equipment operation, the sugar cane consumed and the condition of the equipment may be influencing the speed and performance of these machines.

In order to comprehend the impact of sugar cane characteristics, a statistical study known as linear regression was conducted for each characteristic. Linear regression is a statistical modeling technique applied to determine the relationship between a dependent variable and one or more independent variables. In order to forecast the value of the dependent variable based on the values of the independent variables, it seeks to identify the linear equation that best fits the observed data points. In this case, the dependent variable is the filtration rate and the independent variables are amount of starch, TSS and Dextrane in the sugar cane. Starting with the starch, which appears to be the most important factor in filtration, the graph presented in Figure 3.5 shows the results of the analysis:

According to Figure 3.5, about 35% of the variation in the dependent variable (filtration rate) can be attributed to the amount of starch in the crude. This demonstrates a moderate correlation between the independent and dependent variables.

The filtration rate is the dependent variable, and the amount of suspended particles in the sugar cane explains around 11% of the variation (Figure 3.6). This demonstrates a poor connection between the independent and dependent variables. Opposite to the other features, the investigation

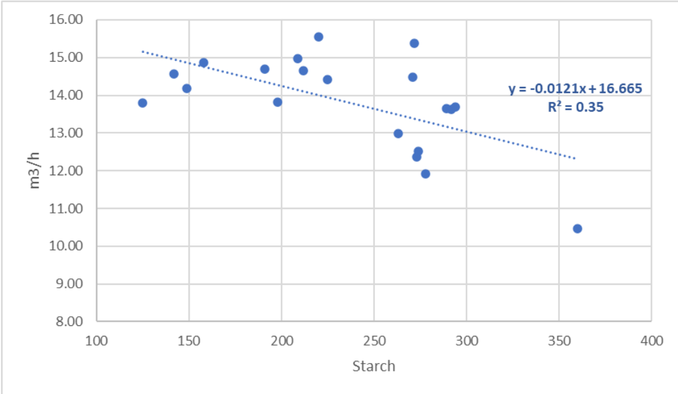


Figure 3.5: Starch- Linear Regression

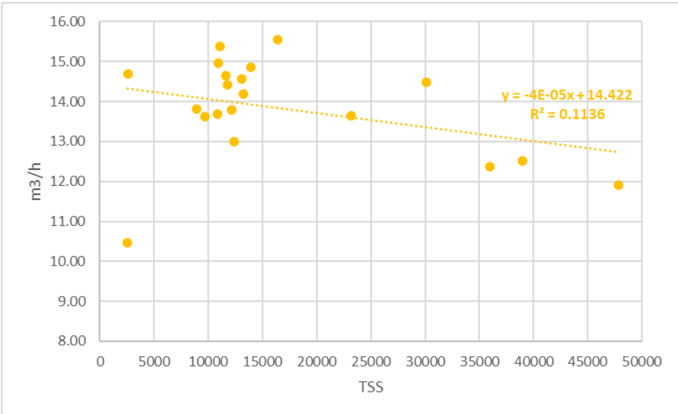


Figure 3.6: TSS- Linear Regression

shows (Figure 3.7) that there is no direct relationship between the filtration rime and the amount of Dextran in the sugar cane.

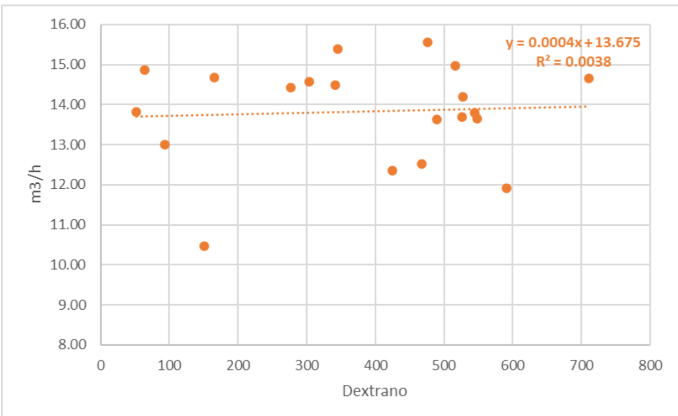


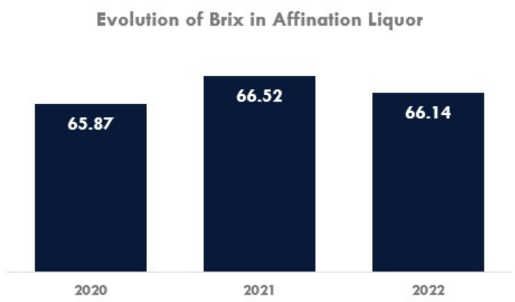
Figure 3.7: Dextrano- Linear Regression

In order to understand the influence of these characteristics together, a multiple linear regression was performed. Multilinear regression, also known as multiple linear regression, is an extension of simple linear regression that allows for the analysis of the relationship between a dependent variable and two or more independent variables. It aims to establish a linear equation that best explains the variation in the dependent variable based on the values of multiple predictors. Therefore, A value of R^2= 41.78% was obtained after completing a multiple linear regression, showing that the amount of starch, Tss, and Dextran contributes 41.78% of the variation in the response variable- Filtration rate.

3.4 Process Parameterization

Brix is one of the most crucial parameters in the primary sugar refining process. Brix is a unit of measurement that quantifies the amount of soluble substances present in a liquid and is used as a measure of concentration. In the case of sugar liquor, Brix is used as a measure of the sucrose concentration present in the solution.

The reduction of a 1º Brix at the entrance of the vacuum pots translates into an increase of the cycle time in average 8 minutes, less 2 ton/h. Thus, there is a decrease (represented in Figure 3.8) of the Brix of the concentrated liquor, the designation used for the solution at the entrance of the pots, in 2022 relative to 2021 of 0.76º, which translates into a decrease in the refining rate of 1.40 ton/h.



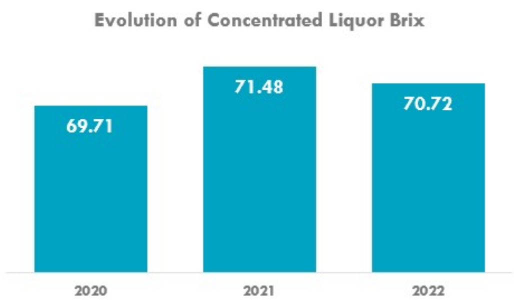


Figure 3.8: Graphic Representation of Affination Brix and Concentrated Brix

Through the graphs in Figure 3.8, it is also concluded that the reduction of the concentrated Brix in 2022 (-0.76º) was higher than the reduction of the refining brix (-0.40º), revealing a higher water input in the process or a lower evaporation capacity, resulting from the decrease of steam pressure in the plant verified in recent years. The increase of water in the process aims to combat the unfavorable characteristics of the raw material verified in recent years. The addition of water causes the dilution of the solution that will be filtered, reducing the concentration of solids in suspension and facilitating the separation of these solids from the liquid. In addition, it facilitates the reduction of suspension viscosity, which makes filtration faster and more efficient. Due to the exceedingly poor quality of the sugar cane consumed during the first four months of this year, additional water was added during the affination process, resulting in a final liquor brix of 59.81º for January, February, March, and April. Nevertheless, and as mentioned previously, this facilitates a portion of the process, but increases the energy required to concentrate the solution further down the line. Therefore, the idea consists in discover a balance that is both advantageous and profitable for the business.

3.5 Equipment breakdowns

To ascertain the equipment's reliability, a historical analysis of equipment failures and maintenance interventions was conducted, by analysing company X's KPIs, including MTBF (mean time between failure), MTTR (mean time to repair), and downtime. MTBF is the average amount of time a system operates without failing, indicating its reliability. MTTR is the average amount of time needed to repair a system following a malfunction; it represents downtime. These metrics assist in evaluating and enhancing system performance and maintenance strategies, striving for longer intervals between failures (higher MTBF) and quicker repair times (lower MTTR). The MTBF increased by 16%, resulting in a cumulative average of 5 hours and 45 minutes. In carbonation, the mean time between failures decreased to 31 hours, indicating a decline in the section's reliability. MTTR, on the other hand, increased to 1 hour and fifteen minutes, a 16% increase over 2021, as shown in Figure 3.9.

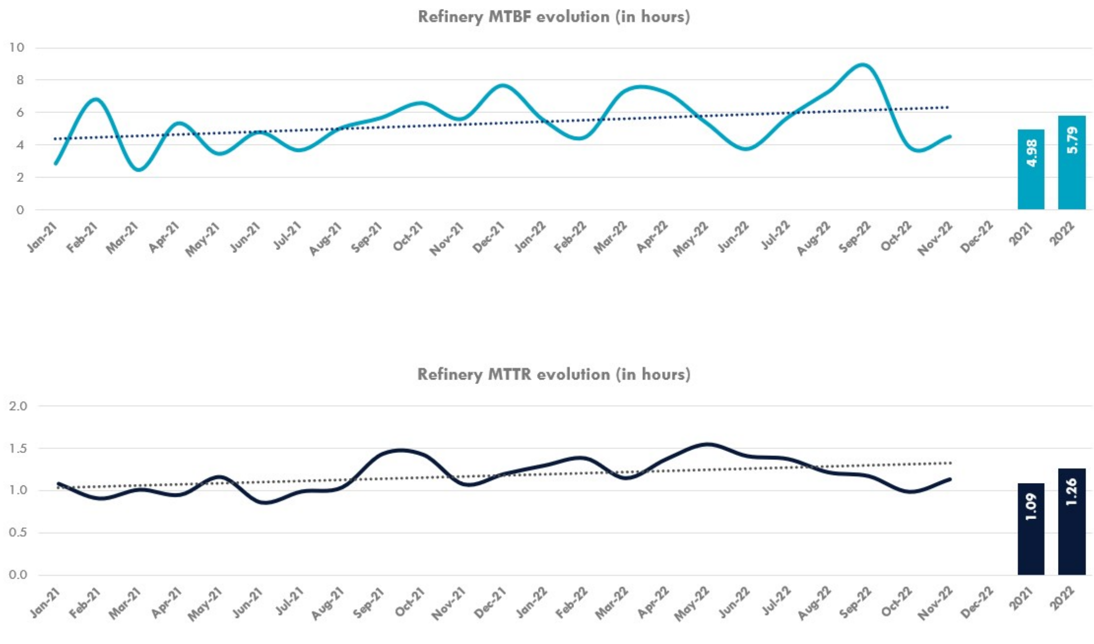


Figure 3.9: Graphical Representation of Maintenance KPI's

In 2022, the quantity of malfunctions and equipment downtime both increased by 100%. We proceeded to divide the breakdowns into three categories for a more accurate analysis of their impact:

- Breakdowns without loss of rhythm
- Breakdowns with loss of rhythm

- Breakdowns with refinery shutdown

As represented in Figure 3.10, the hours of breakdowns with loss of momentum experienced a 23% increase in the last year. Equipment shutdown hours that represented section shutdown increased by 458% (+2388h) and, finally, equipment shutdown hours that led to total refinery shutdown increased by 52% (+55h).

Figure 3.10: Graphical Representation of Equipment's Breakdowns

3.6 Conclusion

Based on the characterization performed, it is evident that the process is impacted by a number of issues. There are indications of bottleneck manifestation at the filtration stage, which are primarily caused by two factors: the consumption of sugar crude with unfavorable qualities and the absence of proper maintenance. This bottleneck hinders the effectiveness and efficiency of the filtration process, causing delays and diminished output and based on the presented data, it is notable that the filtration capacity only tends to decrease. In addition, an energy analysis reveals a decline in concentrated brix, which has a direct impact on the refining rate and indicates an energy bottleneck in the evaporators. The refinery's difficulties are exacerbated by the decrease in concentrated Brix, which slows down the refining process. In addition to these concerns, the reduction in maintenance time has led to a decrease in maintenance efforts and an increase in equipment failures throughout the refinery. Not only do these malfunctions disrupt operations, but they also pose potential safety risks. Understanding these issues and identifying their leading indicators necessitates prompt action to resolve them. Therefore, a comprehensive improvement vision is proposed to address the existing challenges and boost the organization's overall effectiveness and performance.

Chapter 4

Methodology

This chapter focuses on the optimization of sugar refinement procedures. The primary objectives include the development of a mass balance simulator, the implementation of mass analysis, the execution of energy analysis, the identification of process bottlenecks, the optimisation process parameter and the proposal of design solutions to alleviate these bottlenecks, thereby increasing the rate of sugar refining.

This chapter begins with a discussion of the construction of a comprehensive mass balance simulator. This simulator is essential for understanding the complex input-output relationships inherent to the sugar refining process. By facilitating the precise quantification of mass flows, it enables a comprehensive analysis of the system's dynamics, casting light on the interaction between the various components. In conjunction with the mass balance simulator, a comprehensive mass analysis is conducted to quantify the distribution and behavior of various constituents throughout the refining process. This analysis provides invaluable insights into the composition of raw materials, intermediate products, and refined sugar as well as the process parameters for each stage, allowing for the identification of enhancement opportunities.

In addition, an energy analysis is conducted to assess the sugar refining process's energy consumption patterns. This analysis is vital for identifying energy-intensive phases and opportunities for energy optimization, thereby promoting economic efficiency and environmental sustainability.

The identification of bottlenecks within the refining process constitutes a critical aspect addressed in this chapter. By means of systematic analysis and data-driven approaches, potential bottlenecks that impede overall efficiency and constrain the rate of sugar refining are identified.

A investments design solution is proposed to address these constraints. These solutions consist of innovative process modifications, apparatus upgrades, and operational strategies designed to eliminate the identified bottlenecks. The implementation of this solution has the potential to considerably increase sugar refining rates, resulting in increased productivity and enhanced overall process performance.

In conclusion, this chapter serves as a comprehensive exploration of sugar refining processes. Through the development of a mass balance simulator, the execution of mass and energy analysis, the identification of bottlenecks, and the proposal of design solutions, the objective is to optimize the process, elevate the rate of sugar refining, and unlock new avenues for efficiency improvements in the industry.

4.1 Mass Balance Simulation

As discussed in the preceding chapter, Company X confronts significant difficulties in collecting and analyzing process data due primarily to a lack of dependable measuring instruments. Flow meters and continuous brix meters are either absent or insufficient, resulting in significant difficulties in determining flow rates and parameters at each stage, particularly in regions devoid of measuring devices. This lack of data has made it extremely difficult to effectively analyze the process and identify potential bottlenecks. Consequently, the development of a simulator has become essential. By utilizing this simulator, the company hopes to circumvent the limitations imposed by the absence of measuring instruments, enabling a more thorough comprehension of the process and facilitating the identification of bottlenecks. In addition to this advantageous characteristic, the simulator, as its nomenclature implies, offers the capability to simulate hypothetical situations, thus aiding in the quantification of the implemented solutions with greater accuracy and precision.

Referred to as a simulator, the solution implemented is not a conventional simulator in the traditional sense. Instead, it takes the form of a spreadsheet, making it a manual simulator. It operates by performing mass balances for each step based on the provided parameters, inlet and outlet flow rates, and other relevant data. While not as advanced as automated simulators, this spreadsheet-based approach enables the company to perform critical mass balance calculations, aiding in process analysis and problem identification. The mass balance is a fundamental technique for analyzing the mass distribution within a system or process. This analytical method permits the determination of mass quantities entering, departing, and accumulating in a closed system, thereby providing valuable insights into mass dynamics and how process parameters affect within the system. By leveraging this manual simulator, Company X can gain insights and make informed decisions about their operations.

In the scope of this study, the mass balance simulation is conducted under a stationary work regime in an effort to investigate the intricate equilibrium between the various phases of the process under conditions approaching their maximal operational parameters. This stationary regime allows for a thorough examination of the mass distribution and interrelationships between the various system components. By emphasizing mass equilibrium, crucial insights regarding process efficiency, effectiveness, and overall performance can be garnered, thereby facilitating efforts toward improved resource utilization, production optimization, and system design refinements.

This exercise presents seven of the most important process parameters for each stage: color, purity, impurities, total mass, dry mass, brix, and sucrose. However, inputs are only submitted at specific stages of the process; the remaining parameters, which are not deemed inputs, are calculated based on the submitted input.

Next, the mass balance is presented. The simulator has two types of input, those inputs that are considered the annual averages (for this analysis, atypical days due to breakdowns were excluded) of the steady-state process, in other words, they are variables that are measured directly from the process and the other types of input that are the normalized parameters established by the company. These parameters encompass the process instructions or guidelines defined by the company to ensure standardized and consistent simulation outcomes. They serve as a reference for setting specific operational values for the operators. Using predefined equations or formulas, the parameters that are not considered process inputs are calculated. Instead of being expressly defined as process inputs, they are derived through analytical or computational methods. For an effective purging of outliers in the means used as inputs, the method known as removing outliers using interquartile range was applied (Method explain in Appendix E).

The mass balance begins with four distinct inputs: the polarisation, humidity, and color of the sugar cane consumed throughout the year, as well as the consumed tons of sugar cane (variables that are measured directly from the process):

$$Brix = 100 - \%Humidity \tag{4.1}$$

$$Dry\ Matter = Brix * Total\ Mass \tag{4.2}$$

$$Sucrose = TotalMass * Polarisations \tag{4.3}$$

As polarisation represents the percentage of sucrose in a solution

$$Impurities = DryMatter - Sucrose \tag{4.4}$$

It is presumed that everything present in the solution in the form of dry matter that is not sucrose is an impurity because it contributes nothing to the process

$$\%Purity = Sucrose/Dry\ Matter \tag{4.5}$$

The procedure begins with a comprehensive analysis of the consumed sugar cane's properties to determine its characterization. These characteristics have a substantial impact on the overall procedure, necessitating a thorough evaluation of each one. As previously demonstrated (Chapter 2), parameters such as starch content, dry matter quantity, and purity have a significant impact on filtration capacity.

