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Using process modeling and simulation to determine the sustainability of a novel lactic acid biorefinery in Europe: Influence of process improvements, scale, energy source, and market conditions

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Abstract:
The biorefinery concept aims to produce multiple high-value products from bio-based feedstocks. One such product is lactic acid, which is used across several industries such as food, pharmaceuticals, cosmetics, and chemicals. Lactic acid yield is influenced by several factors, and improvements in performance are often first demonstrated at lab-scale, requiring prospective models which make assumptions about how the system will be optimized to understand the potential of a commercial-scale plant. This study assesses the economic and environmental sustainability of a proposed lactic acid biorefinery through the combination of process modelling, techno-economic assessment, and life cycle assessment. The case study proposes producing high purity lactic acid from industrial candy waste and liquid digestate in Denmark through lactic acid fermentation and downstream membrane separations. A base case is first modeled to determine the economic and environmental hotspots of the system; scenarios are then modelled where improvement methods are implemented reducing consumption of energy, chemicals, water, and the production of waste, upscaling the system, and integrating bioenergy. Finally, a cost sensitivity analysis is run varying bioenergy, water, chemical, and labor costs based on the European market. By optimizing unit processes, scale, energy sources and market conditions, the lactic acid unit production cost and global warming potential can be lowered by 94% and 80% compared to the base case, respectively. However, these are optimized in different scenarios, indicating a trade-off in optimizing economic and environmental criteria. As even best-case lactic acid unit production cost is found to be 141-232% higher than market price, this production pathway will require further improvement or market changes before it can be commercially viable.

Keywords: lactic acid, biorefinery, process modeling, scale-up, techno-economic assessment, life cycle assessment

1. Introduction

Biorefineries are both real world deployments and conceptual systems which transform biomass into several valuable products such as food, feed, chemicals, materials, and energy (European

Commission et al., 2021). Biorefinery systems aim to challenge the function of conventional fossil-based refineries, which currently produce the majority of the world's fuels and plastics (Annevelink et al., 2022). While many biorefinery systems have been proposed over the last 10 years in academic and research and development settings, there are still major barriers for turning academic interest into widespread adoption. The main non-technical barrier to deployment of novel biorefineries is due to economic viability, particularly in Europe where there are high energy and labor costs and higher taxes compared to other parts of the world (European Commission et al., 2021). Furthermore, while environmental sustainability is one of the main drivers of this technology, the lack of evidence on the full-life cycle sustainability benefits of biorefinery products has been mentioned as another barrier (OECD, 2018). It is therefore necessary for novel biorefineries to quantitatively prove their economic and environmental sustainability. To evaluate the economic performance of biorefineries, techno-economic assessment (TEA) is commonly used, where market competitiveness is evaluated based on the total production costs calculated for a specified product or products (Humbird et al., 2011). For the evaluation of the environmental impact of biorefineries, life cycle assessment (LCA) is a particularly useful tool, which evaluates the impacts of a product, process or system over its entire life cycle (UNEP/SETAC Life Cycle Initiative, 2011).

One interesting product which can be targeted within a biorefinery system is lactic acid, which has many applications in the food, pharmaceutical, and cosmetic industries (Rodrigues et al., 2017). Lactic acid can be produced through either chemical synthesis or biological fermentation, with fermentation being dominant in commercial production (Joglekar et al., 2006). Lactic acid is biologically produced through the use of bacteria, fungi, yeast, algae or cyanobacteria (Rodrigues et al., 2017). The most used microorganism for the commercial production of lactic acid is *Lactobacillus* sp., a type of bacteria which produces lactic acid as the main product during the fermentation of carbohydrates (de Oliveira et al., 2017). These lactic acid bacteria (LAB) are preferred as they can produce high yields of lactic acid compared to other lactic acid-producing organisms (Manandhar, 2019). LAB need two main things to grow: a carbohydrate source and a nutrient source (Rodrigues et al., 2017). The carbohydrate (sugar) source is a significant factor in the economic performance, as direct sources such as glucose are expensive, while indirect sources such as starches or lignocellulosic biomass require expensive pretreatments such as acid or enzymatic hydrolysis (Ögmundarson et al., 2020). Furthermore, the nutrient source can be a particularly expensive component of production costs (de Oliveira et al., 2017).

Downstream separation and purification is also a critical aspect of lactic acid production, contributing up to 70% of operating and capital costs (Komesu et al., 2017). The conventional downstream process for lactic acid recovery includes lactic acid neutralization with calcium carbonate or calcium hydroxide and acidification with hydrogen sulfide followed by filtration, evaporation, and reactive distillation (Manandhar, 2019; Munagala et al., 2021; Ögmundarson et al., 2020). However, this method produces significant quantities of waste gypsum, which has been highlighted to have a significant economic and environmental impact (Joglekar et al., 2006). Alternative technologies which can be used for separation and purification include membrane technologies such as reverse osmosis, electrodialysis, microfiltration, ultrafiltration, nanofiltration, membrane bioreactors (Alves De Oliveira et al., 2018). Specifically, membrane filtration followed by electrodialysis has been suggested as a method of eliminating the produced salt while also recovering the neutralizing base for reuse (Joglekar et al., 2006).

Previous experimental work by some of the authors explored alternative carbohydrate and nutrient sources for lactic acid production such as candy factory waste, seaweed, grass silage, and digestate (Papadopoulou, 2022; Papadopoulou et al., 2023c). Based on this research, candy factory waste was found to be a good carbohydrate source due to a high sugar content without any hydrolysis needed (279 g/L), whereas digestate was found to be a sufficient low-cost alternative to conventional high-cost nutrient media. Together, a lactic acid yield of 60-70 g/L was achieved. The fermentation broth was subsequently–processed using membrane separation technologies, where experiments were conducted to optimize lactic acid recovery from two downstream processing pathways: microfiltration to nanofiltration (pathway A) or microfiltration to monopolar electrodialysis (pathway B) (Papadopoulou et al., 2023a). It was found that pathway B resulted in a higher lactic acid recovery, while pathway A was less energy demanding.

While the results achieved from experimental trials can be promising, commercial-scale operations can differ significantly from the experimental set-up. It is therefore important to consider the performance of novel production pathways at larger scales to determine whether such as biorefinery system is worth exploring commercially. Process modeling is an important tool at this stage, where designs can be developed using simulators which can predict equipment sizes, energy consumption, and economic costs (Mailaram et al., 2023; Pachón et al., 2020; Phanthumchinda et al., 2018). Furthermore, process models can be used to identify parameters and conditions which have the largest influence on the performance of the system; the process can then be 'optimized' by adjusting these parameters or taking future actions in order to maximize the goals of the project. The use of process modelling and simulation to determine the economic and environmental performance of lactic acid biorefineries is not new. Multiple studies exist which use process modelling to explore the economic, environmental, and both environmental and economic impacts of lactic acid biorefineries (Table 1). However, only some of these studies have considered lactic acid production within Europe, and none have considered candy waste and liquid digestate as feedstocks. Furthermore, while several process models have considered the use of membrane processes for the downstream separation and purification of lactic acid (Kwan et al., 2018; Liu et al., 2015; Peinemann et al., 2019; Phanthumchinda et al., 2018), no study has considered the environmental performance of using such technologies.

This study aims to determine the economic and environmental impacts of a developing pathway for lactic acid production within the European context. The lactic acid production pathway modelled in this study builds on previous research (Papadopoulou et al., 2023b, 2023c, 2023a) considering candy waste and liquid digestate as feedstocks and membrane technologies for downstream separation. To the best of the authors' knowledge, this is the first study determining the techno-economic and environmental life cycle assessment of lactic acid production from a novel feedstock of candy waste and digestate. Furthermore, this is the first environmental life cycle assessment of lactic acid production which considers the use of membrane technologies for downstream separation. Finally, this study focuses on the methodological question of how process optimization is defined and influenced, taking into special consideration the influence of process improvements, scale, energy sources, and market dynamics on the long-term sustainability of the system.

Table 1: Overview of literature using process modelling and simulation to assess the economic and/or environmental performance of lactic acid production and this study.