The present exercise is based on the fundamental principle of The Law of Conservation of Mass, which plays an important role in scientific investigations. This principle states that the total mass of a closed system does not change over time. Therefore, it demonstrates that matter cannot be created or destroyed; it can only undergo transformations or be transferred between different substances. In other words, the mass of all substances entering a system must equal the mass of all substances leaving. Consequently, the application of mass balance facilitates the computation of pertinent parameters at each phase in light of the inflow and outflow of matter at each stage.

In addition, the transfer of energy in the form of heat and work, in conjunction with chemical reactions, is properly accounted for and has a significant impact on the parameters during each step of the procedure.

The primary objective of this mass balance is to perform an exhaustive analysis of the masses involved in the refining process. It makes it easier to determine the precise mass quantities that enter, exit, and accumulate within a closed system. In addition, the six aforementioned parameters fluctuate in accordance with the refining rate and the sequential stages of the process. As a result, the mass balance approach provides valuable insights into the flow rate and associated parameters specific to a given rate, thereby facilitating the identification of potential process bottlenecks. The procedure begins with the structural diagram depicted in Table 4.1,

Table 4.1: Sugar Cane Parameters

	Sugar Cane	
Dry Matter	**Colour**	**Total Mass**
Brix*Total Mass	Colour	Total Mass
	Sucrose	
	Dry Matter-Sucrose	
%Purity	**Impurities**	**Brix**
Sucrose/Dry Matter	Dry Matter- Sucrose	100-%Humidity

Table 4.1 represents the fundamental framework for the exhaustive refining procedure. This process involves the introduction, discharge, and accumulation of matter, followed by chemical and thermodynamic reactions that result in either white or brown sugar as the final product.

As mentioned previously, this exercise relies heavily on inputs, which serve a dual purpose: enhancing exercise efficiency and, therefore, establishing a closer alignment with real-world conditions by incorporating laboratory-measured parameters pertaining to process variables. There are two distinct classes of inputs:

- Laboratory measured parameters
- Company standard parameters

Standardized parameters pertain to predetermined quantities of material added by the company at specific stages, reflecting standardized practices within the operational methodology. As an illustration, in the context of carbonatation, Company X employs a specific ratio of approximately 56% per tonne of dry matter in the affination liquor. This ratio has been established as the optimal choice for process efficiency based on Company X's accumulated experience and comprehensive records of average lime consumption. In the same way, there exist additional working methods that are specified as inputs for other process steps, that also reflect working standardize methods. Regarding the measured parameters, the primary emphasis is placed on brix measurements performed at various stages of the process. When brix is used as an input for a particular step, the calculation of dry matter quantity or total mass is derived from these parameters using Formula

2.6. For example, if the total mass entering the process is known but the dry matter content is not, Formula 4.2 is used to calculate this parameter. On the other hand, Formula 4.6 is used for the opposite situation.

$$Total\ Mass = \frac{Dry\ Matter}{Brix} \tag{4.6}$$

Annex B provides a comprehensive explanation of the calculations involved in executing the mass balance, as well as precise instructions for performing accurate measurements and computations throughout the refining process. Also, it provides an exhaustive account of the intricate calculations entailed in executing the mass balance. This comprehensive document elucidates the deductions made during the process and systematically presents all pertinent parameters for each individual step. To provide contextualization and demonstrate the structure, Figure 4.1 illustrates an initial segment (besides this part there is still all the rest of the process) of a conducted simulation.

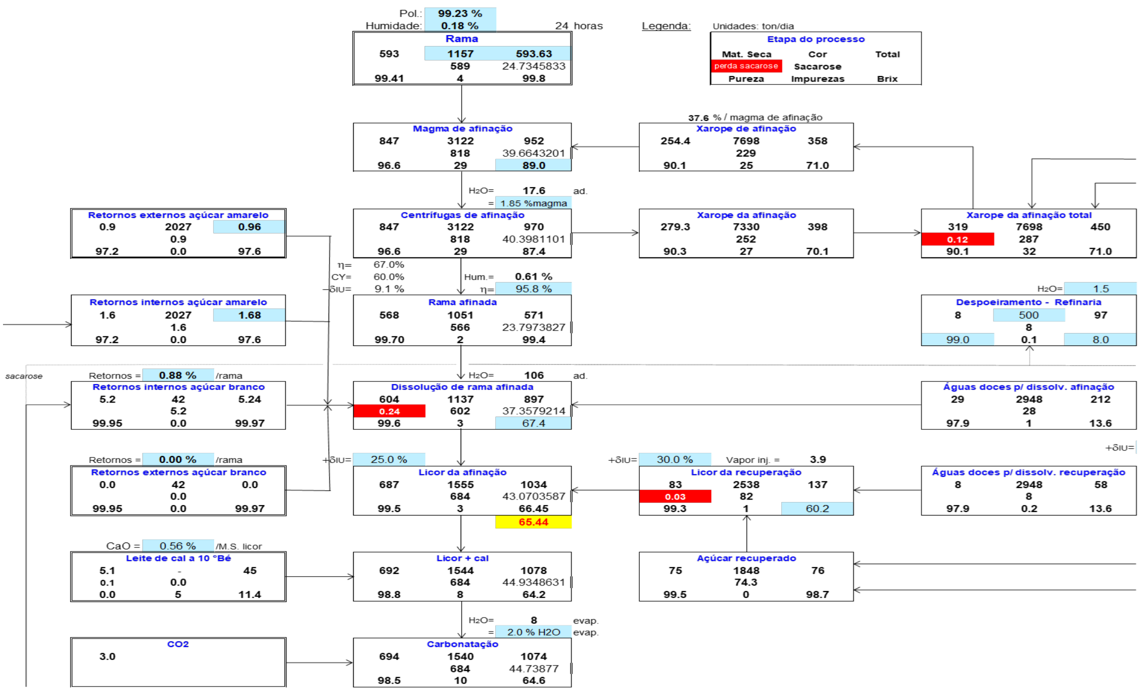


Figure 4.1: Example of a Mass Balance Simulation

The exercise requires the calculation of mass balances for each step, enabling a thorough examination of the refining process. This involves determining the amount of material at each stage and presenting the relevant process parameters on the same format as Table 4.1. These calculations and presentations continue until the desired final state is achieved, whether it be white or brown sugar. Such an approach ensures a comprehensive understanding of the refining process and enables efficient monitoring and control of production as a whole. The exercise concludes with the identical structure depicted in Table 4.2, but with parameters that correspond to the characteristics of the final product after traversing the entire refining process.

Table 4.2: Final White Sugar

Final White Sugar		
Dry Matter	**Colour**	**Total Mass**
Brix*Total Mass	DWS Colour	Total Mass
	Sucrose	
	Dry Matter*%Purity	
%Purity	**Impurities**	**Brix**
DWS %Purity	RCM Impurities	DWS Brix

4.1.1 Conclusions

In conclusion, the development of a mass balance simulator offers substantial benefits for sugar refining processes analysis. The simulator enables a thorough comprehension of flow rates through individual equipment components, thereby facilitating a comprehensive analysis and optimization of equipment performance. It also enables the precise determination of parameters for each stage of the refining process, thereby enhancing the characterization of each stage's processes. The simulator plays a crucial role in identifying process bottlenecks, facilitating the identification of constraints and inefficiencies that hinder the overall efficiency of the process. The simulator provides a versatile platform for comprehensive process evaluation and optimization by simulating volumetric flow under various scenarios, including non-realistic ones. It enables researchers and practitioners to engage in strategic planning, conduct rigorous risk assessments, and investigate hypothetical scenarios in order to evaluate their potential consequences. In conclusion, the mass balance simulator contributes to a better comprehension of the sugar refining process, enhanced optimization strategies, and enhanced performance.

4.2 Mass Balance Analysis-Identification of Bottlenecks

Understanding and optimizing the sugar refining process is significantly facilitated by the conducted mass balance analysis. This analysis incorporates a number of primary goals, including the determination of the mass flow rate traversing each equipment component using the presented mass flow simulator. Additionally, the maximum mass flow rate that each element of equipment can effectively accommodate is determined.

Potential bottlenecks in the refining process can be identified by conducting an exhaustive analysis of the mass flow rate passing through each piece of equipment and comparing it to the equipment's capacity. This comparative analysis facilitates the identification of equipment elements that may face difficulties in adapting to varying refining rates, thereby highlighting areas where production efficiency can be improved. By understanding the dynamics of mass flow rates and the limitations of each equipment unit, it is possible to implement targeted enhancements to eliminate identified bottlenecks and enhance process performance. Identifying and resolving these bottlenecks contributes to increased production capacity and efficacy in the sugar refining process, resulting in increased profitability. The evaluation and enhancement of mass flow within the process were essential to study ways of achieving the desired rhythms. This exercise is essential for addressing one of the OEE efficiency losses discussed in chapter 2, the reduced speed losses, it will enable to eliminate the bottleneck and, as a result, eliminate the root cause of these efficiency losses.

The exercise begins by conducting the mass balance (using the simulation described in Appendix B) for the 2022 refining rates of 22.96 ton/h, as presented in Appendix B Figure C.1 (21,75 ton/h of white sugar and 1,21 ton/h of brown sugar). Table 4.3 displays the results acquired for the flow rates processed by each piece of equipment at the selected frequency.

Table 4.3: 2022 Mass Flow Rate

Sugar Cane conveyor	**Affination centrifuges**	**Affination pumps**	**Licor + cal**	**Filtration**
29.09 m^3/h	28.14 m^3/h	33.1 m^3/h	34.87 m^3/h	34.68 m^3/h

Evaporators	**Vacuum pans**	**Refining centrifuges**	**Wet sugar conveyor**	**Drying sector**
27.86 m^3/h	43.91 m^3/h	34.55 m^3/h	28.55 m^3/h	28.32 m^3/h

Following that, conduct the mass balance for the desired state in order to compare the two scenarios and gain an understanding of the differences in the required steady-state rhythm. Again the exercise involved the simulation of the process (presente in Appendix B Figure C.2), in which the mass flow rate passing through each individual machine was adjusted to match the target rate of 27,28 ton/h (25,39 ton/h of white sugar and 1,89 ton/h of brown sugar) at a stationary state. The primary objective was to identify process bottlenecks and equipment that may not be able to sustain the desired rate (equivalent to an OEE of 80%). Table 4.4 exhibits the outcomes obtained for the volumetric flow rates handled by individual equipment components at the designated frequency.

Table 4.4: Mass Flow Rate for an 80% OEE

Sugar Cane conveyor	**Affination centrifuges**	**Affination pumps**	**Licor + cal**	**Filtration**
34.34 m^3/h	33.21 m^3/h	39.67 m^3/h	41.78 m^3/h	41.51 m^3/h

Evaporators	**Vacuum pans**	**Refining centrifuges**	**Wet sugar conveyor**	**Drying sector**
33.87 m^3/h	52.88 m^3/h	40.87 m^3/h	33.76 m^3/h	33.51 m^3/h

In order to realize the aforementioned analysis, it is imperative to calculate the mass capacity of each equipment unit. This entails a meticulous examination of the characteristics outlined in the equipment's manual or referencing historical records wherein the equipment operated at its rated maximum capacity. Consequently, the values shown in Table 4.5 were derived to denote the respective mass capacities of the equipment components:

Table 4.5: Maximum Equipment Capacity

Sugar Cane conveyor	**Affination centrifuges**	**Affination pumps**	**Licor + cal**	**Filtration**
39 m^3/h	50 m^3/h	50 m^3/h	50 m^3/h	35 m^3/h

Evaporators	**Vacuum pans**	**Refining centrifuges**	**Wet sugar conveyor**	**Drying sector**
30.6 m^3/h	55.7 m^3/h	53.1 m^3/h	37 m^3/h	37 m^3/h

The analysis of mass flow rates consistes of a comprehensive comparison between the mass flow rate in cubic meters per hour passing through each unit, as determined by the target rate of 27,28 ton/h, and the previously examined maximum mass flow rate of each unit (all rates considered are under stationary conditions). This enables the creation of the graph shown in Figure 4.2, which is representative of this comparison:

	Sugar Cane Conveyor	Affination Centrifuges	Affination Pumps	Licor + Cal	Filtration	Evaporators	Vacuum Pans	Refining Centrifuges	Wet sugar conveyor	Drying Sector
Overcapacity (%)	12%	34%	21%	16%	-19%	-11%	5%	23%	9%	9%
Capacity (m3)	39.0	50.0	50.0	50.0	35.0	30.6	55.7	53.1	37.0	37.0
Required Flow Rate (m3)	34.3	33.2	39.7	41.8	41.5	33.9	52.9	40.9	33.8	33.5

Figure 4.2: Available Flow Rate (Difference between maximum flow rate and average flow rate %)

The conducted mass analysis led to the following noteworthy conclusions:

1. **Filtration**: The investigation revealed a capacity shortage of -19% in the filtration unit. Consequently, the filtration process emerged as the primary bottleneck within the overall process. The identified capacity insufficiency indicated that the filtration unit was unable to effectively handle the required flow rate, thereby hindering the attainment of the desired production rate.

2. **Evaporation**: While the evaporation unit was not explicitly identified as a bottleneck, it exhibited an observable lack of capacity amounting to -11%. This insufficiency in capacity indicated that the evaporation unit struggled to adequately support the desired pace. The underlying cause of this capacity deficit was found to be an energy balance issue, which will be explored in detail in the subsequent chapter.

The results of the mass analysis highlighted the crucial role of the filtration unit as the primary bottleneck in the process flow, highlighting its importance. In addition, it was evident that the evaporation unit had a capacity deficit, though not to the same extent as the filtration unit. Inadequate capacity in both units impeded the achievement of the desired production rate.

These results highlight the significance of addressing the capacity limitations in the filtration unit and the underlying energy balance issue in the evaporation unit. Future chapters will delve further into examination and discussion in order to identify appropriate measures for enhancing the overall process's efficiency and achieving the desired production rate.

4.3 Energy Balance Analysis-Identification of Bottlenecks

The undertaking of an energy analysis holds significant importance as it allows for the identification and mitigation of potential mass restrictions within the process arising from inadequate energy capacity to support specific refining rates. So, this chapter examines the energy dynamics within the sugarcane refining process. The analysis concentrates specifically on steam-consuming equipment, as the refining process depends on three types of utilities: electricity, water, and steam.

Assuming ample supply of electric and water resources, as they are provided by the public network, the energy analysis focuses predominantly on steam-consuming equipment. This methodology permits a thorough evaluation of energy utilization and optimization potential in the refining process.

The identified equipment components that heavily rely on steam consumption include the:

- Dissolver
- Evaporator
- Vacuum Pans

These devices hold critical roles at various stages of the refining process and exert a significant influence on energy consumption. Regarding the remaining equipments where is possible to verify an increases temperature, this increase is due to an electric resistances. Despite their limited capacity, there was no evidence that they could potentially impede the process or function as an energy bottleneck. Consequently, these factors were not considered during the analysis. Similarly, despite being steam-powered, the Dissolver's sole purpose is to heat the solution without requiring any evaporation. As a result, it utilizes significantly less steam than the vacuum containers and evaporator. The solution's heating requirements do not necessitate significant quantities of steam, so it was not considered as important as the other two pieces of equipment mentioned.

The evaporator is the starting point for the analysis, as it exhibited the greatest lack of mass capacity. Prior to crystallization, the primary purpose of an evaporator is to concentrate a solution by heating it to its boiling point and evaporating the required quantity of water. In this particular instance, the evaporator has been designed to satisfy the specified parameters, including a 2 bar steam pressure. Figure 4.3 depicts the values for which the equipment is dimensioned.

Figure 4.3: Schematic Representation of the Evaporator

Using Formula 2.9 , it can be determined that the evaporator is capable of evaporating 5.87 ton/h under the provided input and output conditions:

$$\dot{m}_e = 42.30 \cdot (1 - \frac{62}{72}) = 5.87 ton/h \tag{4.7}$$

The value 42.30 ton/h is calculated by multiplying the density, $0.98 \cdot (1 + Brix \cdot \frac{200+Brix}{54000}) = 1.27 ton/\mathrm{m}^3$, by the discharge rate, 33.40 m^3/h. In order to make this evaporation possible it is necessary to heat the solution to saturation temperature and consequently transfer heat capable of evaporating the required flow of water. Table 4.6 outlines the thermodynamic properties required to calculate the total heat transfer (Table D.1 in Appendix D provides the boiling point elevation presented and Table F.2 in Appendix F provides the thermodynamic table for the latent heat presented).