Study	**Type of assessment**	**Region of production**	**Feedstocks**	**Downstream separation and purification**
(Liu et al., 2015)	Economic	China	Corn stover	Membrane technologies
(Phanthumchinda et al., 2018)	Economic	Thailand	Pure nutrient media	Membrane technologies
(Méndez-Alva et al., 2018)	Economic	Unspecified	Sugarcane bagasse	Conventional downstream processes
(Kwan et al., 2018)	Economic	China (Hong Kong)	Food waste	Membrane technologies
(Peinemann et al., 2019)	Economic	Germany	Food waste	Membrane technologies
(Grasa et al., 2021)	Economic	USA, China, Germany	Seaweed	Conventional downstream processes
(Mailaram et al., 2023)	Economic	UK	Bread waste	Conventional downstream processes
(Adom and Dunn, 2017)	Environmental	USA	Pure glucose Corn stover	Conventional downstream processes
(Daful et al., 2016)	Environmental	South Africa	Sugarcane bagasse	Conventional downstream processes
(Ögmundarson et al., 2020)	Environmental	USA, China, Iceland	Corn Corn stover Seaweed	Conventional downstream processes
(Pachón et al., 2020)	Environmental	Switzerland	Vine shoots	Conventional downstream processes
(Mediboyina et al., 2024)	Environmental	Ireland	Whey and de-lactose permeate	Conventional downstream processes
(Daful and Görgens, 2017)	Economic and environmental	South Africa	Sugarcane bagasse	Conventional downstream processes
(Manandhar, 2019)	Economic and environmental	USA	Corn Corn stover Miscanthus	Conventional downstream processes
(Munagala et al., 2021)	Economic and environmental	India	Sugarcane bagasse	Conventional downstream processes
This study	**Economic and environmental**	**Denmark, Europe**	**Candy waste and digestate mixture**	**Membrane technologies**

2. Materials and methods

2.1. Process model development

This section will briefly discuss the experimental work done and data collected for this study, followed by the development of the base case process model.

2.1.1. Experimental work

Fermentations were conducted at the Chemical and Biochemical Engineering Department, Technical University of Denmark (DTU), Kgs. Lyngby, Denmark. In the experimental fermentations, the seed culture (inoculum) was prepared using LAB strain *Lactobacillus plantarum* and 50 ml De Man, Rogosa, and Sharpe (MRS) media (De MAN et al., 1960), incubated for 12 hours at pH 6.5 and 37°C (Papadopoulou et al., 2023c). The LAB strain was previously isolated from the seaweed species *Alaria esculenta* and stored in 20% glycerol stocks at -80°C (Papadopoulou et al., 2023b). Candy waste was obtained by Trolli Iberica A/S (Paterna, Spain) and stored at −20 ◦C. Digestate was collected from Hashøj biogas plant (Dalmose, Denmark), and consisted of 80% animal manure and 20% food or industrial waste. The digestate was manually sieved, diluted 1:1 with distilled water, and sterilized by autoclaving at 121°C for 15 min. The sterilized digestate, candy waste and 5% (v/v) inoculum were then added under sterile conditions to 5 L fermentation bioreactors (BioBench, Biostream International BV, Doetinchem, The Netherlands), and the system was initially flushed with N_2 to create anaerobic conditions. The bioreactors were operated for 48 hours at pH 6.55 (through the addition of NaOH) and 37°C, stirring at 100 rpm. The input volumes and final yields are displayed in Table 2. As NaOH consumption was not measured in the experimental trials, it was estimated based on ion concentrations measured in the downstream processes. An addition of 15 g/l NaOH was calculated, which agrees with that found in literature (Xie et al., 2019).

Table 2: Inputs and outputs to the experimental fermentation process (5 L bioreactor).

Input	Function	Amount	Output	Concentration
Candy waste	Carbon source	2 L	Maltose	40 ± 2.0 g/l
Liquid digestate	Nutrient source	1.5 L	Glucose	0.5 ± 0.5 g/l
Distilled water	Dilution	1.5 L	Lactic acid	65 ± 5.0 g/l
Inoculum	Inoculum	0.25 L	Acetic acid	1.5 ± 0.5 g/l
NaOH (50%)	pH control	150 g	Succinic acid	0.5 ± 0.5 g/l

The fermentation broth was centrifuged and then transported to Technische Universität Wien (TU Wien), Vienna, Austria for downstream processing. Two downstream processing pathways were initially considered (Figures S1-S2, Supplementary information). Pathway A considered microfiltration (MF) followed by nanofiltration (NF) and pathway B considered MF followed by monopolar electrodialysis (MPED). Several experimental runs were first conducted to optimize the processing pathways (Cabrera-González et al., 2022; Papadopoulou et al., 2023a), where it was found that NF was most effective when the temperature was set to 25°C and the pH was controlled to 2.8, while MPED was most effective at a pH of 4.0. It was also found that additional purification would be needed to remove the ions and salts generated from acidification; thus, a final bipolar electrodialysis (BPED) step was added to each pathway. The composition of each input and output stream was determined from high performance liquid chromatography (HPLC) and ion chromatography results (Tables S1-S2, Supplementary information), where it was found that even after BPED significant quantities of ions were leftover in processing pathway B. Pathway B would therefore require additional ion removal, either through ion exchange or ion chromatography. Another method could be a final NF step, as it seemed to work well for removing ions in pathway A. However, this was not explored experimentally for this study and thus there is no experimental data available for the achievable yields for this process. Furthermore, the economic costs are lowest when a minimum number of downstream processing steps are needed (Ecker, 2012; Phanthumchinda et al., 2018). Thus, pathway A was chosen for process modeling.

2.1.2. Process modeling of base case scenario

The chosen experimental pathway (pathway A) produced a final lactic yield of 23.45 g/L and purity of 91.32% considering the presence of 1.21 g/L acetic acid, 0.30 g/L succinic acid, and 0.72 g/L ions. Using the yields determined from each unit process, a base case scenario is modeled using Aspen Plus software (Aspen Technologies, Inc., USA). The base case for process simulation is set as a pilot plant using a total of 1000 tonnes per year (tpy) fermentation substrate, corresponding to 400 tpy of candy waste, 300 tpy of liquid digestate, and 300 tpy of dilution water. Aside from the final purified lactic acid flow, all output streams are considered as wastewater to be treated. In transitioning from a lab scale to pilot scale operation, several changes should be assumed. First, it is assumed that a continuous process would be designed for all scenarios, as continuous flow fermentations correspond to higher annual yields (due to less downtime) and smaller vessel volumes (Peinemann et al., 2019). Second, all manual operations will be assumed to be replaced with automated processes; while this will increase capital cost due to additional storage tanks, piping, and pumping, it reduces the labor intensity of the process, which is a high source of operating cost in developed countries (Kwan et al., 2018). Finally, as commercial lactic acid should be 25-90 wt.% of the solution (Komesu et al., 2017), a final concentration step is needed. The process flow diagram and mass balance for the base case can be seen in Figure 1. Additional details on the modeling can be found in the Supplementary information.

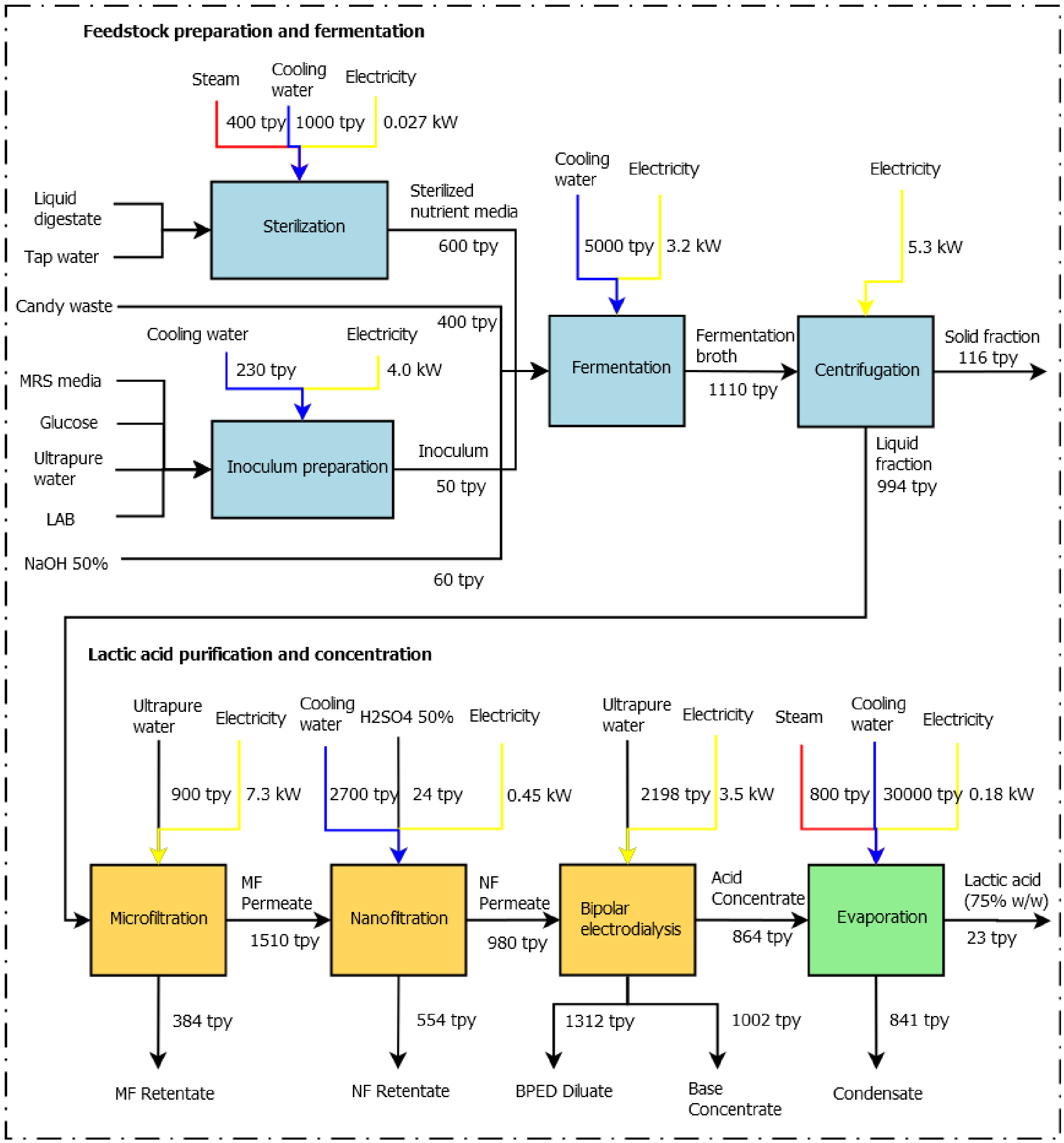


Figure 1: Process flow diagram and mass balance in tonnes per year (tpy) for base case scenario.

For process heating, process steam (2.5 bar) is used for sterilization and concentration, and electricity is used for heating the MF input to 60°C. As lactic acid fermentation is exothermic, process cooling is needed to maintain reactor temperature at 37°C. The fermenter is therefore assumed to be cooled with seawater at 15°C (Falkengaard and Pedersen, 2022). In the experimental work, cooling of the MF permeate prior to NF was done through natural convection; water cooling of NF input is assumed in the process model. Based on international regulations (Danish Government, 2023), the maximum allowable cooling water discharge temperature is 37.8°C.

2.2. Process optimization

2.2.1. Defining an optimal system

It is first important to highlight what is meant by an optimal system. Process optimization is a method of exploring changes in design and operating variables in order to minimize or maximize an objective function according to a set goal (Chaves et al., 2016). In the field of chemical engineering, the set goal is often to maximize profitability through minimizing economic costs while maximizing productivity. However, processes can also be optimized based on other goals. For instance, in the case of studies with environmental sustainability as a clear objective, a process can be optimized based on environmental objectives such as minimizing greenhouse gas (GHG) emissions (Murphy et al., 2016). If more than one goal is set, for example, minimizing economic cost while also minimizing environmental impact, it is often the case that optimization of one goal might compromise the optimization of the other (Bjørn et al., 2018). In this study, two optimization objectives are set: minimize the 1) economic costs and 2) environmental impacts associated with the biorefinery.

As can be seen in Table 3, in this study several improvement methods will be applied and modelled through individual scenarios. These methods are described in the following sections. The purpose of modelling these methods individually is to understand the impact of each proposed change on the sustainability of the process. These changes will later be combined to understand the influence of improvement methods on each other, and finally, to determine an optimal configuration for the biorefinery.

Table 3: Initial scenarios and improvement methods considered.

	Process improvements (PI) applied					Process upscaling (PU) applied (tpy feedstock inputs)			Bioenergy integration (BI) applied		
Scenario names	Triple effect evaporation	Process heat integration	Inoculum recycling	Seed fermentation	Water recycling	10,000	50,000	100,000	Woodchips boiler	Biogas CHP w/o subsidy	Biogas CHP w/ subsidy
Base case	–	–	–	–	–	–	–	–	–	–	–
PI.1	✓	–	–	–	–	–	–	–	–	–	–
PI.2	–	✓	–	–	–	–	–	–	–	–	–
PI.3	–	–	✓	–	–	–	–	–	–	–	–
PI.4	–	–	–	✓	–	–	–	–	–	–	–
PI.5	–	–	–	–	✓	–	–	–	–	–	–
PI.C1	✓	–	✓	–	✓	–	–	–	–	–	–
PI.C2	–	✓	✓	–	✓	–	–	–	–	–	–
PI.C3	✓	–	–	✓	✓	–	–	–	–	–	–
PI.C4	–	✓	–	✓	✓	–	–	–	–	–	–
PU.1	–	–	–	–	–	✓	–	–	–	–	–
PU.2	–	–	–	–	–	–	✓	–	–	–	–
PU.3	–	–	–	–	–	–	–	✓	–	–	–
BI.1	–	–	–	–	–	–	–	–	✓	–	–
BI.2a	–	–	–	–	–	–	–	–	–	✓	–
BI.2b	–	–	–	–	–	–	–	–	–	–	✓

✓ *indicates inclusion of optimization parameters;*

– *indicates no inclusion of optimization parameters*

2.2.2. Improvement method 1: Process improvements (PI) to reduce resource consumption and waste production

To reduce the economic costs and environmental impacts associated with the biorefinery, it is critical to reduce the raw material and utility consumption, i.e., to reduce the number of resources consumed and waste produced. Process improvements aimed at achieving this goal are expected to decrease the operating costs of the system but could also increase the capital costs as well as the complexity of the system, which will increase the operating labor and maintenance necessary. Thus, the trade-offs should be explored for each option. More details for each process improvement method can be found in the Supplementary information.

2.2.2.1. Single to triple effect evaporation (PI.1)

The first proposed improvement is to upgrade the evaporation from single- to triple-effect. In multi-effect evaporation, the first stage is heated with steam, while subsequent stages are heated with the vapor outlet of the previous stage. This reduces both the steam input and final cooling water needed (Cha et al., 2022). For this study, upgrading from single- to triple- effect evaporation leads to a reduction in steam and cooling water input for concentration of 57% and 50%, respectively. However, it also requires two additional evaporation columns and heat exchangers.

2.2.2.2. Process heat integration (PI.2)

It has been shown that many plants use 50% more energy than necessary and can significantly reduce energy and water inputs for heating and cooling through process heat integration (Klemeš et al., 2023). Consequently, a pinch analysis and heat integration model is proposed. Process heat integration involves finding and utilizing wasted heating and cooling streams within a system. The waste heat from sterilization is used to heat the fermenter inputs to 37°C and the waste heat from concentration is used to heat the MF input to 60°C. Similarly, the cooling water is fully utilized to cool the fermenter and the NF input to 25°C. Heat integration reduces the steam, electricity, and cooling water input of the overall plant by 27%, 52%, and 32%, respectively. Furthermore, some capital equipment can be downsized.

2.2.2.3. Inoculum recycling (PI.3)

Other studies have determined separated biomass (LAB) reuse as inoculum to be a promising process improvement (Sikder et al., 2012). The impact of this improvement is therefore also tested. The output from the centrifuge contains 90% of the microbial biomass, which is assumed to be used to directly replace inoculum input to the fermentation stage. An 80% inoculum recycling rate is applied as this rate is found not to significantly compromise the quality of the final lactic acid purity.

2.2.2.4. Seed fermentations (PI.4)

Seed fermentations involve the use of smaller scale fermenters to increase the inoculum volume prior to the main fermentation, thereby reducing the amount of lab-scale inoculum preparation needed (Kwan et al., 2018). To apply this process improvement in the base case, two seed fermentation steps are used to increase the inoculum from 0.025 kg/h to the 5.7 kg/h needed for fermentation. This reduces the inoculum demand by >99%. However, seed fermentation also requires additional capital equipment as well as additional feedstock inputs, electricity, and cooling.

2.2.2.5. Water recycling (PI.5)

According to the experimental results from this study, the diluate and base concentrate from BPED and the condensate from the concentration step are 99.2%, 99.7%, and 99.6% water, respectively. This makes them good candidates for recycling as dilution water streams, which may improve the sustainability of the proposed biorefinery. To test this, the streams are assumed to be combined and then used as inputs to the MF and BPED steps. An 80% water recycling rate is applied as this rate is found not to significantly compromise the quality of the final lactic acid purity.

2.2.2.6. Combined process improvement scenarios (PI.C1-PI.C4)

Finally, it is of interest to see how combining several process improvement scenarios could further reduce economic and environmental costs, and whether certain combinations are more effective than others. Of the five process improvement scenarios, two target reduction of energy inputs (triple effect evaporation and process heat integration), two target reduction of inoculum inputs (inoculum recycling and seed fermentations), and the final one targets reduction of water inputs and wastewater outputs (water recycling). Four combinations are therefore established combining one energy reduction process improvement, one inoculum reduction process improvement, and one water reduction process improvement.

2.2.3. Improvement method 2: Process upscaling (PU) to reduce capital costs and energy demand

Scaling has been shown to have a large impact on the economic and environmental costs of a process, particularly with respect to capital equipment and energy demand (Mainardis et al., 2021). However, there also exists diseconomies of scale, where the scale is so large that local feedstock supply is unavailable, requiring the import of such from farther locations (Manandhar, 2019). There is therefore a balance between increasing scale to minimize capital equipment and energy costs while maintaining a realistic feedstock supply.