Table 4.6: Evaporator Thermodynamic Characteristics

	Exhaust Steam	**1st Effect**	**2nd Effect**
Pressure	2 bar	0.75 bar	0.24 bar
Temperature	120.80ºC	92.40ºC	64.00ºC
Latent Heat	525.4 kcal/kg	544.1 kcal/kg	561 kcal/kg
Boiling Point Elevation	-	3.20ºC	4.10ºC

In order to quantify the heat transfer capacity, an empirical flash value must be assumed first. Assuming an empirical flash value of 1 ton/h (according to LYLE (1957) an usual value for this type of evaporators in this industry, however, it will be subject to confirmation later in this exercise) and considering that the transfer area and heat transfer coefficient are identical in both effects of the double-effect evaporator, it is reasonable to conclude that the evaporation rate in both stages is essentially identical:

$$\dot{m}_{e1} = \frac{5.87 - 1}{2} = 2.44 ton/h \tag{4.8}$$

Therefore, the discharge rate into the second effect is:

$$\dot{m}_{v,2} = \dot{m}_{v,1} - \dot{m}_{e1} = 42.84 - 2.44 = 40.40 ton/h \tag{4.9}$$

According to Table D.2, the self evaporation rate per ºC in this conditions is going to be approximately 1.22 kg/ton of liquor. Using Formula 2.12, the flash's true value will be:

$$\dot{m}_{e,f} = 1.22 * 40.40 * (95.70 - 68.10) = 1360 kg/h \tag{4.10}$$

Each kilogram of vapour from the first effect will evaporate water from the second effect's fluid in proportion to their respective latent heats. Consequently, if the evaporation in the first Effect is equal to:

$$\dot{m}_{e1} \cdot \Delta h_{lv1} = \dot{m}_{e2} \cdot \Delta h_{lv2} \Leftrightarrow (5.87 - 1.36 - \dot{m}_{e1}) \cdot 561 = \dot{m}_{e1} \cdot 544.1 \Leftrightarrow \dot{m}_{e1} = 2.29 ton/h \tag{4.11}$$

The evaporation in the second effect will equal to:

$$\dot{m}_{e2} = 5.87 - 1.36 - 2.29 = 2.22 ton/h \tag{4.12}$$

To quantify the evaporator capacity, only the maximal heat transfer of the first evaporator has been calculated, as the evaporation in the second effect will always be proportional to that of the first effect evaporator. In total, the first evaporation process corresponds to a 1,736.17 Mcals heat transfer.

$$\begin{aligned} Total\ Q &= Sensible\ Heat + Latent\ Heat = \dot{m} \cdot c_p \cdot \Delta T + \dot{m}_{e1} \cdot \Delta h_{lv} \\ &= 42.30 ton/h \cdot 0.70 \cdot (95.70 - 79.30) + 2.30 * 544.10 = 1736 Mcals \end{aligned} \tag{4.13}$$

However, several factors may impede the evaporator's ability to concentrate the solution within the highlighted parameters, making it a bottleneck in the overall process. Initially, a decrease in vapor pressure is possible. It is important to note that higher steam pressure generally provides more energy for vaporization, allowing for more effective liquid concentration. Consequently, a decrease in steam pressure could reduce evaporation efficacy and hinder the concentration process. In addition, a rise in the inlet flow rate, in other words, the rate at which the solution enters the evaporator, can present difficulties. A higher inlet flow rate necessitates a greater heat transfer capacity from the evaporator, which may exceed its design limitations and compromise the desired concentration results.

$$\dot{Q} = \dot{m} \cdot c_p \cdot \Delta T \tag{4.14}$$

Upon analyzing the formula under the assumption of a constant specific heat capacity (c_p) and a constant temperature variation, it becomes clear that the heat transfer to the solution must increase in direct proportion to the input mass increase in order to maintain the same final temperature. This implies that if the solution's mass is increased, a proportional increase in the quantity of heat energy transferred to the solution is required to maintain the desired temperature rise. The observed correlation between mass input and required heat transfer emphasizes the direct relationship between these variables when attempting to achieve the desired final temperature.

Likewise, a lower brix value at the inlet, which alludes to the concentration of sugar in the solution, can affect the performance of the evaporator. A lower Brix value necessitates a prolonged evaporation time because more water must be evaporated to reach the desired concentration. This prolonged process can lengthen the evaporator's residence time and exacerbate the bottleneck effect. Finally, a reduced temperature at the evaporator's inlet can reduce its efficiency. Reduced inlet temperature necessitates the inclusion of additional energy and sensible heat to bring the solution to the boiling point, thereby prolonging the evaporation process.

Reduction in steam pressure, an increase in inlet flow rate, a decrease in brix at the inlet, and a decrease in inlet temperature can all contribute to longer processing times and inhibit the evaporator's ability to concentrate the solution within the specified parameters. These factors highlight the potential function of the evaporator as a bottleneck in the overall process flow.

Regarding the vacuum pans, this equipment uses steam to transfer heat and concentrate a solution, utilizing sensible heat to achieve the evaporation temperature and latent heat to evaporate a specific amount of water. The equipment is designed to accommodate an input flow rate of 23,10 ton/h and an input brix value of 72 in the context depicted. It evaporates 4,72 tons of water per hour effectively, achieving the desired output brix of 90.50°. The steam pressure employed is 2 bars. Under the specified conditions, the cooking set time for the process is 120 minutes per boil. For each individual boil, a total of 18 tons of product is produced. Considering the presence of three vacuum pans, the production output during a full day, adhering to these conditions, would be calculated as 18 tons multiplied by 3 vacuum pans and further multiplied by 12 boilings per day per equipment, resulting in 648 ton/day. However, it is important to note that achieving this production quantity may not always be possible due to deviations or deviations from the optimal conditions.

A brix input value below the predetermined threshold of 72 is one contributing factor. The preparation time increases by approximately 8 minutes for every degree below 72. This situation may occur if the evaporators are incapable of concentrating the solution to the specified brix level due to a energetic bottleneck. Decreased inlet solution temperature is another significant factor. Suboptimal temperature at the entry of the equipment necessitates additional time and sensible heat to accomplish the desired evaporation temperature, thereby extending the cooking process. Inadequate liquor availability, as a result of defective processes in the refining centrifuges, can also result in longer cooking times. Inadequate distillate production hinders the concentration process, thereby reducing overall efficiency. The absence of adequate vapor pressure, which makes it impossible to evaporate the intended volume of water, is also highlighted. This restriction may result from insufficient steam supply or problems with steam generation, preventing the apparatus from operating at its maximum capacity. The cumulative effect of these variables results in cooking times that exceed the specified 120-minute threshold. In such a scenario, the vacuum pans, which comprise the aforementioned apparatus, become the energy bottleneck in the overall process, thereby limiting system efficiency and throughput.

4.4 Solution Design

Based on the analysis conducted, in order to achieve the desired production rate, it is necessary to implement measures that alleviate the bottleneck caused by the filtration process. Three potential solutions to this problem have been identified:

- Acquisition of a new Putsch filter: One possible solution is to invest in new filtration equipment from Putsch, which is anticipated to have improved capabilities and performance. It is anticipated that by upgrading the filtration equipment, the bottleneck caused by the current filtration unit can be reduced or eliminated.

- Purchase of sugar cane with improved filtration characteristics: Obtaining sugar cane with more favorable characteristics for the filtration procedure is a prospective alternative course of action. This involves procuring sugar cane with characteristics that facilitate smoother filtration, such as a lower starch content or a more uniform particle size distribution. By choosing sugar cane with these desirable characteristics, it is anticipated that the filtration process will be more efficient and effective.

- The addition of water during the refining phase results in a decrease in Brix. This decrease in Brix not only affects the carbonated beverages but also decreases the solution's viscosity. Therefore, the addition of water, along with the resultant decrease in Brix and viscosity, improves the filtration rate. The decreased viscosity allows for a smoother and more efficient passage of fluid through the filtration system, thereby increasing the filtration rate.

However, the disadvantages associated with the reduction in Brix obtained by adding water make this option less desirable. A notable disadvantage is the substantial increase in specific steam consumption, which results in increased costs. The substantial costs associated with evaporating the entire volume of added water make this method economically challenging.

In addition, according to a study conducted by Company X, a one-degree decrease in Brix at the inlet of the vacuum containers results in a 9-minute increase in cooking time on average. This prolonged simmering time has negative effects on process efficiency and could potentially disrupt the production rhythm. Furthermore, the evaporator's energy limitations prevent it from adequately concentrating the solution, prohibiting Company X's predetermined Brix value of 72 from being achieved at the vacuum pans' inlet. As a result, the extended cooking time considerably reduces the production rate, resulting in a substantial decrease in overall production efficiency.

In light of these drawbacks, the strategy of diminishing Brix by adding water may not be financially viable or operationally effective. In order to increase filtration rates without compromising energy consumption, production efficiency, or cost-effectiveness, the only alternative solutions is the investment in a new putsch filter or the Purchase of sugar cane with improved filtration characteristics. Due to the fact that Company X is presently unable to regulate the procurement of raw materials, only one of the three options remains: a new Putsch filter. In consequence, a business case addressing the different advantages of this option is presented in the following chapter.

4.5 Business Case Putsch Filter

The current filtration capacity of 35 m^3/h is determined by combining the capabilities of individual putsch filters, each of which has a filtration rate of 14 m^3/h, and a rotating filter with a capacity of 7 m^3/h. The introduction of a new putsch filter could increase the filtration capacity to 49 m^3/h. This additional filter would have the same filtration rate as the existing filters (m^3/h), thus increasing the overall capacity.

The expansion of filtration capacity yields three significant benefits. First, it causes an increase in the rate of refining, which fosters increased productivity and output. Secondly, the enhanced capacity alleviates the prevalent practice of reducing the brix during affination to achieve a higher filtration rate, which reduces the specific energy cost. Incorporating redundancy into the filtration system provides a valuable safety net. Due to the availability of secondary filters, the impact of a filter failure on the production rate is reduced by 50%. Therefore, the expansion of filtration capacity not only improves refining efficiency and output, but also optimizes cost-effectiveness by decreasing energy consumption. In addition, the implementation of redundancy ensures operational reliability, thereby enhancing the system's resilience to unforeseen circumstances.

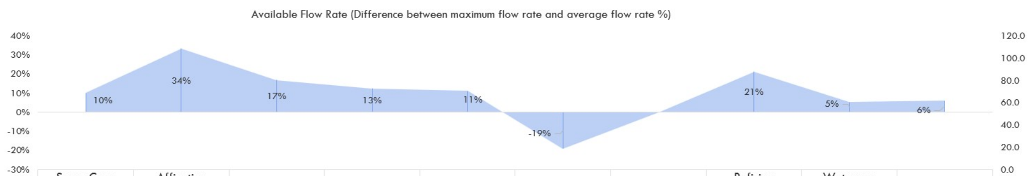
4.5.1 Refining Rhythm

In order to quantify these gains, the first step was to compute the refining rate in a hypothetical scenario in which company X invested in new filtration equipment. In the scenario where the identified bottleneck in the filtration process is successfully eliminated, as determined through the previous analysis of mass and energy, it is evident that certain equipment would be incapable of accommodating such elevated refining rates. Consequently, new bottlenecks arise within the process, impeding the desired pace of refinement. Therefore, the first stage was to determine the next mass bottleneck, in the event that filtration was no longer a problem for the process. A comprehensive examination of past data revealed a significant finding from 2008 during the historical analysis. The use of an Australian sugar cane with extremely favorable filtration characteristics was observed to have effectively eliminated the obstruction at the filtration stage. Consequently, the vacuum pans were identified as the refinement process's subsequent impediment. A comprehensive analysis of the vacuum pans revealed that they were the primary impediment in the refining process. Notably, they demonstrated the ability to conduct 39 boilings per day, which corresponded to 715 kilograms of sugar cane consumption. This resulted in a rate of 29.38 ton/h of white and brown sugar production. During the analysis, the significance of maintaining a brix level of approximately 72° at the entrance of the vacuum containers became apparent. This specific Brix level was found to be crucial for attaining the desired 39 boilings per day production rhythm (as mentioned before, in case of low brix the baking time increases, one less brix degree corresponds to 8 more minutes of boilings).

In sum, historical analysis of the moment when filtration ceased to be a bottleneck in the process, as a result of the consumption of a crude with characteristics favorable to filtration (low starch content and lower percentage of solids in suspension), revealed that the vacuum pans were the new bottleneck that dictated the rate of refining. According to historical documents, this equipment carried out 39 boilings per day, which corresponded to a raw material consumption of 715 tonnes, a refining rate of 29,38 ton/h, and a brix of 72º for the concentrated liquor (Brix at the entrance of the vacuum pans).

Thereby, the next step was to analyse whether the equipment responsible for providing a 72º brix in the concentrated liquor, the evaporators, would have the capacity to do so. To this end, an assessment about the energy capacity of the evaporators was subsequently perfomed. This evaluation sought to determine whether or not the evaporators had sufficient energy capacity to meet this requirement. The evaluation of evaporator energy capacity was essential for determining the viability and practicability of maintaining the intended refining rate.

As demonstrated by the energy analysis, the evaporator in question has a heat transfer limitation of 1,736.17 Mcals. Using the developed mass simulation, it is determined that the input flow rate of the evaporators under the specified conditions is 53.50 ton/h with an input brix of 61.50º. To achieve an output brix of 72º, which is a requirement for 39 boilings per day, the evaporator must be able to vaporize 7.81 ton/h of water from the solution. This rate of evaporation corresponds with a heat transfer of 2304.77 Mcals (calculations shown in excel spreadsheet in Figure G.2 in Appendix G). As a result, it is evident that the evaporator cannot sustain this rate, thereby manifesting as an energy constraint in this scenario.



	Sugar Cane Conveyor	Affination Centrifuges	Affination Pumps	Licor + Cal	Filtration	Evaporators	Vacuum Pans	Refining Centrifuges	Wet sugar conveyor	Drying Sector
Overcapacity (%)	10%	34%	17%	13%	11%	-19%	0%	21%	5%	6%
Capacity (m3)	39.0	50.0	50.0	50.0	49.0	30.6	55.7	53.1	37.0	37.0
Required Flow Rate (m3)	35.0	33.2	41.5	43.7	43.5	36.4	55.7	41.7	35.0	34.7

Figure 4.4: Schematic Representation of The Mass Balance

In order to determine the revised refining rhythm following the procurement of the new putsch filter, the revised bottleneck imposed by the evaporator would be required. To obtain a brix of 72 at the exit of the evaporators, the input flow rate must be decreased. The objective of this task is to determine the necessary adjustments based on a total heat transfer of 1,736.20 Mcals, which requires a backwards approach to mathematical calculations. To accomplish this, the discharge rate must be reduced to 45.73 ton/h. In addition, by obtaining new filtration equipment, the addition of water during the affination stage is rendered unnecessary, resulting in an input brix of 62.79º (which will be discussed in greater detail subsequently). These modifications are essential for optimizing the system and attaining the desired result. In Appendix G Figure G.4 is presented the excel spreadsheet used for the described exercise.

To sum up, Due to the presence of an energy blockage in the evaporator, it is necessary to reduce the refining rate to 45.73 ton/h in order to reach the daily target of 39 boilings. This adjustment is crucial for maintaining the intended brix level and optimizing the refinement process as a whole. Based on the exhaustive analysis performed, it can be concluded that the evaporators lack the capacity to accommodate the increased filtration rate, primarily due to their inability to attain a brix value of 72º in the concentrated liquor. As a result, the revised refining cadence is governed primarily by the newly identified bottleneck imposed by the evaporators. Utilizing the advanced mass simulator discussed previously, it was determined that the recalibrated refining rate under stationary conditions will be 25.60 ton/h, comprised of 24.53 ton/h of white sugar and 1.07ton/h of brown sugar.

Personnel costs, fixed costs, and structure costs remain unchanged in the scenario with the new putsch filter compared to the scenario without the filter. However, the costs of energy and basic materials will increase proportionally with the amount of sugar produced. As a consequence, personnel costs, fixed costs, and structure costs are dispersed, resulting in a decrease in the cost per ton (€/ton) of sugar. In particular, this reduction amounts to 17.93€ per ton, resulting from subtracting the 2022 €/ton from that obtained for the new scenario, as shown in Table 4.7.

Table 4.7: Table representing the specific cost reduction

	2022	**New Scenario**
Personnel Costs (Direct)	38.03€	34.23€
Utilities Costs	91.73€	82.93€
Sugar Cane Costs	600.00 €	600.00 €
Boiling Point	53.23€	47.91€
Total	**783.00 €**	**765.07 €**

4.5.2 Specific Energy Costs

The analysis of specific costs centered predominantly on the increased energy required to evaporate the water added to the process in order to increase the filtration rate. In the current scenario, however, the use of raw sugar with markedly unfavorable characteristics for the filtration section forces Company X to add around 1 ton of water per 20 ton of solution in the affination stage, resulting in an average brix of 59.81º in the evaporator's entrance. Conversely, analysis of the data pertaining to the months of October, November, and December, during which no water addition occurred in the affination area, reveals a higher brix measurement of 62.79º for the concentrated liquor. Assuming this comparison to compare the costs associated with this addition of water and based on the steam costs recorded at the intervals mentioned, there is a discernible increase in steam consumption per unit of sugar produced, amounting to an increase of 0,0982 tonnes of steam per tonnes of sugar produced. At a cost of 5,01 € per tonne of consumed steam, the impact is evident as daily savings of 5.01 € per ton produced are realized. This cost reduction is attributable to the modified method of lowering the brix by adding water, which aims to increase filtration rates.

4.5.3 Maintenance

Maintenance-wise, the addition of a new putsch filter would make the filtration procedure redundant. This redundancy would serve as a crucial mechanism in the event of maintenance shutdowns, permitting a transient reduction in the processing rate to 22.96 ton/h for 1.50 weeks. The maintenance team has designated this period of time for preventive maintenance tasks. Currently, the filtration process relies on three filtration units capable of filtering 14mm^3/h + 14m^3/h + 7m^3/h, for a total capacity of 35m^3/h; however, the capacity is reduced to 21m^3/h if one putsch filter fails. In the scenario with the newly acquired putsch filter, however, the filtration capacity in the event of a stoppage would still equate to 35 m^3/h, as a result of the addition of 14 m^3/h plus 14 m^3/h plus 7 m^3/h. Therefore, in the event of a clog, the reduction in filtration capacity would be reduced by 50% compared to the current configuration. This enhancement reduces the impact of maintenance-related interruptions on the overall filtration process, ensuring smoother operations and mitigating their disruptive effects. The new putsch filter's redundancy not only improves the system's dependability, but also facilitates more efficient and manageable maintenance procedures.

4.6 Optimisation and Normalisation of Process Parameters

In industrial processes, parameter optimization and normalization play a significant role in enhancing production efficiency. By establishing and maintaining process parameters within predetermined ranges, organizations can ensure the smooth operation of their apparatus and increase their output levels. In industries where efficiency is crucial, such as sugar production, optimizing and normalizing parameters becomes essential. This chapter's objective is to establish a reference point based on the average quality of primary materials and to identify optimal parameter ranges while taking into account potential variations. The advantages of efficient parameter optimization and standardization include a decrease in specific costs, an increase in the refinery's OEE, and ultimately a rise in production output. Sugar producers can maximize operational efficiency and output by meticulously optimizing their parameters to achieve the optimal balance. By fine-tuning these parameters, businesses can unleash the full potential of their production processes, ensuring cost-effectiveness, increased productivity, and overall success.

The brix level is one of the most important control parameter for the production process. By closely monitoring and sustaining this parameter within the optimal range, production can be significantly enhanced while specific costs are reduced. Controlling the brix levels in the refining liquor, final liquor, and concentrated liquor is crucial. Any deviation at the beginning of the process, especially with the tuning liquor, can have a negative impact on the brix levels of the final and concentrated liquor, making it difficult to accomplish the desired parameters. In situations where these parameters deviate, the lack of control over this parameter results in two significant losses:

- A decrease of 2.98° brix in the final liquor causes a significant increase of 5.01 € per ton in specific steam costs

- A one-degree decrease in brix levels in the concentrated liquor adds approximately eight minutes of cooking time in the vacuum containers. This prolonged preparation time results in a -2 ton/h loss

Controlling the brix parameter with care and precision is essential for optimizing production efficiency, minimizing costs, and achieving optimal results in the sugar production process. Furthermore, the brix parameter is closely linked to mass balance, as it is associated with the addition/evaporation of water in the process and the amount of dry matter in each stage.

Determining the optimal brix range for a process is a difficult and complex endeavor. This is especially true regarding carbonatation, the intermediate phase in the production of sugar. Carbonatation involves the inclusion of lime and subsequent filtration to separate impurities from raw sugar juice.

One factor to consider is the effect of reduced brix levels during affination. Lowering the brix levels can result in a decrease in viscosity, which, in turn, improves the efficiency of filtration. When the viscosity is reduced, the liquid flows more easily through the filtration system, allowing for faster and more effective impurity removal.

However, this advantage comes at a price. Reducing the brix levels during affination increases steam consumption per unit, resulting in increased operating expenses. The additional steam required to sustain the required temperature and pressure during the process can negate the benefits of enhanced filtration effectiveness.

In addition, reducing the brix levels during affination has a domino effect on the final liquor (inlet liquor in the evaporators). The brix level in affination influences the brix level of the concentrated liquor, which is the output of the sugar production process. If the input flow rate to the evaporator (the final liquor flow rate) is comparatively high, the evaporator may not be able to sufficiently concentrate the solution to achieve a desired concentrated liquor brix of 72º (or any other target value). Consequently, this circumstance results in an increase in cooking time, as the solution requires more time to attain the desired concentration, and a decrease in overall production efficiency.

However, excessively high brix concentrations in the affination stage can also be problematic. When the brix concentration is too high, the solution's viscosity increases. This increased viscosity slows down the filtration rate, producing a bottleneck in the process. As a consequence, the decreased filtration rate has a negative impact on the final output.

Finding the optimum brix intervals is, in conclusion, a delicate balancing act. The objective is to strike a balance between obtaining a good filtration rate and a concentrated liquor with a brix of 72º or as close to this value as feasible. This equilibrium guarantees effective filtration while minimizing boiling time and maximizing output. To determine the optimal brix range for a particular sugar production system, it is necessary to take into account variables such as viscosity, steam consumption, input flow rate, and overall process efficiency.

Determining the optimal brix values is a meticulous process that begins with identifying the concentrated brix. The primary objective is to attain the highest feasible brix in the concentrated liquor parameter without exceeding the critical threshold of 72-73º brix. This restriction is vital in preventing the undesirable occurrence of sugar crystallization. If the brix value exceeds this range, there is a possibility that sugar crystals will form prematurely, resulting in the loss of sucrose. These crystals can become entrapped in the deposit between the evaporators and pans, leading to the loss of valuable sucrose. To err on the side of caution, a conservative reference value of 72º brix is chosen for the concentrated liquor in an effort to establish a balance between maximizing concentration and preventing sugar crystallization.

Using Formula 4.15, it is possible to ascertain the brix of the final liquor by deriving it from the brix of the concentrated liquid.

$$\dot{m}_e = \dot{m}_{l,0} \cdot (1 - \frac{Final\ Brix}{Concentrated\ Brix}) \tag{4.15}$$

This formula provides a method for calculating the brix value from inputs that are already known. By contemplating the input flow and the quantity of water evaporated, the optimal value for the final brix can be determined. The capacity of the evaporator, as calculated in Chapter 3, determines its value, with a maximal capacity of 5.95 ton/h. Regarding the flow rate, It was employed the reference value derived from the mass balance analysis conducted for the period of 2022, which is 42.40 ton/h. As the mass balance analysis is conducted under optimal conditions to determine the optimum filtration rate, the value of 42.40 ton/h serves as a reference value in this exercise. By applying the formula to these values, the optimal brix value for the final liquor in the given context is obtained:

$$Final\ Brix = 72 * (1 - \frac{5.87}{42.40}) = 62.03 \tag{4.16}$$

Once the optimal values for the final brix and the concentrated brix have been determined, the optimal range for the brix for refining remains to be determined. The exercise initiates by identifying the factors responsible for these disparities:

- Inclusion of lime during the carbonatation process results in the production of water (chemical reaction in the mass balance page 61- Appendix B). This addition of water to the process has a diluting effect, resulting in a decrease in brix value.

- The washing of filters in each filtration cycle requires the inclusion of water during the filtration stage. This washing procedure, which is necessary for removing impurities and maintaining the efficiency of the filters, contributes to the system's overall water content and consequently influences the brix value.

- The resins utilized in the treatment cycles are able to retain the colored compounds present in the liquor. At the conclusion of each treatment cycle, the polymers must be cleansed and regenerated. This cleansing procedure adds additional water to the system, which further affects the brix value.

These three factors explain the difference in brix between the affintion liquor and the final liquor. According to the information provided in Annex 5, the observed decrease in brix is approximately 3.56º. Although there may be modest deviations due to equipment failures, non-compliance with standards, inaccurate caudalimeter calibration, or other factors, these deviations are comparable and can be considered constant. Notably, this decrease in brix value persists regardless of production rates, branch-specific consumption, or other factors. By recognizing this relatively constant value, the overall process can be better comprehended and managed. Based on the mentioned information, it is possible to formulate an equation that relates the brix in the affination and the brix of the final liquor:

$$Affination\,Brix = Final\,Brix + 3.56 = 62.03 + 3.56 = 65.59 \tag{4.17}$$

In conclusion, the determination of the optimal brix ranges is crucial to the optimization of various production process operational parameters. It requires striking a delicate balance between the filtration rate, evaporator capacity, and subsequent cooking time in the vacuum vessels. This effort attempts to minimize or eliminate potential bottlenecks, thereby optimizing the overall efficiency of the process. Through careful consideration of the evaporation capacity, cooking time, and filtration capacity, precise values have been determined to facilitate this equilibrium. It has been determined that maintaining the concentrated liquor at a temperature of 72º yields an optimal reference of 62.03º for the final liquor, resulting in a value of 65.59º for the affination liquor. These discerned values provide invaluable insights for enhancing the production process's effectiveness and efficiency.

Considering the potential variability of these values, an interval of ±0.7 (value considered by Company X in defining brix variability) around the nominal values was adopted. Consequently, the optimal range for the brix at tuning was determined as 64.89 to 66.29, obtained by subtracting and adding 0.70 to the nominal value of 65.59. Likewise, the optimal range for the final brix was computed as 61.33 to 62.73, while for the concentrated brix, it was established as 71.3 to 72.7. These ranges allow for accommodating fluctuations in the brix values, ensuring a more flexible and adaptable approach in attaining the desired outcomes within the given process.

Chapter 5

Discussion and Future Work

5.1 Discussion

The primary objective of this project is to bolster the market competitiveness of the company through the enhancement of its capabilities to effectively compete in the commodities sector. This entails maintaining low costs per tonne and offering products at competitive prices. In addition, the company aims to augment production output and optimize operational efficiency in order to strengthen its market position. The specific targets set for this endeavor involve achieving an Overall Equipment Efficiency (OEE) of 80% for white sugar and 48.50% for brown sugar. Attaining these OEE targets will enable the company to enhance its competitive edge, leverage cost advantages, and elevate its production performance, positioning it favorably in the market.

The execution of a meticulous mass and energy balance analysis played a pivotal role in identifying critical constraints within the process and accurately quantifying their impact. The study revealed filtration as a significant bottleneck, leading to a considerable 19% capacity shortage in relation to the desired white sugar OEE state of 80%. This capacity shortfall was determined through a rigorous comparison between the stationary flow rate required to achieve the desired rate and the maximum rated mass of the equipment. Consequently, the primary objective of this study was to explore and implement optimal strategies to mitigate or eliminate the adverse effects of this bottleneck on the overall process.

The primary objective of this project is to identify and quantify the optimal solution to address the identified bottleneck and effectively increase production. By implementing this solution, the company aims to enhance its market competitiveness. This is achieved through improvements in Overall Equipment Efficiency (OEE) and the optimization of production processes. The enhanced OEE and optimized processes not only lead to increased production output but also contribute to energy savings, promoting sustainability and operational efficiency. The comprehensive strategy pursued by the company demonstrates its commitment to energy conservation and resource optimization, thereby bolstering its market position.

In order to address the existing limitation, the proposed solution entails making an investment in a new filter that possesses similar capacities to the current ones. This strategic measure aims to increase the filtration capacity from 35m3/h to 42m3/h. Although the rotating filter could potentially enhance the filtration capacity to 49 m3/h, it was not taken into account in this analysis. This is because the objective of the endeavor is to improve the filtration system, as the rotating filter degrades the quality of the liquor. Consequently, the rotating filter should be utilized only as a last resort. The augmentation of the filtration capacity is expected to yield a corresponding increase in the overall efficiency of the refining process. However, it is important to acknowledge that this enhancement in filtration capacity will unavoidably introduce new bottlenecks, specifically in the evaporation process. As the filtration capacity is elevated, the subsequent stages of the production process may encounter difficulties in accommodating the heightened input originating from the filtration stage. Consequently, the findings of this study indicate that the forthcoming primary bottleneck is anticipated to manifest in the evaporation stage.According to the findings of this study, the evaporation stage is anticipated to be the site of the upcoming primary obstruction.

The projected consequences on the evaporation process are predicted to impose limitations on the overall process rate, capping it at 25.60 ton/h for both white and brown sugar. Despite the resulting trade-offs, the implementation of this proposed solution is anticipated to yield noteworthy enhancements in Overall Equipment Efficiency (OEE). Specifically, the OEE of white sugar is forecasted to escalate from its initial value of 69.00% to an impressive 76.80%, signifying a remarkable increase of 11.30%. Similarly, the OEE of brown sugar is expected to experience a substantial upturn from 42.00% to 46.10%, reflecting a noteworthy improvement of 9.76%.

From a financial standpoint, implementing the proposed solution results in substantial cost reductions. Primarily, personnel costs, fixed costs, and structural costs are eliminated, resulting in a decrease in the price per ton (€/ton). This reduction amounts to 17.93 €/ton, only related to the increase in production and therefore dissolved fixed costs equal to the year 2022 and the energy savings of 5.10 €/ton, related to the elimination of brix reduction during the refining process contributes to an energy. Combining these cost reductions, the new solution yields a total of 23.03€/ton, thereby justifying the expenditure.

In conclusion, the proposed solution effectively eliminates the existing process impediment, resulting in a significant increase in Overall Equipment Efficiency (OEE) for the production of both white and brown sugar. Additionally, the implemented solution results in significant reductions in steam consumption per unit. Consequently, the company's market competitiveness improves substantially, as evidenced by a significant reduction in specific costs of 23.03 euros per ton. This accomplishment demonstrates the solution's efficacy in mitigating process limitations and delivering tangible benefits, thereby strengthening the company's competitive market position.

5.2 Future Works

Despite the fact that the project's objectives were not achieved with this solution, future work can increase these indicators to the desired levels. Consideration of potential future research and development avenues reveals three noteworthy areas of concentration. Investing in a new evaporator to effectively resolve the identified energy bottleneck is an essential step towards improving the studied process. This strategic investment aims to maximize operational efficacy, eradicate bottlenecks, and boost performance. Second, a transition to a continuous operation model has the potential to streamline the overall workflow and maximize productivity within the investigated process. This transition necessitates the investigation of methodologies that incorporate various stages of the process seamlessly, thereby minimizing downtime and maximizing resource utilization. The application of Total Productive Maintenance (TPM) provides a valuable opportunity for optimizing equipment reliability, maintenance procedures, and overall operational effectiveness. By adopting TPM principles, organizations can achieve greater levels of productivity, reduced downtime, and improved overall equipment effectiveness. These three focal areas have significant potential for advancing the studied process and establishing the groundwork for future endeavors with enhanced performance and productivity.

Investing in a new evaporator is a strategic solution to the identified energy barrier in the refining process. The investigation into the investment in a new putsch filter has revealed the existence of an energy bottleneck in the evaporation stage, thereby steering the prioritization of refinery system investments. In light of the preceding findings, the following investment strategy is outlined. Priority is given to the procurement of a new putsch filter due to the numerous previously discussed factors. The new putsch filter is essential for overcoming existing limitations and optimizing the overall system's efficacy. Once the new putsch filter is installed, it paves the way for the next investment, which consists of new equipment capable of sustaining the increased refining rates made possible by this enhanced scenario. This suggested solution holds the key to eliminating bottlenecks in the refining process at the mass and energy levels. By integrating a new evaporator to resolve the energy bottleneck, the system can operate more efficiently and effectively. In addition, the elimination of these obstructions allows for the achievement of new goals within the refining process. In addition, it is possible to optimize parameters that may hinder production rates, thereby facilitating a more streamlined and productive workflow. The investment in a new evaporator emerges as a crucial stage for eliminating energy bottlenecks in the process. This strategic decision, guided by the study's findings, not only opens up new opportunities for achieving refining goals, but also assures the optimization of critical parameters that might otherwise limit production rates.

Within the context of the investigated refinery process, the adoption of a continuous operation paradigm offers numerous benefits. One of the primary reasons for instituting continuous operation is to maximize the available maintenance time at the refinery. Significantly, in 2022, the stationary refining time increased by 25 days. Theoretically, this extended duration should correspond to a 9.60% increase in white sugar extraction, or 13,000 tons. Due to a slower production rate of -1.20 ton/h per hour, the actual quantity of white sugar increased by only 4,200 tons, or 3.10%. This disparity in production volume growth despite the extended operating time highlights the significance of 5.80 ton/h rhythm losses, which are predominantly observed during "normal" refining periods. It becomes apparent that the impact on refining rhythm is most pronounced during suspension and start-up days, highlighting the opportunity to improve the overall refining rhythm by minimizing these transitional periods. By transitioning to continuous operation, the reduction in stoppage and start-up days would substantially improve the refinery's rhythm. As a result, overtime labor would no longer be necessary, allowing for improved maintenance planning and extended open time, ultimately resulting in a reduction in breakdown occurrences. Adopting continuous operation not only streamlines the production process, but also enables improved maintenance strategies, resulting in increased operational efficiency and decreased downtime.

Total Productive Maintenance (TPM) is usually implemented to address the challenges of refinery downtime and equipment failures. Aligned with Lean Thinking principles, TPM offers compelling benefits for the refining process. First, TPM improves product quality through proactive maintenance practices and timely inspections, thereby reducing equipment-related defects and breakdowns. It assures optimal equipment performance, thereby decreasing the likelihood of quality deviations and customer discontent. Second, TPM improves equipment availability by implementing preventive and predictive maintenance strategies, thereby minimizing unscheduled delay and ensuring continuous production. By addressing prospective equipment failures in advance, TPM reduces production schedule disruptions and improves overall operational efficiency. Furthermore, TPM improves equipment reliability by nurturing a proactive maintenance culture and employee participation. Regular inspections, prompt maintenance, and efficient equipment management reduce unscheduled equipment failures, resulting in a more stable and reliable production environment. In addition, TPM has a positive effect on Overall Equipment Effectiveness (OEE), by optimizing availability, performance, and quality, TPM increases OEE by reducing equipment outage, enhancing equipment performance, and minimizing quality defects. This results in increased production effectiveness, decreased waste, and enhanced operational performance.

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Appendix A

Refining Process

This appendix provides a thorough overview of the complete procedure described in Chapter 2. It incorporates the journey of raw sugar from its initial entry into the process to the final stages, where either white or brown sugar can be produced. Every process step is presented meticulously, providing a thorough comprehension of the sequential transformations and operations that occur. By examining this appendix, readers acquire a comprehensive understanding of the entire process, enabling them to comprehend the complexities and subtleties involved in producing white and brown sugar variants.