The two main feedstocks used in this biorefinery are candy waste and liquid digestate, both of which are in abundant supply within Europe. European countries are some of the world's largest consumers of sugary foods: Belgium, Poland, Malta, and Denmark are all reported to have an annual sugar consumption of over 40 kg per capita, deriving from the consumption of processed foods such as confectionaries, baked goods and fizzy beverages (World Population Review, 2023). The production of these confectionaries generates a large volume of sugar-rich (~300 g/L) wastewater, where factories producing confectionaries have been found to produce approximately 300-500 m3 of wastewater per month, or 3,600-6,000 m3/year (Zajda and Aleksander-Kwaterczak, 2019). Digestate is produced by biogas plants, of which there are approximately 20,000 in Europe. These plants vary in capacity and type but produce on average 610 kW of electricity (European Biogas Association, 2020). In 2020, the largest biogas producers in the EU were Germany, UK, Italy, France, Czechia, Denmark, and the Netherlands. A 500 kW biogas plant produces more than 10,000 tpy of digestate (Wiśniewski et al., 2018). Typically, solid digestate is used as fertilizer; to reduce the weight, the liquid fraction is normally removed via centrifugation (Mainardis et al., 2021). Assuming a liquid fraction of 80-90% (Kisielewska et al., 2022), that is 8,000-9,000 m3 of wasted liquid digestate per year.

Based on these metrics, three commercial scale scenarios are assessed. In the first (PU.1), it is assumed that all candy waste is supplied by one factory and all digestate is supplied by one biogas plant, corresponding to 4,000 tpy candy waste, 3,000 tpy liquid digestate, and 3,000 tpy dilution water for a total of 10,000 tpy feedstock. In the second (PU.2), it is assumed that 5 candy factories supply the candy waste and 2 biogas plants supply the liquid digestate for a total of 50,000 tpy feedstock. The final scale tested (PU.3) corresponds to 10 candy factories and 4 biogas plants for a total of 100,000 tpy feedstock. The spatial density of candy factories and biogas plants are determined from real Danish candy production plants and biogas plants (Table S12, Supplementary information).

2.2.4. Improvement method 3: Bioenergy integration (BI) to reduce environmental impacts

A large contributor to environmental impact is the energy source. The Danish energy market is rapidly transitioning towards low-carbon sources, with the electricity production from fossil fuels dropping from 97% in 1990 to 16% in 2020 (IEA Bioenergy, 2022a) and heat production from fossil fuels dropping from 79% in 1990 to 17% in 2020 (IEA Bioenergy, 2022b). Biomass and waste are currently the largest sources of heat in Denmark while wind power and biomass make up the largest proportion of electricity generation. It is therefore reasonable to assume that the electricity and heating demands of lactic acid production plant could be supplied by bioenergy. For this study, two alternative energy scenarios will be applied. The first (BI.1) will assume that steam is generated on-site from a boiler using 100% woodchips. The second (BI.2) will assume that steam and electricity are supplied from a combined heat and power (CHP) plant using biogas.

While the environmental impact of the alternative energy scenarios is easily implemented through background data available in the ecoinvent database (see section 3.3.2), the economic impact of switching energy suppliers is less straightforward. Producing steam from a woodchips boiler has been found to be 23% more expensive than steam produced by a natural gas boiler (Jeswani et al., 2020), and biogas production through anaerobic digestion has been found to be over 100% more expensive than the natural gas market price (Jacobsen et al., 2014). On the other hand, there are several economic policies in place which benefit bioenergy, such as increasing carbon taxes on high-emitting energy sources and government support and subsidies for bioenergy (Kost et al., 2021). In Denmark, biogas subsidies have been relatively high, at approximately 56 €/MWh if used for CHP (Al Seadi, 2019). However, these subsidies are not guaranteed, with the aforementioned biogas supports only available for plants commissioned before 2020 (Danish Energy Agency, 2023). To understand how these economic policies impact bioenergy integration, two sub-scenarios will be considered for biogas use, excluding (BI.2a) and including (B1.2b) subsidies.

2.3. Sustainability assessment

Sustainability is assessed through the use of the LCA methodology (ISO, 2006a, 2006b). The goal is to calculate the economic and environmental impacts of producing 1 kg lactic acid (LA) solution for the base case and considering several improvement scenarios to determine the optimal biorefinery configuration. The system boundaries include all raw material (feedstocks, chemicals) and utility (electricity, steam, water, waste) flows as can be seen in Figure 1 as well as capital equipment and labour. The final use and disposal of the produced lactic acid is not considered; thus, it is a cradle-to-gate LCA. Economic impacts will be assessed through a TEA calculating the total annualized unit production cost, while environmental impacts will be assessed through an environmental LCA calculating the total global warming potential (GWP).

2.3.1. Techno-economic assessment (TEA)

To conduct the TEA, Aspen Plus Economic Analyzer (Aspen Technologies, Inc., USA) is used. This software estimates the capital and operating costs of a process based on the process units contained in an Aspen Plus model. As the Aspen Plus inventory does not include equipment blocks for newer technologies such as MF, NF, and ED, a placeholder separator was used for each of these process units in the Aspen Plus model. Capital cost for each of these units were later found based on literature (Åkerberg and Zacchi, 2000; Giordano et al., 2017; González et al., 2007; Sikder et al., 2012; Tejayadi and Cheryan, 1995; Vineyard et al., 2020) (Table S11, Supplementary information). It was also found that the Aspen Plus inventory does not vary centrifuge capital costs with scale, so this cost was also updated from literature (C1D1 Labs, 2023). All costs were converted to USD using a conversion rate of 7.69 DKK to 1 € and 0.91 € to 1 USD. Total capital cost was then estimated based on total equipment cost and the additional costs listed in Table 4.

To calculate the operating costs, the annual costs of raw materials, chemicals, utilities, and labor should first be determined. Raw materials, chemical, and water inputs were determined from the Aspen Plus process model. Electricity inputs for pumping, heating, fermenter stirring, and BPED were determined in this study as well as from literature (Junker, 2004; Szepessy and Thorwid, 2018). Labor was determined based on the methodology outlined by Verret et al. (2020). However, it was found that this calculation was not representative for a pilot scale operation; thus, for the base case, operating labor costs were assumed based on 4 full-time equivalent operators. The inputs needed per year were then multiplied by the price per unit (Table 5). All prices were taken from the Danish market perspective when data was available; if not, European or global prices are used. Inoculum cost was calculated based on cultivation media, electricity, water, and labor inputs at lab-scale (Table S13, Supplementary information). As waste streams, it is assumed that the candy waste and digestate are free of cost, thus only the cost of transport is considered.

To obtain the total annual costs, additional operating costs should be considered, such as additional operating supplies, maintenance costs, laboratory costs, and supervision (Table 4). Capital cost is also converted to an operating cost through straight-line depreciation. The assumed lifetime of the plant is 20 years. It is assumed that all capital equipment has a lifetime of 20 years except for the membranes, which need to be replaced every 5 years (for a total of 3 replacements in years 5, 10 and 15). Total unit production cost is calculated by dividing the total annual cost by the total annual lactic acid production (kg LA/year).

Table 4: Capital and operating cost calculations.

Capital costs	**Estimate**	**Sources**
(A) Total equipment cost	From process model and literature	See Supplementary information (Table S11) for more details.
(B) Procurement and installation cost (% of total equipment cost)	37.5	(Munagala et al., 2021; Phanthumchinda et al., 2018)
(C) Total installed equipment cost	A+B	
(D) Piping, electrical, and instrumentation cost (% of total installed equipment cost)	45	(Daful and Görgens, 2017; Kwan et al., 2018)
(E) Buildings, land improvements, and auxiliary facilities cost (% of total installed equipment cost)	47.5	(Daful and Görgens, 2017; Kwan et al., 2018; Sikder et al., 2012)
(F)Total direct cost	C+D+E	
(G) Engineering, construction, contingencies, and contractor fees (% of total direct cost)	75	(Kwan et al., 2018; Sikder et al., 2012)
(H) Total fixed capital	F+G	
(I) Working capital (% of total fixed capital)	10	(Daful and Görgens, 2017; Kwan et al., 2018; Liu et al., 2015; Munagala et al., 2021)
Total capital investment	H+I	
Operating costs	**Estimate**	**Sources**
(J) Raw materials and utilities costs	From process model and Table 2	
(K) Replacement costs	From capital costs	
(L) Total variable costs	J+K	
(M) Maintenance costs (% of total fixed capital)	5	(Giordano et al., 2017; Kwan et al., 2018)
(N) Operating supplies (% of total fixed capital)	1.5	(Kwan et al., 2018; Tejayadi and Cheryan, 1995)
(O) Operating labour	From calculations	See Supplementary information for more details.
(P) Laboratory costs (% of operating labour cost)	20	(Kwan et al., 2018; Munagala et al., 2021)
(Q) Admin/supervision (% of operating labour hours)	20	(Giordano et al., 2017; Kwan et al., 2018)
(R) Plant overhead (% of operating labour cost)	50	(Kwan et al., 2018; Munagala et al., 2021; Tejayadi and Cheryan, 1995)
(S) Capital depreciation (% of total fixed capital)	5	(Kwan et al., 2018; Munagala et al., 2021)
(T) Insurance, taxes and royalties (% of total fixed capital)	5	(Kwan et al., 2018; Munagala et al., 2021)
(U) Total fixed costs	M+N+O+P+Q+R+S+T	
Total annual costs	L+U	

Table 5: Unit cost of chemicals, utilities and labor.