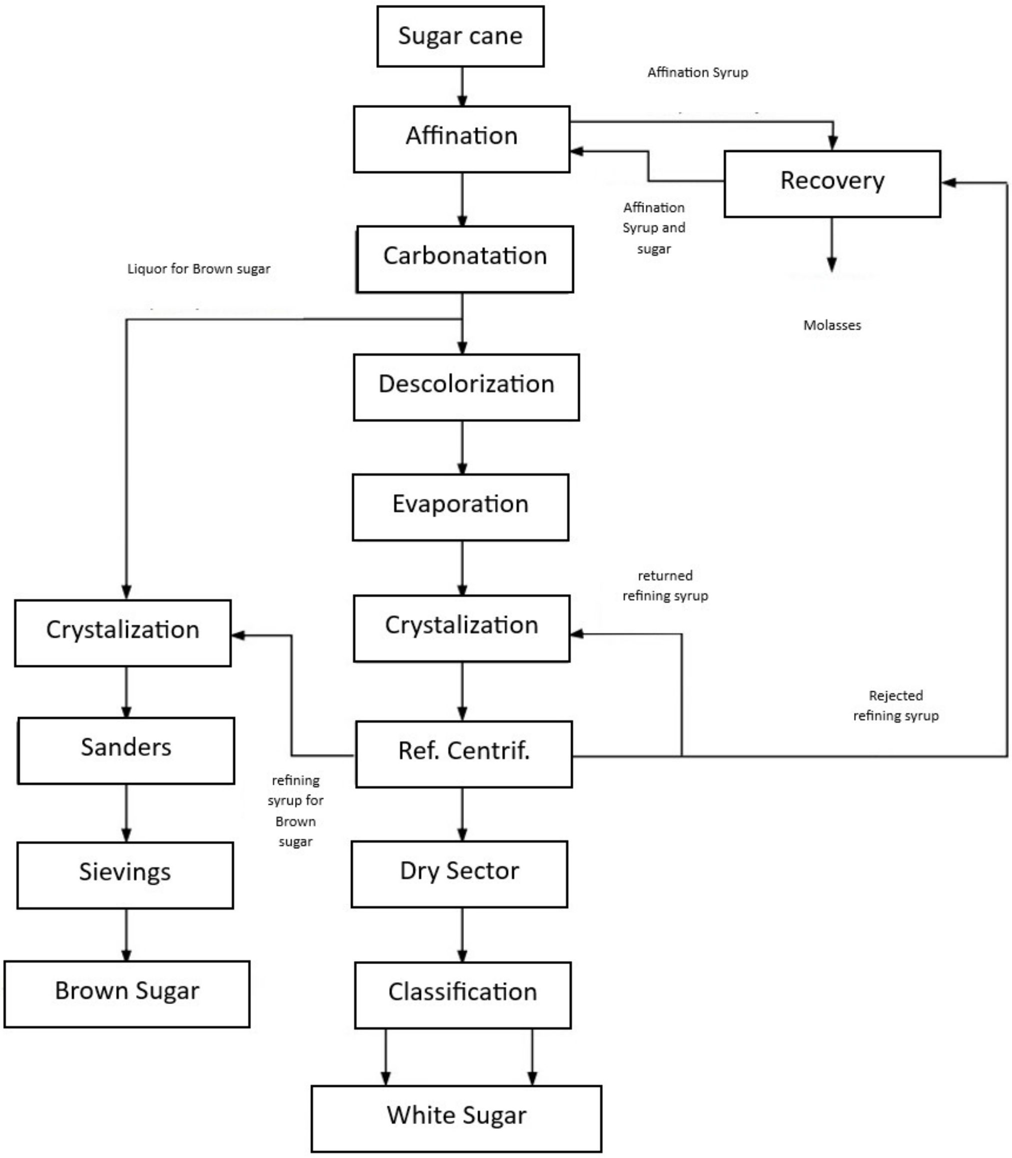

Figure A.1: Schematic representation of the refining process

Appendix B

Mass Balance

Following the previously described flow, sugar cane (SC) is transported from the raw material warehouse to the area where it is crushed. In the first magma mingler, the raw material is combined with the affination syrup(AS) produced by the centrifuges (downstream stage). At this point, the mass balance is conducted based on the brix input, with the remaining parameters being calculated as a result of this input. In other words, knowing that the average brix will be Xº (average value obtained from the laboratory analyses throughout the year at a steady rate), it is possible to compute the total amount of matter in the mingler using the formula deduction below and assuming that the total amount of dry matter in the magma solution is given by the sum of the sugar cane dry matter and the affination syroup dry matter:

$$\frac{Sugar\ Cane\ Dry\ Matter + Affination\ Syroup\ Dry\ Matter}{Magma\ Brix} = Magma\ Total\ Mass$$

$$\frac{Sugar\ Cane\ Dry\ Matter + Affination\ Syroup\ Dry\ Matter}{Magma\ Brix} = \frac{Magma\ Dry\ Matter}{Magma\ Brix}$$

$$Sugar\ Cane\ Dry\ Matter + Affination\ Syroup\ Dry\ Matter = Magma\ Dry\ Matter$$

$$SC\ Total\ Mass * SC\ Brix + AS\ Totl\ Mass * AS\ Brix = Mag\ Total\ Mass$$

$$SC\ Total\ Mass + AS\ Brix * (Mag\ Total\ Mass - SC\ Total\ Mass) = Mag\ Brix * Mag\ Total\ Mass$$

$$\begin{gathered} SC\ Dry\ Matter - SC\ Total\ Mass * AS\ Brix = \\ Mag\ Brix * Mag\ Total\ Mass - AS\ Brix * Mag\ Total\ Mass \end{gathered}$$

$$Mag.\ Total\ Mass = \frac{SC\ Dry\ Matter - SC\ Total\ Mass * AS\ Brix}{Mag\ Brix - AS\ Brix} \tag{B.1}$$

Except for the syrup's brix value, all other values have already been calculated. The brix value of the affination syrup is derived from the laboratory's average value (in all average values outliers are eliminated), and is therefore deemed an input for this simulation.

Regarding color, whenever there is no variation in the internal energy of the process, the color is the result of the weighted average of the two mingled materials and is therefore determined as follows:

$$Mag\ Colour = \frac{SC\ Dry\ Matter * SC\ Colour + AS\ Dry\ Matter * AS\ Colour}{Mag\ Dry\ Matter} \tag{B.2}$$

The quantity of sucrose is determined by summing the amount of sucrose present in each material introduced into the process, while the other parameters are calculated using the formulas described previously. In summary, the parameters of this stage will be as follows:

Table B.1: Affination Magma

Affination Magma		
Dry Matter	**Colour**	**Total Mass**
Brix*Total Mass	Equation B.2	Equation B.1
	Sucrose	
	SC Sucrose+ AS Sucrose	
%Purity	**Impurities**	**Brix**
Sucrose/Dry Matter	Dry Matter- Sucrose	Input

The liquor is then transferred to the centrifuges via overflow. The primary objective of this stage is to remove the molasses coating from the sugar crystals obtained from the initial extraction procedure. This coating of molasses contains non-sugar residues, including organic and inorganic impurities. Consequently, given that there is no loss of dry matter during the process, the quantity of sucrose resulting from centrifugation is identical to that of the previous step. However, as a result of the addition of water to the process for the purpose of cleaning and rinsing the separated solids or products, the brix decreases and the quantity of matter increases.

The amount of water added to a process can be determined by comparing the resultant affined sugar cane (ASC) and affination syrup (AS) after the centrifugation step. For determining the syrup's brix, laboratory measurements are used as a guide (so it is considered as an input). Using the expressions B.5, it is possible to repeat the calculations until the desired brix value is reached:

$$AS\,Dry\,Matter = Aff.\,Centrifuges\,Dry\,Matter - ASC\,Dry\,Matter \tag{B.3}$$

$$AS\,Total\,Mass = Magma\,Total\,Mass + Water\,Added - ASC\,Total\,Mass \tag{B.4}$$

$$AS\,Brix = \frac{AS\,Dry\,Matter}{AS\,Total\,Mass} \tag{B.5}$$

The process of affination in the refining operation yields two main products: affined sugar cane and affination syrup. While the brix of the syrup can be determined through laboratory measurements, the exact amount of water added in the centrifuge remains unknown. To calculate the precise quantity of water added, an iterative process is employed. Considering that the total mass in the centrifuge is a result of water addition, and the brix of the syrup is determined by equation B.5, which relies on equations B.3 and B.4, an iterative approach is employed. This iterative process aims to ensure that the calculated brix matches the recorded result, ensuring accuracy and consistency in the thinning process.

Table B.2: Affination Centrifuges

Affination Centrifuges

Dry Matter	**Colour**	**Total Mass**
Magma Dry Matter	Magma Colour	Mag. Total Mass*(1+%Water Added)
	Sucrose	
	Mag Sucrose	
%Purity	**Impurities**	**Brix**
Mag %Purity	Mag Impurities	Dry Matter/Total Mass

Centrifugation facilitates the segregation between the liquid fraction and solid fraction of the solution, resulting in the production of two distinct materials at the outlet: affined sugar cane, which is obtained through the elimination of the aforementioned impurities, and affination syrup. In terms of mass balance for this stage, it is assumed that the separation exclusively occurs within the dry matter of the cane. Consequently, no particle removal takes place in the dry matter introduced from the affination syrup, as these particles have already undergone the centrifugation process. Accordingly, to obtain the dry matter content in the refined cane, it is estimated that merely approximately 4.2% of the impurities present in the dry matter of the cane are separated during the centrifugation process, resulting in a total dry matter equal to SC Dry Matter* 95,8/100 (value based on the literature review for centrifuge efficiency in affination). As it pertains to brix, the humidity parameter is measured at this stage of the procedure and is thus an input. At this stage, the brix value can be obtained by subtracting 100 from the percent relative humidity.

Table B.3: Affined Sugar Cane

Affined Sugar Cane

Dry Matter	**Colour**	**Total Mass**
Magma Brix	Magma Colour	Mag. Total Mass*(1+0.0185)
	Sucrose	
	Mag Sucrose	
%Purity	**Impurities**	**Brix**
Mag %Purity	Mag Impurities	100-%Humidity

In the phase of dissolution, returns from the classifiers, drier, Yellow Sanded Sugar sieves, Packaging, Warehouses, and customers are added. Affined Liquor is the name of the ultimate product. Therefore, the amount of dried matter and sucrose results from the addition of all the quantities of matter mentioned. In addition to the aforementioned components, water must be added to this composition in order for it to be thoroughly combined.

As the brix is again an input, the total mass is calculated on the basis of this value and of the value obtained for dry matter. The remaining parameters are also calculated using the same formulas used previously. Therefore, the simulation parameters for this phase are set as follows:

The Recovering Liquor (RL) is added to the mixture from the previous stage in order to obtain the final product, Affined Liquor (AL). The sugars recovered by crystallization of the affination and Refining Syrups are the source of this liquor. This new mixture is heated to kill any microorganisms that may be present in the solution. Consequently, the mixture's internal energy fluctuates, and the corresponding energy is added to the formula used to calculate the color, which is

Table B.4: Dissolution of affined sugar cane

Dissolution of affined Sugar Cane		
Dry Matter Sum of all rejected components	**Colour** Weighted Average colour	**Total Mass** Dry Matter/Brix
	Sucrose Sum of all the sucrose present in the rejected components	
%Purity Sucrose/Dry Matter	**Impurities** Dry Matter-Sucrose	**Brix** Input

then computed as follows:

$$AF\ Colour = \frac{ASC\ Dry\ Matter * ASC\ Colour(1+\delta U) + RL\ Dry\ Matter * RL\ Colour}{AL\ Dry\ Matter} \tag{B.6}$$

Therefore, both the total amount of matter and the amount of sucrose in the solution are the result of adding the previously calculated matter to the matter from the recovery fluid, with the remainder parameters being calculated using the previously described formulas.

Table B.5: Affined Liquor

Affined Liquor		
Dry Matter Sucrose+Impurities	**Colour** Equation B.6	**Total Mass** ASC Total mass+RL Total Mass
	Sucrose ASC Sucrose+RL Sucrose	
%Purity Sucrose/Dry Matter	**Impurities** ASC Impurities+ RL Impurities	**Brix** Dry Matter/Total Mass

Carbonation is the next stage; The Affination Liquor is sent from the Affination to the Carbonation. In this step the liquor is first mixed with Milk of Lime (a mixture of calcium hydroxide in water) and then carbonated in Saturators (Saturation Towers). This procedure is crucial for effective carbonation, as will be explained in the following section.

At this stage, based on literature review, lime (CaO) should be added at a ratio betwwen 0.5% to 0.8% per tonne of dry matter present in the affination liquor. Typically, lime is used in the beet refining process, but, as this is a sugar cane refining process, the conversion to the chemical component used must be made. In the case of company X, they resort to the use of hydrated lime ($Ca(OH)_2$), and based on Company X's experience and records of average lime consumption, the percentage used is about 56% per tonne of dry matter in the affination liquor. Consequently, the conversion is performed as follows:

Beginning with the chemical reaction's equation, how to obtain hydrated lime from lime is described

$$CaO + H_{20} \longrightarrow Ca(OH)_2$$

As the molar mass of Ca=40, O=16 and H=1, to simplify the explanation, the equation can be described as follows:

(40+16) + (2+16) -> 74

According to the preceding equation, 56 units of CaO plus 18 units of H2O yield 74 units of Ca(OH)2. It indicates that the ratio between hydroxide and CaO is 1.32 (74/56). This implies that 1 ton of CaO and 0.32 tons of water will produce 1.32 tons of Ca(OH)2, which is, therefore, the minimum amount of water required for calcium hydroxide preparation. So calcium hydroxide contains 75.7% CaO and 24.3% H2O. In sum, the dry matter of hydrated cal (considering dry matter only the amount of $Ca(OH)_2$) is calculated with the following formula:

$$Hydrated\ Cal = 0.0056 * AF\ Dry\ Matter * 74/56 \tag{B.7}$$

When preparing hydrated lime in accordance with company X's standard parameters, the final solution must have a concentration of 10º Baumé. Using the reference table F.3 in appendix F to determine the calcium oxide (CaO) concentration in the solution at 10º Baumé, a value of 8.74% solute is obtained. This CaO percentage is then multiplied by a factor of 1.32 to effectuate the conversion to hydrated lime (Ca(OH)2). The resultant proportion acquired from the conversion delineates that in a solution of 100 tons, an estimated mass of 11.54(8.74*1.32) tons of hydrated lime, constituting the solute, is present. Given a desired quantity of dry matter in tons of hydrated lime, the total mass is calculated by making the proportion, as shown:

$$Hydrated\ Cal\ Total\ Mass = \frac{Hydrated\ Cal\ Dry\ Matter * 100}{11,54} \tag{B.8}$$

Table B.6: Liquor+Cal

	Liquor+Cal	
Dry Matter AL Dry Matter+ equation B.7	**Colour** Weighted Average Colour	**Total Mass** AL Total Mass+ equation B.8
	Sucrose AL Sucrose	
%Purity Sucrose/Dry Matter	**Impurities** Sucrose-Impurities	**Brix** Dry Matter/Total Mass

The next step is to transfer the solution to the saturators. Carbon dioxide (commonly known as Carbonation Gas) is injected into the liquor in this step. This gas reacts with lime to produce calcium carbonate precipitate, as illustrated below:

$$Ca(OH)_2 + CO_2 \longrightarrow Ca(CO)_3 + H_2O$$

When calculating the dry matter at this stage, it is necessary to subtract the amount of water formed by the reaction between carbon dioxide and hydrated lime, as represented in the equation shown. Due to the fact that the ratio of calcium hydroxide molecules to water formed is 1/1, the dry matter of this step can be calculated as follows (assuming 44 g/mol of H_2O and 18 g/mol of CO_2):

$$Carbonation\ Dry\ Matter = (Licor + Cal)\ Dry\ Matter + CO2\ Mass * (1 - 18/44) \tag{B.9}$$

Regarding the total mass, it is calculated by adding the amount of CO_2 to the amount of mass from the previous stage and subtracting a percentage of water that is evaporated during the carbonation process (around 2% based on the flow meter of the condensate present in the equipment). Consequently, the parameters for this phase will be:

Table B.7: Carbonatation

Carbonatation		
Dry Matter	**Colour**	**Total Mass**
Equation B.9	Weighted Average Colour	Equation XX
	Sucrose	
	Liquor+Cal Sucrose	
%Purity	**Impurities**	**Brix**
Sucrose/Dry Matter	ASC Impurities+ RL Impurities	Dry Matter/Total Mass

Still in the carbonation section, this precipitate and a portion of the Affination Liquor's impurities are separated by filtration to produce the Filtered Liquor. In order to facilitate the removal of impurities, it is necessary to cleanse the responsible filters after each filtering cycle. To accomplish this, a specific quantity of water must be added, namely 480 percent of the resulting dry cake. During this phase, approximately twenty percent of the total quantity of water added is redirected to the process itself. On the other hand, the remaining washing water and carbonated sediment produced by the process are expelled from the system. This expulsion is essential for the elimination of undesirable waste, allowing the treatment process to continue in a sustainable and efficient fashion.

Thus, in the filters, the quantities of sucrose and impurities will equal those determined in the preceding step, and the remaining parameters will be calculated based on these values (table B.8, with the exception of the total matter, which requires the addition of washing fluids.

Table B.8: Filtration

Filtration		
Dry Matter	**Colour**	**Total Mass**
Sucrose+Impurities	Weighted Average Colour	Carbonatation Total Mass*(1+%Water)
	Sucrose	
	Carbonatation Sucrose	
%Purity	**Impurities**	**Brix**
Sucrose/Dry Matter	Carbonatation Impurities	Dry Matter/Total Mass

This process will result in the formation of two categories of solutions: filtered liquor (FL) and carbonated sludge (CS). As previously indicated, only 20% of the washing waters enter the main process, so focusing solely on the filtered liquor will result in a decrease in color and mass in comparison to the previous stage. At this juncture, both the brix and the color are company-recorded annual averages, so all other parameters will be calculated based on these two. Except the impurities, this parameter is calculated presuming that the difference between the impurities that enter the carbonation process (affination liquor) compared to those that exit (filtered liquor) is around 60%.

After filtration, the sugar undergoes a decolorization process. Decolorization is a pivotal step in the sugar cane refining process, aimed at eliminating impurities responsible for the coloration of the final sugar product. This process assumes paramount importance in achieving the desired level of whiteness, purity, and visual attractiveness of refined sugar. The objectives of decolorization

Table B.9: Filtered Liquor

Filtered Liquor		
Dry Matter	**Colour**	**Total Mass**
F. Dry Matter+ CS Dry Matter	Input	FL Total Mass+0.2*Water Added
	Sucrose	
	Dry Matter-Impurities	
%Purity	**Impurities**	**Brix**
Sucrose/Dry Matter	0.6* AL Impurities	Input

can be summarized as follows: improving aesthetic quality by reducing color intensity, enhancing purity by eliminating non-sugar compounds and pigments, standardizing color specifications to meet market demands, optimizing sugar yield by selectively removing impurities, and enhancing the stability and shelf life of the sugar product. Decolorization plays a pivotal role in producing high-quality, visually appealing, and pure sugar products that align with consumer expectations and industry standards.

In order to decolorize sugar solutions, company X passes them through columns containing anionic polymers. These resins have a rigid structure, forming small porous spheres with active sites where the dye is exchanged for the resin ion, thereby removing the color from the sugar. The calculation of the total mass is give by the following equation:

$$Decolorization\,Total\,Mass = FL\,Total\,Mass + Water - Brown\,Liquor(BL) \tag{B.10}$$

Resin passage is achieved through treatment cycles, at the conclusion of which the resin must be cleansed and regenerated. According to the labor standards, 42m3 of water must be added per regeneration; thus, the amount of water added equals the amount of water added multiplied by the number of regenerations. To calculate the average number of cycles per day, the calculation involves solving the formula (formula mentioned in the literature review for calculating the colour load):

$$NCycles = \frac{\frac{FL\,Dry\,Matter \cdot (FL\,Color - Final\,Liquor\,Color)}{Color\,Load}}{7000} \tag{B.11}$$

with the color charge being a known process input and 7,000 representing the daily quantity of resin used.

Table B.10: Decolorization

Decolorization		
Dry Matter	**Colour**	**Total Mass**
FL Dry Matter-BL Dry Matter+ CS Dry Matter	FL Color	Equation B.10
	Sucrose	
	FL Sucrose-BL Sucrose	
%Purity	**Impurities**	**Brix**
Sucrose/Dry Matter	Dry Matter-Sucrose	Dry Matter/Total Mass

At the conclusion of the procedure, a solution with less color, designated final liquor (FIL), is obtained. As the color and brix of the solution are again measured in this phase, they are considered simulation inputs and the other parameters will be calculated based on them. Due to the fact that decolorization produces not only final liquor (FIL) but also fresh water and effluent, the total mass and quantity of sucrose of this solution can be calculated by subtracting the total mass and sucrose present on the effluent and sweet water, as shown in the subsequent.

$$FIL\,Total\,Mass = Decolorization\,Total\,Mass - Efluent\,Total\,Mass - Sweet\,Water\,Total\,Mass \tag{B.12}$$

$$FIL\,Sucrose = Decolorization\,Sucrose - Efluent\,Sucrose - Sweet\,Water\,Sucrose \tag{B.13}$$

Comparing the solution entering and exiting the decolorisation procedure, a X% reduction in color can be observed:

$$\frac{FIL\,Color - Filtered\,Liquor\,Colour}{Filtered\,Liquor\,Colour} * 100 = X\% \tag{B.14}$$

As mentioned previously, the decolourisation process is aimed at eliminating impurities responsible for the coloration of the final sugar product, therefore, it is feasible to presume a proportion of the contaminant drop equal to the proportion observed in the color decline at this stage. Consequently, the quantity of impurities is calculated as follows:

$$FIL\,Impurities = (Descolourisation\,Dry\,Matter - Water\,Dry\,Matter) * \\ (1 - \frac{X\%}{100}) * \frac{Decolourisation\,Imp.}{Decolourisation\,Dry\,Matter} \tag{B.15}$$

The remaining parameters are calculated as shown in the following table:

Table B.11: Final Liquor

Final Liquor		
Dry Matter	**Colour**	**Total Mass**
Sucrose+Impurities	Input	Equation B.12
	Sucrose	
	Equation B.13	
%Purity	**Impurities**	**Brix**
Sucrose/Dry Matter	Equation B.15	Input

Following the decolorization process, it becomes necessary to enhance the concentration of the solution, thereby reducing its water content and increasing its Brix value. This concentration step is carried out utilizing a double effect evaporator system. In the initial effect, the Final Liquor obtained from the decolorization stage is subjected to heat through a dedicated heating furnace employing steam as the heating medium. After passing through the evaporators, the final liquor will become a more concentrated solution, which is known as concentrated liquor(CL). As there is no addition of material, only the evaporation of water from the solution, the dry matter remains the same and the total mass is calculated by dividing the dry matter by the brix (again, a company-recorded average and thus one of the process inputs). As for the colours, they are barely altered by variations in temperature. The other parameters are determined by the calculations described above.

Subsequent to the process of concentration, it is necessary to crystallize the sucrose. In the crystallisation section, the sugar dissolved in the Concentrated Liquor and Refining Syrup is partially crystallized, resulting in the formation of a heated mass. This crystallization is processed in

Table B.12: Concentrated Liquor

Concentrated Liquor		
Dry Matter	**Colour**	**Total Mass**
FL Dry Matter	Input	Dry Matter/Brix
	Sucrose	
	FL Sucrose	
%Purity	**Impurities**	**Brix**
Sucrose/Dry Matter	FL Impurities	Input

Vacuum Pots, which are discontinuous crystallisers. In this apparatus, the liquor is concentrated to the point of oversaturation, where the Seed can be extracted. After the sugar crystals have formed, more Liquor or Syrup, depending on the type of sugar to be produced, is added to the pan until it is filled normally. At the conclusion, a cooked mass is acquired and stored in the cooked mass stirrers until centrifugation in the Refining Centrifuges.

Due to the addition of refining syrup (RS) with specific characteristics from the refining centrifuges (process explained below), the total mass of the solution in the vacuum pots will equal the aggregate of the masses of the two solutions, as well as the quantities of sucrose and dry matter. According to the formulas used in the preceding stages, the brix, impurities, and %purity will be calculated. The final hue will be determined by a weighted average of the two solutions.

Table B.13: Vaccum Pans

Vaccum Pans		
Dry Matter	**Colour**	**Total Mass**
CL Dry Matter+RS Dry Matter	Weighted Average Colour	CL Total Mass+RS Total Mass
	Sucrose	
	CL Sucrose+RS Sucrose	
%Purity	**Impurities**	**Brix**
Sucrose/Dry Matter	Dry Matter-Sucrose	Dry Matter/Total Mass

As derived from the vacuum pans, a solution referred to as refined cooked mass(RCM) is obtained. The process of crystallization involved in this phase entails an elevation in temperature, thereby resulting in a marginal modification in color. As the objective lies in maintaining a constant quantity of dry matter while exclusively facilitating sugar crystallization, it is evident that the quantities of dry matter and sucrose will correspond to those calculated for the solution derived from the vacuum pans. Concerning the parameter of Brix, since it is once again registered by the laboratory, it is considered an input. Consequently, we rely upon this parameter in conjunction with the quantity of dry matter to compute the total mass (by dividing the dry matter by the Brix value). Thus, the ensuing parameters for the solution subsequent to the vacuum pans are ascertained:

In the Refining Centrifuges, the Wet Sugar and the Refining Syrup are separated from the cooked dough arriving from the cooked dough mixer. This syrup is comprised of the liquid portion of the boiled mass, the water used to cleanse the sugar during centrifugation, as well as dilution water. The syrup is stored and piped to the Storage Deposits, where it is used to crystallize white or brown sugar. A portion of the nectar is periodically sent to the Recovery section. The Wet

Table B.14: Refined Cooked Mass

Refined Cooked Mass		
Dry Matter Vaccum Pans Dry Matter	**Colour** Weighted Average Colour	**Total Mass** Dry Matter/Brix
	Sucrose Vaccum Pans Sucrose	
%Purity Sucrose/Dry Matter	**Impurities** Dry Matter-Sucrose	**Brix** Input

Sugar from the Refining Centrifuges is conveyed to the Sugar Dryer via cat conveyor, lift, and belt conveyor.

As with Affination centrifuges, the parameters of these centrifuges remain constant, with the exception of the total mass and, consequently, the brix. The operation of the centrifuges in the refining process is analogous to that of affination centrifuges, as they both involve the addition of water. The calculation of the water quantity in the centrifuges is determined based on the following considerations: the total mass within the centrifuges is equal to the cumulative mass from the preceding step combined with the amount of washing water added. Furthermore, the average Brix value of the refining syrup is recorded as an input parameter. Iterative processes are subsequently employed to ensure that the simulated Brix value aligns with the registered Brix value. Therefore, the parameters for this stage of the process are the following:

Table B.15: Refining Centrifuges

Refining Centrifuges		
Dry Matter RCM Dry Matter	**Colour** RCM Colour	**Total Mass** RCM Total Mass+ Water Total Mass
	Sucrose RCM Sucrose	
%Purity RCM %Purity	**Impurities** RCM Impurities	**Brix** Dry Matter/Total Mass

This procedure results in refined syrup and wet sugar. Initially focusing on the wet white sugar (WWS), the laboratory at this point records the percentage of humidity in the solution, which is regarded as an input for the simulation. It is possible to calculate the brix from the humidity value by subtracting the registered number from 100. Regarding the quantity of dry matter, this is determined by multiplying the dry matter present in the previous phase by the yield recorded of 56%. As stated in the literature review, approximately 50%, however 56% is inferred based on historical records. The remaining parameters will be calculated as follows:

The sugar from the Refining Centrifuges is dried in two phases in a rotary dryer. The sugar is exposed to co-current heated air in the first portion of the dryer and co-current cold air in the final portion. Thus, the sugar is desiccated and cooled prior to being sorted, wrapped, packaged, or sent to silos for bulk filling. The air used to dry the sugar is filtered and heated with steam in a heat exchanger. Hot air is circulated through the dryer by fans at the dryer's entrance and departure. The frigid air is sucked into the dryer and passes through filters before entering the dryer's top. The suction is performed in the center (within the Dryer), causing the sugar to come into contact

Table B.16: Wet White Sugar

Wet White Sugar		
Dry Matter	**Colour**	**Total Mass**
$Centrifuges\ Dry\ Matter * \eta$	Input	Dry Matter/Brix
	Sucrose	
	Dry Matter*%Purity	
%Purity	**Impurities**	**Brix**
Input	RCM Impurities	100-%Humidity

with the chilly air. The air leaving the drier is passed through a cyclone, where sugar powder is separated by sprinkling hot sugar water. The surplus of fragrant water is refined.

For the remaining 2 phases of the procedure, the parameters color and %purity will remain unaltered, as impurities will be nearly eliminated and only the dry matter content will be altered. Regarding the rest of the parameters, as the only difference from the previous stage is the sugar humidity, it is possible to verify an increase in the brix (dehydration of the product at this stage) and a consequent decrease in the total mass (water evaporation).

Table B.17: Dry White Sugar (DWS)

Dry White Sugar		
Dry Matter	**Colour**	**Total Mass**
%Purity+Sucrose	WWS Color	Dry Matter/Brix
	Sucrose	
	WWS Sucrose	
%Purity	**Impurities**	**Brix**
WWS %Purity	WWS Impurities	100-%Humidity

In the sugar exit zone of the Dryer, there is a net whose function is to extract the agglomerated crystals (moulds) that are returned to the affination dissolver. The aforementioned network will generate sugar returns, resulting in a decrease in the solution's total mass. As the Company X has recalls for the quantity of returns, the total mass can be calculated as follows:

$$Final\ White\ Sugar\ Total\ Mass = DWS\ ToTal\ Mass - White\ Sugar\ Returns \tag{B.16}$$

Table B.18: Final White Sugar

Final White Sugar		
Dry Matter	**Colour**	**Total Mass**
Brix*Total Mass	DWS Colour	Equation B.16
	Sucrose	
	Dry Matter*%Purity	
%Purity	**Impurities**	**Brix**
DWS %Purity	RCM Impurities	DWS Brix

Appendix C

Mass Balance Simulation

This appendix contains Excel sheets illustrating the simulated patterns for the dissertation's discussed analysis. Figure C.1 illustrates the cadence for the year 2022 and provides valuable insight into the performance during that time period. Figure C.2 depicts the rhythm corresponding to an Overall Equipment Effectiveness (OEE) of 80%, allowing a comprehensive evaluation of the system's efficacy under such conditions. Figure C.3 depicts a scenario in which a three-putsch filter is acquired, highlighting the potential impact and benefits of this acquisition on the overall operations. These figures serve as visual aides to support the dissertation's findings and enhance comprehension of the topic.