Chemicals, utilities and labor	**Unit cost**	**Source**
NaOH 100%	0.27 \$/kg	(Peinemann et al., 2019)
H_2SO_4 93%	0.09 \$/kg	(Daful and Görgens, 2017)
Inoculum	11.82 \$/L	See Table S13, Supplementary information for more details.
Electricity (Danish grid)	0.10 \$/kWh	(Statista, 2023)
Electricity (biogas CHP)	Without subsidy: 0.14 \$/kWh With subsidy: 0.03 \$/kWh	See Table S15, Supplementary information for more details.
Steam (industrial average)	30.00 \$/tonne	(Bonatsos et al., 2020)
Steam (on-site woodchips boiler)	56.02 \$/tonne	See Table S14, Supplementary information for more details.
Steam (biogas CHP)	Without subsidy: 39.59 \$/tonne With subsidy: 8.74 \$/tonne	See Table S15, Supplementary information for more details.
Cooling water	0.05 \$/m^3	(Manandhar, 2019)
Tap water	Pilot scale: 2.88 \$/ m^3 Commercial scale: 1.44 \$/ m^3	(DANVA, 2022)
Ultrapure water	Pilot scale: 3.54 \$/m^3 Commercial scale: 2.10 \$/ m^3	(Jonsson and Mässgård, 2021)
Wastewater	Pilot scale: 5.78 \$/m3 Commercial scale: 2.89 \$/ m^3	(DANVA, 2022)
Transport by truck	0.40 \$/tonnekm	(van der Meulen et al., 2020)
Salary plant operator	54,000 \$/year	(Salary Explorer, 2023)
Salary plant supervisor	70,000 \$/year	(ERI Economic Research Institute, Inc, 2023)

2.3.2. Environmental life cycle assessment

The environmental LCA is modeled in SimaPro (PRé Sustainability B.V., 2023) using the ecoinvent database version 3 (Weidema, B.P et al., 2013) for background processes such as chemicals production and waste management (Table S17, Supplementary information). The impacts considered are limited to GWP and calculated through total GHG emissions (in kg CO2-eq) via the IPCC 2013 method considering a 100-year timeline (Muñoz and Schmidt, 2016). As waste streams, the candy waste and liquid digestate are considered as avoided waste flows. Inoculum impact is modelled based on cultivation media, electricity, water, and seed bacteria, where the preparation and storage of the seed bacteria is modeled according to Penicaud et al. (2018).

Capital goods are often excluded from biorefinery LCAs; consequently, there is a lack of available life cycle inventory (LCI) data for biorefinery capital equipment (Vance et al., 2022b). Nonetheless, capital infrastructure can be an important contributor to life cycle impacts for some processes (Frischknecht et al., 2007). To include the impact of the biorefinery capital equipment in the LCA, a proxy is used. The ecoinvent database contains data on the construction of an ethanol fermentation plant with a production output of 900,000 tpy (Steubing et al., 2016). To adjust for the smaller production scales considered in this study (23-23,000 tpy), impacts are considered as a function of total capital equipment weight (Table S10, Supplementary information). The calculated capital equipment weight is taken from the Aspen Plus Economic Analyzer.

2.3.3. Sensitivity analysis

While this study focuses on a Danish context, as discussed in section 2.2.2, candy waste and digestate is readily available in several European countries. Consequently, a sensitivity analysis is conducted on the optimal configuration, applying a range of unit costs and unit environmental impacts representing the variability in European market conditions to understand their impacts on the biorefinery system.

3. Results and discussion

3.1. Economic and environmental performance of base case

The base case (pilot plant) scenario produces 23,030 kg/year of lactic acid of 75% concentration at 95% purity. The total unit production cost is 61.16 \$/kg LA and the GWP is 19.82 kg CO_2-eq/kg LA. It should be noted that pilot scales are not comparable to commercial scale scenarios; thus, the base case is simply used to identify economic and environmental hotspots and understand the effectiveness of the improvement methods. Figure 2 shows the relative contribution of each process step to capital equipment costs, raw material and utility costs, and GWP. The highest capital costs come from the capital equipment needed in the separation steps (60% of total capital equipment cost), while the highest operating costs come from the fermentation step (87% of total raw material and utility costs). Environmental impacts are attributed primarily to the concentration step (50% of total GWP). The input which contributes most to raw material and utility costs is inoculum (85% of total), followed by steam (5% of total), while the input which contributes most to GWP is steam (75% of total), followed by electricity (9% of total) and NaOH (8% of total). More details can be found in the Supplementary information.

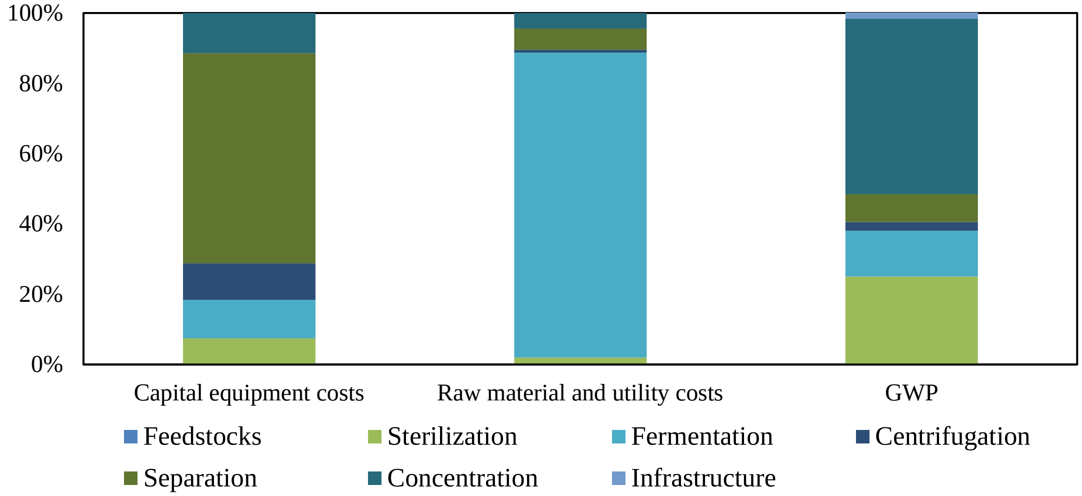


Figure 2: Capital equipment costs, raw material and utility costs, and GWP breakdown by processing step for the base case scenario.

3.2. Economic and environmental performance of improvement methods

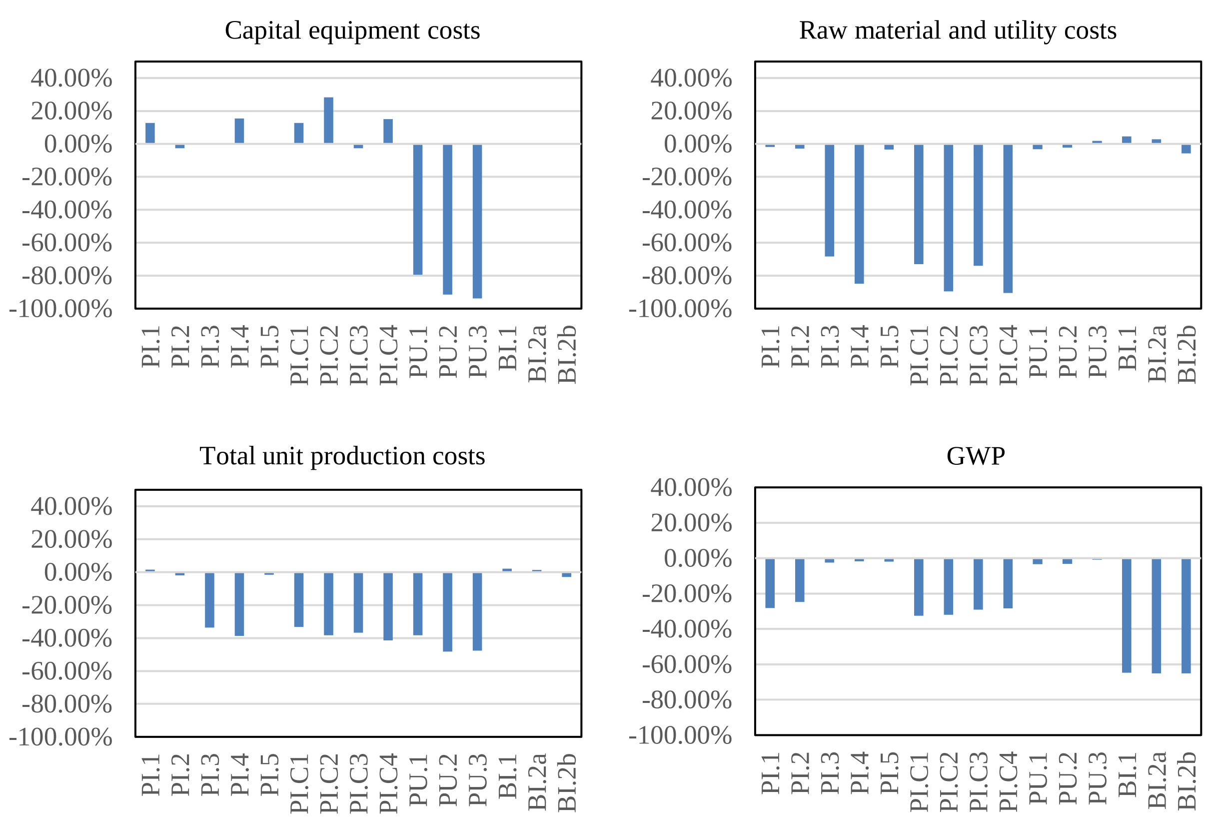


Figure 3: Change in capital equipment costs, raw material and utility costs, total unit production costs and GWP for each optimization scenario compared to base case.