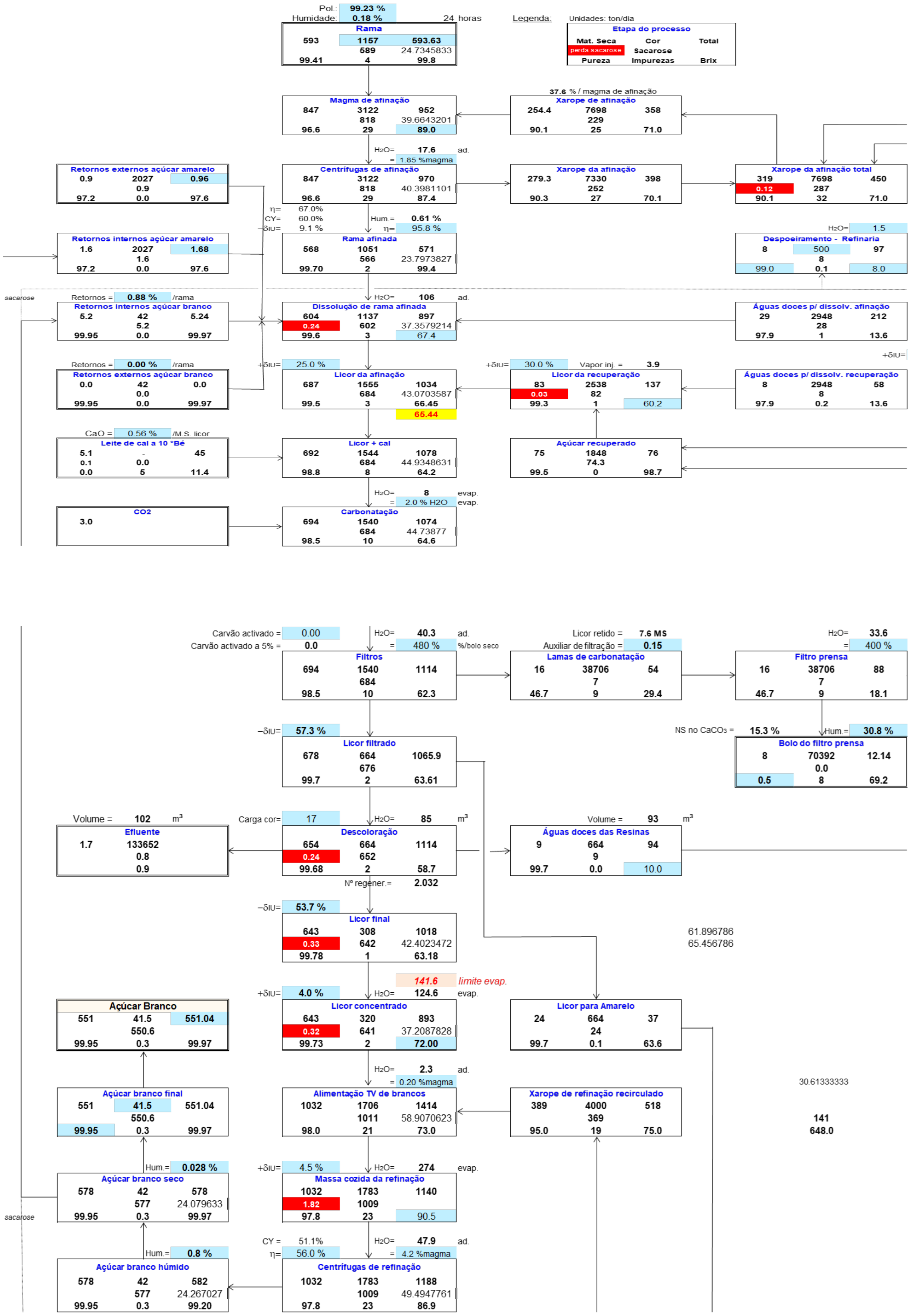

Figure C.1: Mass Balance simulation for 2022 rhythms of refinement

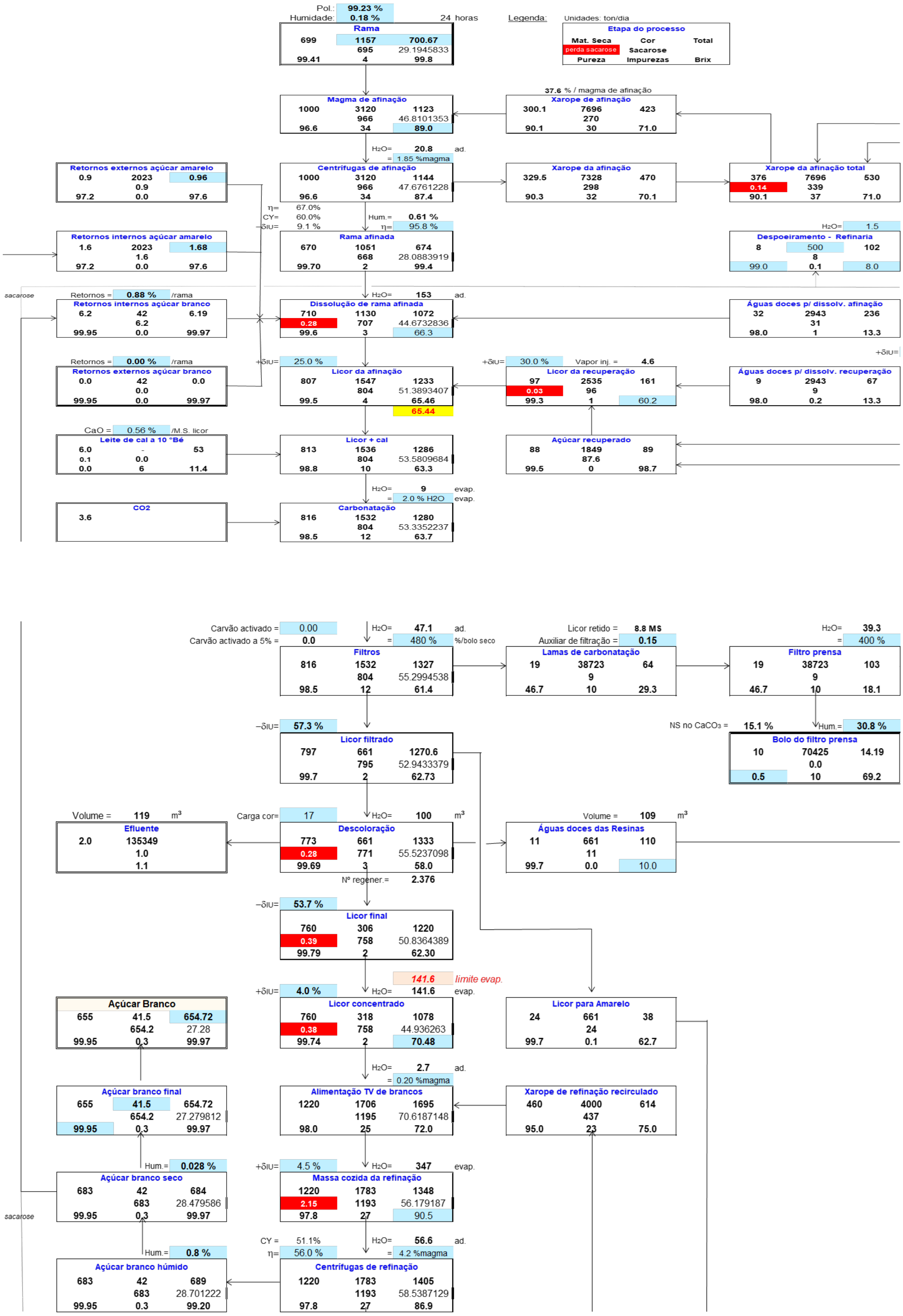

Figure C.2: Mass Balance simulation for an OEE of 80%

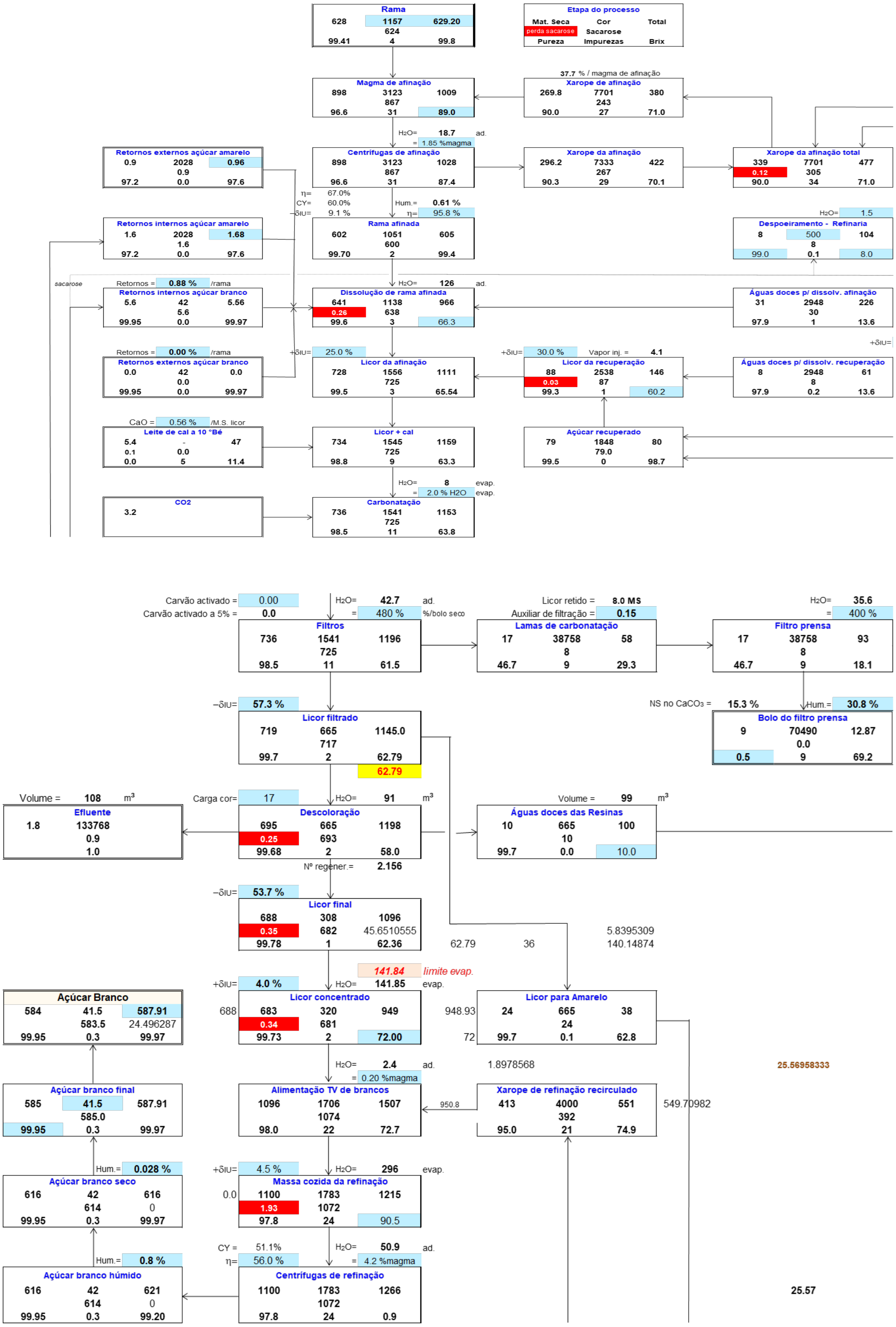

Figure C.3: Mass Balance simulation

Appendix D

Evaporator Labels

The supplied appendix contains the formulas used in the evaporator sizing calculations. Formula D.1 relates specifically to the mass balance, which is essential for determining the correct amounts and proportions of substances involved in the evaporation process. This formula ensures that the input and output mass flow rates are accounted for accurately, allowing for precise sizing calculations. Alternatively, Formula D.2 is related to the energy balance, which accounts for heat transfer and energy requirements during evaporation. This formula aids in the evaluation of energy inputs and outputs, ensuring that the evaporator operates efficiently and according to the intended parameters. These formulas play a crucial role in the meticulous calculation process, ensuring the evaporator system's accurate and reliable sizing.

$$\dot{m}_{v,0} + \dot{m}_{f,0} = \dot{m}_{v,n} + \dot{m}_{f,n} + \sum \dot{m}_{c,i} + \sum \dot{m}_{vb,i} \tag{D.1}$$

$\dot{m}_{v,0}$= Steam Input Mass in the Evaporator
$\dot{m}_{f,0}$= Liquor Input Mass in the Evaporator
$\dot{m}_{v,n}$= Steam Input Mass in the Effect n
$\dot{m}_{f,n}$= Liquor Outlet Mass in the n effect of the Evaporator
$\dot{m}_{c,i}$= Steam Input Mass in the Effect n
$\dot{m}_{vb,i}$= Vapour Outlet Mass in the i effect of the Evaporator

$$\dot{m}_{v,0} \cdot h_{v,0} + \dot{m}_{f,0} \cdot h_{f,0} = \dot{m}_{v,n} \cdot h_{v,n} + \dot{m}_{f,n} \cdot h_{f,n} + \sum \dot{m}_{c,i} \cdot h_{c,i} + \sum \dot{m}_{vb,i} \cdot h_{v,i} \tag{D.2}$$

By applying formula 2.9 and having an understanding of the input and output brix and the input mass flow rate, it is possible to calculate the amount of water evaporated:

$$\dot{m}_e = \dot{m}_{l,0} \cdot (1 - \frac{Brix\ Inlet}{Brix\ Outlet}) \tag{D.3}$$

$\dot{m}_e$= Total Mass of Evaporated Water

The formula 2.10 is used to calculate sensible heat, which is the heat responsible for the temperature rise to saturation and the formula 2.11 is used to calculate latent heat, which is the heat required to evaporate a solution.

$$\dot{Q} = \dot{m} \cdot c_p \cdot \Delta T \tag{D.4}$$

c_p= Specific Heat of the Solution

$$\dot{Q} = \dot{m}_{e,1} \cdot \Delta h_{lv} \tag{D.5}$$

$\dot{m}_{e,1}$= Total Mass of Evaporated Water in 1 effect
Δh_{lv}= Enthalpy of Vaporisation

TABLE VIII.

BOILING POINT ELEVATION OF SUGAR SOLUTIONS.

Per cent. Sugar. ° Brix.		*Boiling Point Elevation °C. at Vacuum of:*				
		Atmos.	15″	20″	25″	27″
10	...	·1	·1	·1	·1	·1
20	...	·3	·3	·3	·3	·2
30	...	·6	·5	·5	·5	·4
40	...	1·0	·9	·9	·8	·8
45	...	1·4	1·2	1·2	1·1	1·1
50	...	1·8	1·6	1·5	1·4	1·4
55	...	2·3	2·1	1·9	1·8	1·8
60	...	3·0	2·7	2·5	2·4	2·3
65	...	3·8	3·4	3·2	3·0	2·9
70	...	5·1	4·5	4·3	4·0	3·9
75	...	7·0	6·2	5·9	5·5	5·4
80	...	9·4	8·4	7·9	7·4	7·2
82	...	10·5	9·4	8·9	8·4	8·1
84	...	12·0	10·8	10·3	9·6	9·3
86	...	14·0	12·6	12·0	11·2	10·8
88	...	16·5	14·8	14·1	13·2	12·8
90	...	19·6	17·5	16·6	15·5	15·1
91	...	21·6	19·3	18·3	17·2	16·6
92	...	24·0	21·4	20·3	19·0	18·4
93	...	26·9	24·0	22·8	21·4	20·6
94	...	30·5	27·2	25·8	24·2	23·4

Figure D.1: Table of Boiling Point Elevation Values According to Brix and Pressure (Adapt from LYLE (1957))

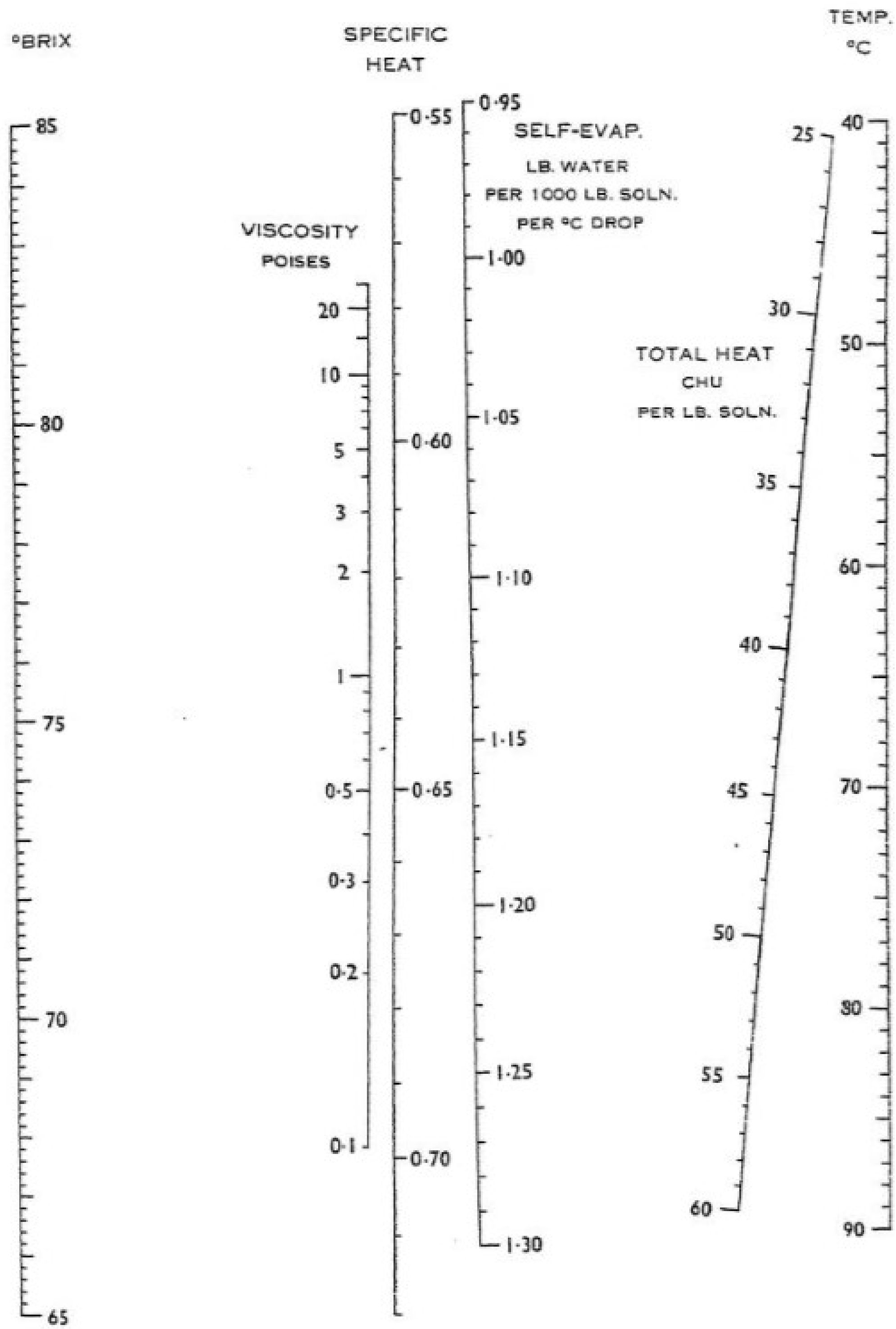

Figure D.2: Table of Viscosity, Self Evaporation Rate and Specific Heat According to Brix (Adapt from LYLE (1957))

Table D.1: Evaporator Thermodynamic Characteristics

	Exhaust Steam	**1st Effect**	**2nd Effect**
Pressure	2 bar	0.75 bar	0.24 bar
Temperature	120.80ºC	92.40ºC	64.00ºC
Latent Heat	525.4 kcal/kg	544.1 kcal/kg	561 kcal/kg
Boiling Point Elevation	-	3.20ºC	4.10ºC

Appendix E

Removing Outliers

Interquartile Range (IQR) is a statistical technique used to identify and eliminate data points that deviate considerably from the rest of the dataset when removing outliers. A statistical outlier is an observation that deviates significantly from the preponderance of data points and does not fit the overall pattern or distribution.

The interquartile range (IQR) is a measure of statistical dispersion and is defined as the difference between the first and third quartiles of a dataset. It represents the center fifty percent of the data and is resistant to extreme values. To identify anomalies using the IQR method, we perform the following procedures:

1. Sort the dataset by increasing value.
2. Determine the first and third quartiles (Q1 and Q3).
3. To determine the IQR, subtract Q1 from Q3.
4. Determine the minimum and maximum values for outliers- Q1: (1.5 * IQR) is the minimum acceptable value. Upper limit: Q3 plus (1.5 * IQR).
5. Outliers are any data points that fell below the lower bound or above the upper bound.
6. Remove or flag these outliers for further investigation or exclusion from future calculations.

By eliminating outliers, we can reduce their impact on statistical measures that can be skewed by extreme values, such as the mean, standard deviation, and correlation coefficients. Outliers can be caused by a variety of factors, including measurement errors, data entry mistakes, or truly uncommon events that differ from the majority of the observations.