Figure 3 shows the difference in capital costs, raw material and utilities costs, total unit production costs, and GWP for each scenario implementing an individual improvement method compared to the base case. All process improvement scenarios (PI.1-PI.5) are found to reduce the raw materials and utilities costs and GWP. However, only process heat integration (PI.2) is found to also reduce capital equipment costs (-3%), while triple effect evaporation (PI.1) and seed fermentations (PI.4) are found to increase capital equipment costs (+13% and +16%, respectively). Despite seed fermentations having the largest increase in capital cost, the savings achieved in raw materials and utilities costs (-85%) lead to the largest reduction in total unit production costs of all process improvement scenarios (-39%). On the other hand, the savings in raw materials and utilities costs by switching to triple effect evaporation (-2%) are not enough to offset the higher capital costs and lead to an overall increase in total unit production costs (+2%). Finally, both triple effect evaporation and process heat integration result in the largest reduction in GWP compared to the base case (-28% and -25%, respectively) due to the reduction in energy inputs.

Considering scenarios which combine process improvement methods (PI.C1-PI.C4), it can be seen that total unit production costs and GWP improve significantly. However, there is a trade-off between optimizing economic and environmental performance. Implementing triple effect evaporation, inoculum recycling, and water recycling (PI.C1) has the least reduction in unit production cost (-33%) but the greatest reduction in GWP (-33%) compared to the base case, while implementing process heat integration, seed fermentations, and water recycling (PI.C4) has the greatest reduction in unit production cost (-41%) but the least reduction in GWP (-28%). Implementing triple effect evaporation, seed fermentations, and water recycling (PI.C2) or implementing heat integration, inoculum recycling, and water recycling (PI.C3) have moderate unit production cost and GWP reductions.

Per kg LA produced, process upscaling to commercial scales (PU.1-PU.3) demonstrates a significant decrease in capital equipment costs (-79% for PU.1, -91% for PU.2, and -94% for PU.3). It also leads to decreases in labor costs and taxes per kg LA produced, as seen in Table 6. Furthermore, upscaling leads to slight decreases in raw materials and utilities costs per kg LA produced for the 10,000 tpy (PU.1) and 50,000 tpy feedstock (PU.2) scenarios, due to improvements in utility prices for large consumers and in electricity efficiency of equipment (Figure 4). However, for the commercial scale scenario considering 100,000 tpy feedstock (PU.3), an increase in the raw material and utility costs per kg LA produced is observed, due to the increased cost of feedstock transport. The increased transport distances at larger scales also contribute to the slight increase in GWP from PU.1 to PU.2 and PU.2 to PU.3. However, as the main contributor to environmental impact is steam use (see section 4.1), which is not affected by scale, GWP does not vary significantly. Ultimately, unit production cost reaches its lowest point when the feedstock input is 50,000 tpy (PU.2), while GWP reaches its lowest point when the feedstock input is 10,000 tpy (PU.1).

Table 6: Total unit production cost (incl. breakdown) and GWP for base case and improvement scenarios.

Scenario	Unit production cost breakdown ($/kg LA)				Total unit production cost ($/kg LA)	Total GWP (kg CO_2-eq/kg LA)
	Capital[a]	Insurance and taxes	Labor[b]	Raw materials and utilities		
Base case	8.65	3.65	18.86	30.05	61.21	19.82
PI.1	9.72	4.11	18.86	29.50	62.18	14.21
PI.2	8.44	3.55	18.86	29.21	60.06	14.90
PI.3	8.65	3.65	18.86	9.50	40.66	19.32
PI.4	9.95	4.21	18.86	4.55	37.57	19.46
PI.5	8.65	3.65	18.86	29.04	60.20	19.44
PI.C1	9.72	4.11	18.86	8.11	40.80	13.35
PI.C2	8.44	3.55	18.86	7.82	38.67	14.04
PI.C3	11.02	4.67	18.86	3.15	37.70	13.50
PI.C4	9.92	4.19	18.86	2.87	35.84	14.21
PU.1	1.85	0.75	6.05	29.15	37.80	19.14
PU.2	0.80	0.31	1.21	29.42	31.74	19.19
PU.3	0.58	0.22	0.61	30.65	32.06	19.65
BI.1	8.65	3.65	18.86	31.40	62.56	7.00
BI.2a	8.65	3.65	18.86	30.92	62.08	6.91
BI.2b	8.65	3.65	18.86	28.31	59.47	6.91

a-depreciation and maintenance; b- operating, laboratory, supervision, plant overhead;

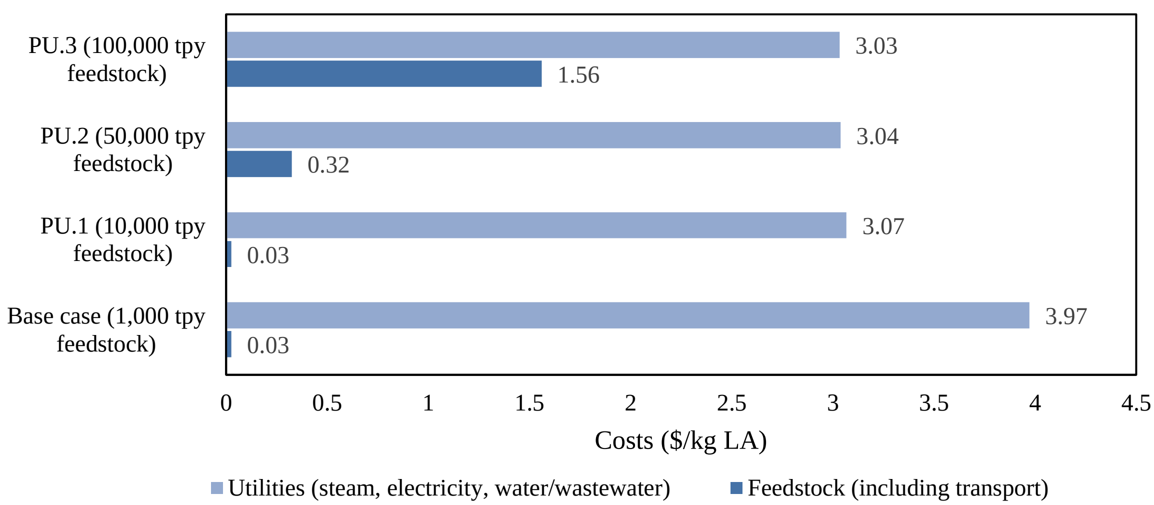


Figure 4: Comparison of feedstock and utilities costs for base case, PU.1, PU.2, and PU.3.

Finally, as can be seen in Figure 3, use of steam generated from an on-site woodchips boiler (BI.1) or electricity and steam generated from a local biogas CHP plant (BI.2a, BI.2b) significantly reduce the environmental impacts of lactic acid production (-65% of GWP compared to base case for both scenarios). In terms of economic impacts, both scenarios result in higher economic costs compared to the base case when excluding subsidies, with use of the woodchips boiler (BI.1) resulting in a higher raw material and utilities cost (+5%) compared to use of a biogas CHP plant (BI.2a, +3%). If subsidies are provided, however, the scenario of electricity and steam provided by biogas CHP (BI.2b) results in a reduction in raw material and utilities cost compared to the base case (-6%).

3.3. Influence of process improvements, scales, and bioenergy integration on selection of optimal configuration

Further investigations are carried out on different combinations of process improvements, scales, and bioenergy integration to assess their collective influence on the sustainability of the system. Table 7 describes the final scenarios considered, and Figure 5 displays the resulting unit production costs and GWP for each scenario in the form of a Parent front. The final scenarios tested have unit production costs ranging from 3.89-15.15 \$/kg LA and GWP ranging from 3.97-5.14 kg CO_2-eq/kg LA. Compared to other LCA studies on lactic acid production, the GWP calculated from these scenarios is reasonable. GWP has been reported in the range of 0.65-7.90 kg CO_2-eq/kg LA when using corn or corn stover (Adom and Dunn, 2017; Ögmundarson et al., 2020), 0.41-4.62 kg CO2-eq/kg LA when using sugarcane bagasse (Daful et al., 2016; Munagala et al., 2021), 5.69-11.28 kg CO_2-eq/kg LA when using macroalgae (Ögmundarson et al., 2020) and 1.33-23.98 kg CO2-eq/kg LA when using dairy side streams (Mediboyina et al., 2024). On the other hand, unit production costs are significantly higher than the lactic acid market price of 1.14-1.57 \$/kg LA (ChemAnalyst, 2023) and to what has been reported in other studies, which have been determined to be in the range of 0.5-4.4 \$/kg LA (Daful and Görgens, 2017; Grasa et al., 2021; Kwan et al., 2018; Liu et al., 2015; Mailaram et al., 2023; Manandhar, 2019; Munagala et al., 2021; Peinemann et al., 2019; Phanthumchinda et al., 2018).

Based on the findings of the study thus far, it could be assumed that the optimal economic configuration should be a biorefinery handling 50,000 tpy feedstock which implements process heat integration, seed fermentations, and water recycling and is supplied steam and electricity by a subsidized biogas CHP plant (F8). The best configuration in terms of environmental performance could be assumed as a biorefinery handling 10,000 tpy feedstock implementing triple effect evaporation, inoculum recycling, and water recycling and utilizing steam and electricity generated from a biogas CHP plant (F1). However, as can be seen in Figure 5, while F8 is indeed identified as the optimal configuration in terms of the economic objective, F4 is found to have the best environmental performance. This shows that optimization is complex, as applying one improvement method can influence the effectiveness of another. As F8 has only a marginally higher environmental impact compared to F4, it is selected as the optimal configuration for the sensitivity analysis.

Table 7: Final optimization scenarios considered.

Scenario names	Process improvements (PI) applied: Triple effect evaporation	Process heat integration	Inoculum recycling	Seed fermentations	Water recycling	Process upscaling (PU) applied (tpy feedstock inputs): 10000	50000	100000	Bioenergy integration (BI) applied: Biogas CHP w/ subsidy
F1	✓	–	✓	–	✓	✓	–	–	✓
F2	–	✓	✓	–	✓	✓	–	–	✓
F3	✓	–	–	✓	✓	✓	–	–	✓
F4	–	✓	–	✓	✓	✓	–	–	✓
F5	✓	–	✓	–	✓	–	✓	–	✓
F6	–	✓	✓	–	✓	–	✓	–	✓
F7	✓	–	–	✓	✓	–	✓	–	✓
F8	–	✓	–	✓	✓	–	✓	–	✓
F9	✓	–	✓	–	✓	–	–	✓	✓
F10	–	✓	✓	–	✓	–	–	✓	✓
F11	✓	–	–	✓	✓	–	–	✓	✓
F12	–	✓	–	✓	✓	–	–	✓	✓

✓ *indicates inclusion of optimization parameters;*

– *indicates no inclusion of optimization parameters*

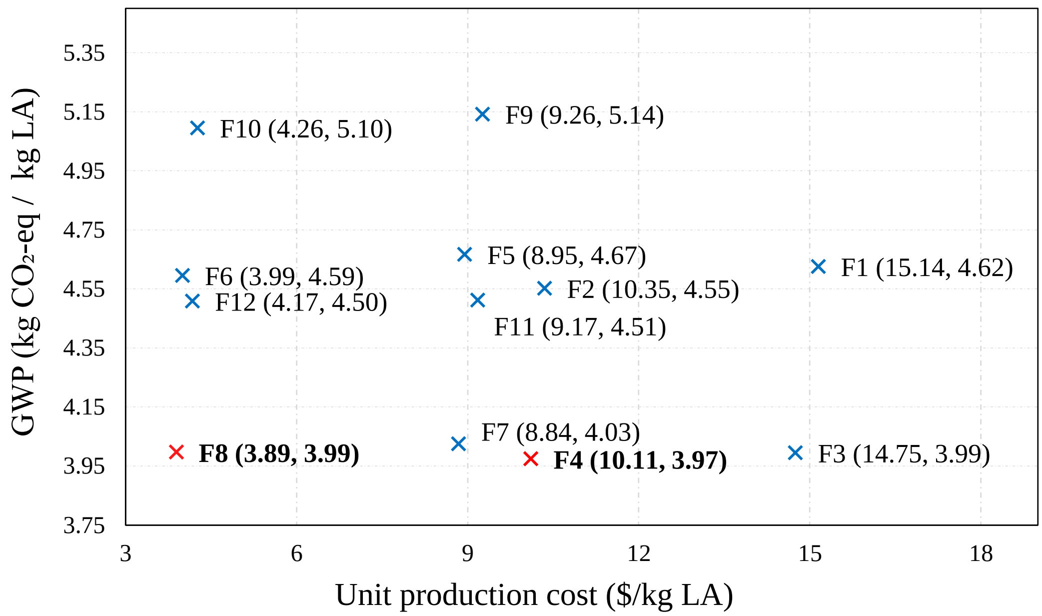


Figure 5: Pareto front for evaluating optimal biorefinery configuration based on unit production cost and GWP for final scenarios considered.

3.4. Sensitivity analysis: Influence of market conditions on optimal configuration

The main contributors to total unit production cost in the optimal configuration (F8) are raw materials and utilities (38%), followed by labor (30%). Of the raw material and utilities costs, the largest contributor is NaOH (24%), followed by steam (23%). The main contributors to GWP are also found to be NaOH (42%) and steam (41%). Consequently, the economic impacts of labor, NaOH, and electricity and steam from biogas CHP (including subsidies) as well as the environmental impacts of NaOH and electricity and steam from biogas CHP are varied according to market conditions seen within a European context (Table 8).

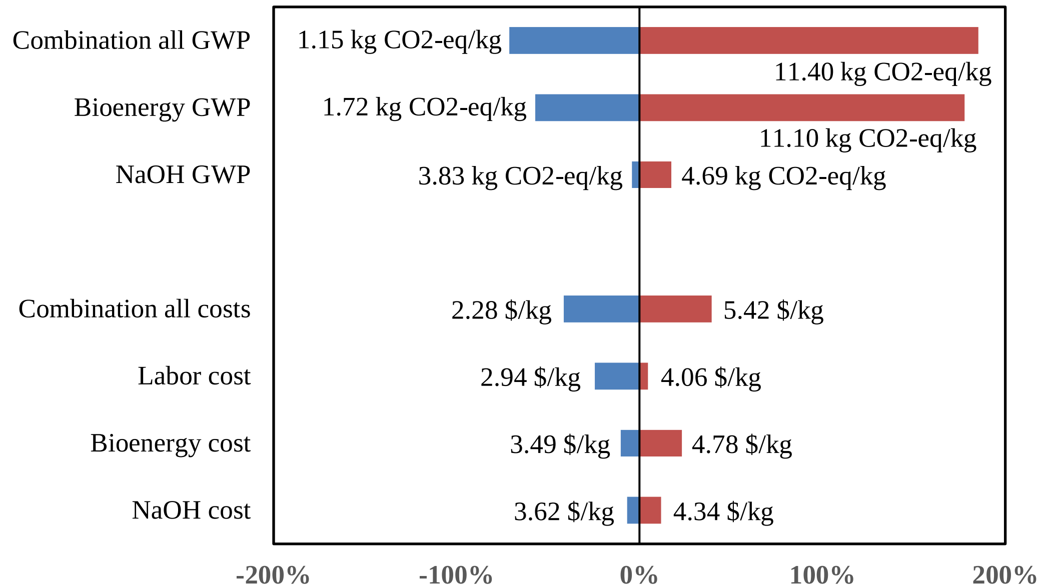
Table 8: Parameters varied in the sensitivity analysis.

Parameter	Current value	Minimum value	Maximum value	Source and details
Labor cost (operator)	53,571 $/year	10,629 $/year	61,600 $/year	(Eurostat, 2024) Minimum: Bulgaria, 2023. Maximum: Luxembourg, 2023.
NaOH cost	0.27 $/kg	0.07 $/kg	0.62 $/kg	(Business Analytiq, 2024) Minimum: Imported from China, March 2021. Maximum: European market, November 2022.
Electricity and steam cost (from biogas CHP with subsidies)	Electricity: 30.11 $/MWh Steam: 8.74 $/tonne	Electricity: 0 $/MWh Steam: 0 $/tonne	Electricity: 149.60 $/MWh Steam: 26.22 $/tonne	(Banja et al., 2019; Kost et al., 2021) Minimum: no heat credit, minimum feedstock price and operating cost, maximum subsidies. Maximum: heat credit, maximum feedstock price and operating cost, minimum subsidies. See Supplementary information (Table S16) for more details.
NaOH GWP	1.29 kg CO_2-eq/kg	0.86 kg CO_2-eq/kg	1.51 kg CO_2-eq/kg	ecoinvent Minimum: European production, membrane cell production process. Maximum: Non-European production, diaphragm cell production process.
Electricity and steam GWP (from biogas CHP)	Electricity: 222.00 kg CO_2-eq/MWh Steam: 40.70 kg CO_2-eq/tonne	Electricity: -395.00 kg CO_2-eq/MWh Steam: 0 kg CO_2-eq/tonne	Electricity: 1,288.80 kg CO_2-eq/MWh Steam: 160.65 kg CO_2-eq/tonne	(Fusi et al., 2016; Shimako et al., 2016) Minimum: cow slurry as feedstock. Maximum: cultivated microalgae as feedstock.