However, caution must be exercised when removing outliers. Occasionally, outliers can provide valuable insights or indicate underlying data phenomena. Before completely eliminating outliers, it is essential to scrutinize their nature and comprehend the context of the data. In some instances, outliers may be valid and indicative of extreme or uncommon occurrences, and removing them could result in the loss of crucial data.

Due to its simplicity and reliability, the IQR method is a popular technique for detecting outliers. Nevertheless, alternative methods such as z-scores, modified z-scores, and other domain-specific approaches may also be used depending on the characteristics of the dataset and the objectives of the analysis.

Using the Interquartile Range to eliminate outliers enables us to identify and manage extreme values that deviate substantially from the majority of the data. By doing so, we can ensure that our statistical analyses and modeling efforts are founded on a more precise and representative data set, resulting in more trustworthy conclusions and insights.

Appendix F

Process Parameters

This appendix comprises a compilation of all of Company X's process records. Figure F.1 displays the quality analysis of the unrefined material consumed over the past three years for this collection. It provides a graphical representation of the variations in the quality parameters, providing valuable insight into the characteristics and trends of the source material. In addition, Figures F.2 and F.3 illustrate the tables utilized in the mass balance calculations. These tables are indispensable for accurately estimating mass flow rates and ensuring a comprehensive equilibrium throughout the process. Lastly, Figure F.4 displays the brix readings from previous years. This information is essential for characterizing the current state and permits a detailed analysis of the concentration levels, thereby facilitating comprehension of process performance. The inclusion of these numbers and documents in the appendix enhances the overall comprehension of Company X's operations and provides a firm foundation for further analysis and interpretation.

	Starch	Dextrano	TSS
Bona	125	544	12170
Alice Star 50% + Bona 50%	142	304	13040
Bona 75% + Regius 25%	149	527	13220
Alice Star	158	64	13910
Double Diamond	191	165	2630
Livadi	198	52	8920
Bona 50% + Global Harmony 50%	209	517	10920
Benjamin Confidence	212	711	11620
Regius	220	476	16380
Alice Star 50% + Global Harmony 50%	225	277	11790
Pelagiani	263	93	12380
Avellaneda 50% + Pelagiani 50%	271	342	30090
Giorgos 50% + Strandja 50%	272	345	11090
Avellaneda 67% + Pelagiani 33%	273	425	36000
Avellaneda 75% + Pelagiani 25%	274	467	38950
Avellaneda	278	591	47810
Avellaneda 33% + Santamaria 67%	289	548	23180
Global Harmony	292	489	9680
Santamaria	294	526	10870
Santa Virginia	360	151	2530
	234.75	380.7	16859

Figure F.1: Sugar Cane qualities

Pressure	Temp	volume (m^3/kg)		energy (kJ/kg)		enthalpy (kJ/kg)			entropy (kJ/kg.K)		
MPa	°C	vf	vg	uf	ug	hf	hfg	hg	sf	sfg	sg
0.001	6.97	0.00100	129.18	29.3	2384.5	29.3	2484.4	2513.7	0.1059	8.8690	8.9749
0.0012	9.65	0.00100	108.67	40.6	2388.2	40.6	2478.0	2518.6	0.1460	8.7622	8.9082
0.0014	11.97	0.00100	93.90	50.3	2391.3	50.3	2472.5	2522.8	0.1802	8.6720	8.8522
0.0016	14.01	0.00100	82.74	58.8	2394.1	58.8	2467.7	2526.5	0.2100	8.5935	8.8035
0.0018	15.84	0.00100	74.01	66.5	2396.6	66.5	2463.4	2529.9	0.2366	8.5242	8.7608
0.002	17.50	0.00100	66.99	73.4	2398.9	73.4	2459.5	2532.9	0.2606	8.4620	8.7226
0.003	24.08	0.00100	45.65	101.0	2407.9	101.0	2443.8	2544.8	0.3543	8.2221	8.5764
0.004	28.96	0.00100	34.79	121.4	2414.5	121.4	2432.3	2553.7	0.4224	8.0510	8.4734
0.006	36.16	0.00101	23.73	151.5	2424.2	151.5	2415.1	2566.6	0.5208	7.8082	8.3290
0.008	41.51	0.00101	18.10	173.8	2431.4	173.8	2402.4	2576.2	0.5925	7.6348	8.2273
0.01	45.81	0.00101	14.67	191.8	2437.2	191.8	2392.1	2583.9	0.6492	7.4996	8.1488
0.012	49.42	0.00101	12.36	206.9	2442.0	206.9	2383.4	2590.3	0.6963	7.3886	8.0849
0.014	52.55	0.00101	10.69	220.0	2446.1	220.0	2375.8	2595.8	0.7366	7.2945	8.0311
0.016	55.31	0.00102	9.431	231.6	2449.8	231.6	2369.0	2600.6	0.7720	7.2126	7.9846
0.018	57.80	0.00102	8.443	242.0	2453.0	242.0	2363.0	2605.0	0.8036	7.1401	7.9437
0.02	60.06	0.00102	7.648	251.4	2456.0	251.4	2357.5	2608.9	0.8320	7.0752	7.9072
0.03	69.10	0.00102	5.228	289.2	2467.7	289.3	2335.2	2624.5	0.9441	6.8234	7.7675
0.04	75.86	0.00103	3.993	317.6	2476.3	317.6	2318.5	2636.1	1.0261	6.6429	7.6690
0.06	85.93	0.00103	2.732	360.0	2489.0	359.9	2293.0	2652.9	1.1454	6.3857	7.5311
0.08	93.49	0.00104	2.087	391.6	2498.2	391.7	2273.5	2665.2	1.2330	6.2009	7.4339
0.1	99.61	0.00104	1.694	417.4	2505.6	417.5	2257.4	2674.9	1.3028	6.0560	7.3588
0.12	104.78	0.00105	1.428	439.2	2511.7	439.4	2243.7	2683.1	1.3609	5.9368	7.2977
0.14	109.29	0.00105	1.2366	458.3	2516.9	458.4	2231.6	2690.0	1.4110	5.8351	7.2461
0.16	113.30	0.00105	1.0914	475.2	2521.4	475.4	2220.6	2696.0	1.4551	5.7463	7.2014
0.18	116.91	0.00106	0.9775	490.5	2525.5	490.7	2210.7	2701.4	1.4945	5.6676	7.1621
0.2	120.21	0.00106	0.8857	504.5	2529.1	504.7	2201.5	2706.2	1.5302	5.5967	7.1269
0.3	133.52	0.00107	0.6058	561.1	2543.2	561.4	2163.5	2724.9	1.6717	5.3199	6.9916
0.4	143.61	0.00108	0.4624	604.2	2553.1	604.7	2133.4	2738.1	1.7765	5.1190	6.8955
0.6	158.83	0.00110	0.3156	669.7	2566.8	670.4	2085.7	2756.1	1.9308	4.8284	6.7592
0.8	170.41	0.00112	0.2403	720.0	2576.0	720.9	2047.4	2768.3	2.0457	4.6159	6.6616
1	179.88	0.00113	0.1944	761.4	2582.7	762.5	2014.6	2777.1	2.1381	4.4469	6.5850

Figure F.2: Steam Thermodinamic Table

Degree Beaume	Density (gm/ml)	Grams CaO/Lt	CaO Percentage by weight
1	1.007	7.5	0.75
2	1.014	16.5	1.65
3	1.022	26	2.55
4	1.037	36	3.50
5	1.045	46	4.43
6	1.052	56	5.36
7	1.060	65	6.18
8	1.067	75	7.08
9	1.075	84	7.87
10	1.083	94	8.74
11	1.091	104	9.60
12	1.1	115	10.54

Figure F.3: Baume Table

	Affination	Final	Concentrated
2020	65.91	61.95	70.25
Oct	65.64	61.07	69.27
Nov	66.31	62.13	70.37
Dec	65.66	62.54	71.05
2021	65.84	62.03	70.59
Jan	66.13	63.61	71.59
Feb	66.28	62.53	70.39
Mar	65.90	62.54	70.65
Apr	66.30	62.55	70.97
May	66.06	61.21	70.18
Jun	66.21	61.43	70.77
Jul	66.17	62.28	70.41
Aug	65.95	61.59	70.02
Sep	65.94	61.82	70.40
Oct	65.03	61.96	70.44
Nov	64.35	61.60	70.64
Dec	66.17	61.42	70.52
2022	64.91	61.37	69.90
Jan	66.35	62.29	70.86
Feb	63.89	61.33	70.12
Mar	65.69	61.55	70.64
Apr	64.60	60.53	69.67
May	64.99	60.26	69.50
Jun	64.85	60.52	68.90
Jul	62.35	61.70	69.39
Aug	64.28	61.97	69.57
Sep	65.61	61.64	70.40
Oct	65.46	61.58	69.91
Nov	65.56	61.91	70.09
Dec	65.37	60.98	69.49
2023	65.30	61.98	69.84

Figure F.4: Brix Table

Appendix G

Evaporation Calculations

This appendix contains the calculation documents utilized during the evaporator sizing procedure. The calculations conducted for the parameters associated with the sizing of the evaporator are illustrated in detail in Figure G.1. Figures G.2 and G.3 depict the calculations performed for a 715 ton flow rate. In the first scenario, it is evident that the required heat transfer exceeds the evaporator's capacity, indicating a performance limitation. Figure G.3 depicts the result of this non-ideal scenario by displaying the corresponding brix at the equipment's outlet. Figure G.4 demonstrates the necessary flow reduction required to obtain a brix output of 72º, thereby presenting a potential solution to meet the specified parameters. These calculation pages in the appendix provide crucial information and facilitate comprehension of the sizing procedure and its associated difficulties in relation to the evaporator.

DATA							
1 effect				**2 effect**			
Entrance		**Exit**		**Entrada**		**Saida**	
Licor				**Licor**			
Brix=	62 º	Brix=	65.79 º	Brix=	65.79 º	Brix=	72.00 º
Temp.=	79.3 ºC	Temp.=	95.7 ºC	Temp.=	95.7 ºC	Temp.=	68.00 ºC
m=	42.30 ton/h	m=	39.86 ton/h	m=	39.86 ton/h	m=	36.42 ton/h
d=	1.27 ton/m3	d=	1.27 ton/m3	d=	1.27 ton/m3	d=	1.34 ton/m3
V=	33.40 m3/h	V=	31.44 m3/h	V=	31.44 m3/h	V=	27.24 m3/h
Specific heat=	0.7			Specific heat=	0.7		
Vapour				**Vapour**			
P=	1.75 Kg/cm2a	P=	0.748 Kg/cm2a	P=	0.748 Kg/cm2a	P=	0.24 Kg/cm2a
Temp.=	120.8 ºC	Temp.=	92.4 ºC	Temp.=	92.4 ºC	Temp.=	64 ºC
LT=	525.4 Kcal/Kg	LT=	544.1 Kcal/Kg	LT=	544.1 Kcal/Kg	LT=	561 Kcal/Kg
U=	900 Kcals/m2/h/ºC			U=	700 Kcals/m2/h/ºC		
A=	90 m2			A=	90 m2		

Evap. Required	
Evap.=	5.87 ton/h

Flash	
Flash=	1.347047 ton/h

2 Effect Evaporation	
Evap.2=	2.229 ton/h

1 Effect Evaporation	
Evap.1=	2.30 ton/h
Heat from 79.3 to 95.7	
Heat=	485.58 Mcals
Heat to evap. 2.328 ton/h	
Heat=	1250.59 Mcals
Steam required	
Steam=	3.47 ton/h
Total Heat	
Heat=	1736.17 Mcals

Figure G.1: Evaporator Capacity

DATA							
1 effect				2 effect			
Entrance		Exit		Entrada		Saida	
Licor				Licor			
Brix=	61.5 º	Brix=	65.68 º	Brix=	65.68 º	Brix=	72.00 º
Temp.=	79.3 ºC	Temp.=	95.7 ºC	Temp.=	95.7 ºC	Temp.=	68.00 ºC
m=	53.57 ton/h	m=	50.17 ton/h	m=	50.17 ton/h	m=	45.76 ton/h
d=	1.26 ton/m3	d=	1.27 ton/m3	d=	1.27 ton/m3	d=	1.34 ton/m3
V=	42.40 m3/h	V=	39.59 m3/h	V=	39.59 m3/h	V=	34.22 m3/h
Specific heat=	0.7			Specific heat=	0.7		
Vapour				Vapour			
P=	1.75 Kg/cm2a	P=	0.748 Kg/cm2a	P=	0.748 Kg/cm2a	P=	0.24 Kg/cm2a
Temp.=	120.8 ºC	Temp.=	92.4 ºC	Temp.=	92.4 ºC	Temp.=	64 ºC
LT=	525.4 Kcal/Kg	LT=	544.1 Kcal/Kg	LT=	544.1 Kcal/Kg	LT=	561 Kcal/Kg
U=	900 Kcals/m2/h/ºC			U=	700 Kcals/m2/h/ºC		
A=	90 m2			A=	90 m2		

Evap. Required	
Evap.=	7.81 ton/h

Flash	
Flash=	1.695296 ton/h

2 Effect Evaporation	
Evap.2=	3.012 ton/h

1 Effect Evaporation	
Evap.1=	3.11 ton/h
Heat from 79.3 to 95.7	
Heat=	615.01 Mcals
Heat to evap. 2.328 ton/h	
Heat=	1689.65 Mcals
Steam required	
Steam=	4.61 ton/h
Total Heat	
Heat=	2304.66 Mcals

Figure G.2: Evaporator for a 715 ton sugar cane at a 72º brix at the exit

DATA							
1 effect				2 effect			
Entrance		Exit		Entrada		Saida	
Licor				Licor			
Brix=	61.5 º	Brix=	64.38 º	Brix=	64.38 º	Brix=	68.95 º
Temp.=	79.3 ºC	Temp.=	95.7 ºC	Temp.=	95.7 ºC	Temp.=	68.00 ºC
m=	53.57 ton/h	m=	51.18 ton/h	m=	51.18 ton/h	m=	47.78 ton/h
d=	1.26 ton/m3	d=	1.26 ton/m3	d=	1.26 ton/m3	d=	1.32 ton/m3
V=	42.40 m3/h	V=	40.63 m3/h	V=	40.63 m3/h	V=	36.25 m3/h
Specific heat=	0.7			Specific heat=	0.7		
Vapour				Vapour			
P=	1.75 Kg/cm2a	P=	0.748 Kg/cm2a	P=	0.748 Kg/cm2a	P=	0.24 Kg/cm2a
Temp.=	120.8 ºC	Temp.=	92.4 ºC	Temp.=	92.4 ºC	Temp.=	64 ºC
LT=	525.4 Kcal/Kg	LT=	544.1 Kcal/Kg	LT=	544.1 Kcal/Kg	LT=	561 Kcal/Kg
U=	900 Kcals/m2/h/ºC			U=	700 Kcals/m2/h/ºC		
A=	90 m2			A=	90 m2		

Evap. Required	
Evap.=	5.79 ton/h

Flash	
Flash=	1.729486 ton/h

2 Effect Evaporation	
Evap.2=	1.999 ton/h

1 Effect Evaporation	
Evap.1=	2.06 ton/h
Heat from 79.3 to 95.7	
Heat=	615.01 Mcals
Heat to evap. 2.328 ton/h	
Heat=	1121.32 Mcals
Steam required	
Steam=	3.47 ton/h
Total Heat	
Heat=	1736.32 Mcals

Figure G.3: Evaporator for a 715 ton sugar cane

DATA											
1 effect						**2 effect**					
Entrance			**Exit**			**Entrada**			**Saida**		
Licor						**Licor**					
Brix=	62.79	º	Brix=	66.31	º	Brix=	66.31	º	Brix=	72.00	º
Temp.=	79.3	ºC	Temp.=	95.7	ºC	Temp.=	95.7	ºC	Temp.=	68.00	ºC
m=	45.73	ton/h	m=	43.30	ton/h	m=	43.30	ton/h	m=	39.88	ton/h
d=	1.27	ton/m3	d=	1.27	ton/m3	d=	1.27	ton/m3	d=	1.34	ton/m3
V=	35.98	m3/h	V=	34.08	m3/h	V=	34.08	m3/h	V=	29.82	m3/h
Specific heat=	0.7					Specific heat=	0.7				
Vapour						**Vapour**					
P=	1.75	Kg/cm2a	P=	0.748	Kg/cm2a	P=	0.748	Kg/cm2a	P=	0.24	Kg/cm2a
Temp.=	120.8	ºC	Temp.=	92.4	ºC	Temp.=	92.4	ºC	Temp.=	64	ºC
LT=	525.4	Kcal/Kg	LT=	544.1	Kcal/Kg	LT=	544.1	Kcal/Kg	LT=	561	Kcal/Kg
U=	900	Kcals/m2/h/ºC				U=	700	Kcals/m2/h/ºC			
A=	90	m2				A=	90	m2			

Evap. Required		
Evap.=	5.85	ton/h

Flash		
Flash=	1.463318	ton/h

2 Effect Evaporation		
Evap.2=	2.159	ton/h

1 Effect Evaporation		
Evap.1=	2.23	ton/h
Heat from 79.3 to 95.7		
Heat=	524.93	Mcals
Heat to evap. 2.328 ton/h		
Heat=	1211.39	Mcals
Steam required		
Steam=	3.47	ton/h
Total Heat		
Heat=	1736.32	Mcals

Figure G.4: Evaporator for a 72º brix at the exit