The sensitivity analysis indicates high variability in market conditions which have direct impacts on establishing a European biorefinery. The environmental benefits of utilizing electricity and heat generated from biogas CHP plants are highly dependent on the feedstock, whereas the economic benefits are dependent on the feedstock price, selling price of the generated electricity and heat, and subsidies available, all of which vary significantly by country. Chemical costs are also regionally dependent, with imported NaOH from China costing 3-4 times less than NaOH from Europe (Business Analytiq, 2024). On the other hand, NaOH produced in Europe is found to have a lower GWP than the rest of the world, which could be due to high energy demand of chemical production and the increased penetration of renewable energies in Europe (Perez Sanchez et al., 2023; Ritchie and Rosado, 2024). Finally, labor costs have significant variability within Europe, with the average laborer in the highest-income country making 6 times more than the average laborer in the lowest-income country.

Figure 6 displays a tornado plot of the impacts of the sensitivity analysis on the optimal configuration. Considering the lowest labor costs, NaOH costs, and bioenergy costs, the lactic acid unit production cost lowers to 2.28 $/kg. On the other hand, if labor costs, NaOH costs, and bioenergy costs are at the higher range, unit production cost increases to 5.42 $/kg. An even higher variability can be seen in GWP, where considering a lower GWP for NaOH and bioenergy production would result in a lactic acid GWP of 1.15 kg CO_2-eq/kg, whereas a higher NaOH and bioenergy GWP would increase the GWP of lactic acid to 11.40 kg CO_2-eq/kg.

Figure 6: Impacts of sensitivity analysis on environmental and economic impacts of optimal configuration (F8).



4. Discussion

In this study, a novel lactic acid production process is modelled, using a feedstock mixture of candy waste and liquid digestate, applying LAB fermentation and membrane processes for downstream separation and purification. The study first uses lab-scale experimental results to determine the yield of unit processes. The benefit of utilizing experimental data is that it can capture losses which would otherwise be idealized by the process model. For example, some studies have assumed that all or a significant portion of the base (NaOH in this study) is recoverable from the BPED membrane and can be reused in the fermentation process (Åkerberg and Zacchi, 2000). In this study, it was found that many of the ions were separated by NF, and the base concentrate recovered from BPED was primarily water (Table S1, Supplementary information). Thus, even with further concentration, direct recycling of this stream would contribute little to nothing to the base demands of the process. Future work could focus on running the NF retentate through BPED to determine whether both acids and bases could be effectively recovered. A drawback of using experimental data is that processes are often configured to improve the yield or energy consumption of that specific unit process, without considering the consequences to other unit processes. For example, in this study, water was added to the MF process to increase the flux and thus reduce the operating time and energy used in the process. However, this decision would increase energy required in the concentration process, when that same water would need to be removed. In future studies it is suggested that experimental optimization also consider impacts from a systems level. For this case, the impact of dilution water could be studied more closely.

The biggest drawback of utilizing experimental results is the limited amount of data available. First, the experiments were run as a series of batch processes; however, as discussed in section 2.1.2, the system was modelled considering continuous processes, as they are more typical in commercial systems. This increases the uncertainty of the results, particularly when modelling continuous fermentation, which tends to achieve higher yields than batch fermentation but also has an increased risk of contamination (Ingledew and Lin, 2011). Improving the accuracy of the model would require experimental validation of the continuous system; however, as the results showed that the production pathway considered in this study was not economically viable, it is unlikely such a system would be built. On the other hand, this study only considered downstream lactic acid separation and purification via membrane processes. This makes comparison with other studies difficult, particularly with environmental LCAs which up until this point have only considered the conventional fermentation and downstream processing method of lime neutralization and acidification followed by reactive distillation. Future experimental work should therefore explore the production of lactic acid using the feedstocks considered in this study but applying conventional processes to see if better performance could be achieved. Furthermore, future experimental work could explore the many additional fermentation and downstream processing pathways which have been suggested in the literature, such as non-sterile fermentation (Peinemann et al., 2019), the use of acid tolerant bacteria (Daful and Görgens, 2017), membrane bioreactors (Sikder et al., 2012), ion exchange (González et al., 2007), and reverse osmosis (Phanthumchinda et al., 2018).

Membrane technologies are still developing (Li et al., 2021); thus, there is limited data available related to their commercial operation, such as economic costs and environmental impacts of capital equipment. In particular, only a few articles were found which included data on the capital costs of microfiltration and nanofiltration (Giordano et al., 2017; González et al., 2007; Sikder et al., 2012; Tejayadi and Cheryan, 1995) and even fewer articles included data on the capital costs of electrodialysis (Åkerberg and Zacchi, 2000; Vineyard et al., 2020). Only one article was found which assessed the environmental impact of microfiltration membrane manufacturing (Tangsubkul et al., 2006), and no data could be found on the environmental impacts of nanofiltration and electrodialysis capital equipment. This is therefore another source of uncertainty within the results obtained from this study, as the capital costs assumed for the membrane unit processes were based on limited data points, and the environmental impacts of the capital equipment could not be considered on a unit basis but had to be approximated on an aggregate basis. Future work could attempt to more clearly define the economic and environmental impacts of capital equipment for membrane technologies.

In this study, the base case is assumed as a pilot scale operation utilizing 1,000 tpy fermentation feedstock, which can produce 23 tpy LA. While average production capacities in studies assessing the economic or environmental impacts of lactic acid production are typically higher, from 100 tpy (Phanthumchinda et al., 2018) to over 30,000 tpy LA (Adom and Dunn, 2017), it is necessary for a novel production process to first be proven at pilot scale. It is therefore interesting to find that process improvements implemented at pilot scale performs differently than process improvements implemented at commercial scales (Section 3.3). While process improvements can successfully reduce operating costs and environmental impacts by reducing energy, chemical, and water consumption, some improvements require significant increases in capital equipment investment, which at a smaller capacity might not be worth the additional cost. The main benefit of upscaling is that the share of costs which is attributed to capital equipment decreases, meaning more capital-intensive processes such as multi-effect evaporation and seed fermentations can be included. However, there are limitations to upscaling, as longer feedstock transport distances are needed which can result in higher environmental impacts and offset the cost reductions achieved.

Bioenergy integration is found to be successful in reducing environmental impacts, but only results in lower economic costs if subsidies are available. Furthermore, obtaining both feedstocks and energy at favorable prices requires positive relationships between stakeholders to promote a win-win scenario (Vance et al., 2022a). This could be analyzed in future work.

Even after process improvements, upscaling, and bioenergy integration including subsidies, the unit production costs found in this study were higher than other studies. While some of this could be attributed to geographical differences and others could be attributed to inflation, it was found that several unit costs were estimated at significantly lower values in previous studies, including capital equipment, chemicals, water, and steam. Inoculum costs in particular were found to be significantly lower in other studies (Daful and Görgens, 2017; Mailaram et al., 2023; Peinemann et al., 2019), and no other study with the exception of Kwan et al. (2018) considered the cost of waste treatment. In their study Furtado Amaral et al. (2020) provide an insightful discussion on how manufacturing costs in bioenergy systems have been underestimated in many studies, and the consequences this can have on economic support policies. The authors of this study also feel that without quantifying the full scope of impacts, decision-makers can have an unrealistic view of the cost of producing lactic acid in the European context. As seen in the sensitivity analysis, unit production costs could be decreased by shifting production to a country with low labor costs and high biogas subsidies and importing the chemicals needed from markets with lower prices. However, it is important to consider that production processes in countries with lower prices may have increased environmental impacts due to the energy sources used (Ritchie and Rosado, 2024). Furthermore, such countries may perform worse in terms of social impacts such as offering a living wage and workers' rights (UNEP/SETAC Life Cycle Initiative, 2011). While such aspects were not considered in this study, in future studies it could be worth exploring the value of establishing a European biorefinery in terms of social objectives. Ultimately, despite the poor economic performance found in this study, the model and results are valuable for future studies, which could simulate further scenarios from the baseline reported in this work.

5. Conclusions

This study assesses the economic and environmental sustainability of a novel lactic acid production process in Denmark using a feedstock mixture of candy waste and liquid digestate, applying LAB fermentation and membrane processes for downstream separation and purification. Along with the base case, several scenarios are modelled varying unit processes, scales, and energy sources to understand the independent and collective effects of improvement methods on process optimization. The study shows that while improvements in economic and environmental performance can be achieved through process improvements, upscaling, and bioenergy integration, there are still challenges to overcome regarding the use of novel feedstocks and downstream processes in industrial lactic acid production, such as limited supply chains and uncertain market conditions. The hope is that this research can help inform future studies on European lactic acid biorefinery systems, particularly with regards to process optimization strategies.

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