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Integrated biorefinery for bioethanol and succinic acid co-production from bread waste: Techno-economic feasibility and life cycle assessment

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ABSTRACT

In this study, an advanced decarbonization approach is presented for an integrated biorefinery that co-produces bioethanol and succinic acid (SA) from bread waste (BW). The economic viability and the environmental performance of the proposed BW processing biorefinery is evaluated. Four distinctive scenarios were designed and analysed, focusing on a plant capacity that processes 100 metric tons (MT) of BW daily. These scenarios encompass: (1) the fermentation of BW into bioethanol, paired with heat and electricity co-generation from stillage, (2) an energy-optimized integration of Scenario 1 using pinch technology, (3) the co-production of bioethanol and SA by exclusively utilizing fermentative CO_2, and (4) an advanced version of Scenario 3 that incorporates carbon capture (CC) from flue gas, amplifying SA production. Scenarios 3 and 4 were found to be economically more attractive with better environmental performance due to the co-production of SA. Particularly, Scenario 4 emerged as superior, showcasing a payback period of 2.2 years, a robust internal rate of return (33% after tax), a return on investment of 32%, and a remarkable net present value of 163 M$. Sensitivity analysis underscored the decisive influence of fixed capital investment and product pricing on economic outcomes. In terms of environmental impact, Scenario 4 outperformed other scenarios across all impact categories, where global warming potential, abiotic depletion (fossil fuels), and human toxicity potential were the most influential impact categories (−0.344 kg CO_2-eq, −16.2 MJ, and −0.3 kg 1,4-dichlorobenzene (DB)-eq, respectively). Evidently, the integration of CC unit to flue gas in Scenario 4 substantially enhances both economic returns and environmental sustainability of the biorefinery.

1. Introduction

European Union's Renewable Energy Directive sets a goal to replace 10% of transport fuel with biofuels, such as bioethanol. Likewise, within the United Kingdom (UK), the Renewable Transport Fuel Obligation (RTFO) enforces fuel suppliers to secure a proportion of their fuel supply from renewable sources. Furthermore, RTFO has set an increased biofuel production target, from 4.75% in 2020 to 12.4% by 2032 [1]. They have also added targets for advanced waste-based renewable fuels, increasing from 0.1% in 2019 to 2.8% by 2032 [2]. These endeavours are in accordance with the aim of the UK government, which seeks to reduce its overall greenhouse gas emissions by 80% by the year 2050 [3]. In the UK, bioethanol production primarily relies on first-generation edible feedstocks, specifically wheat and sugar beet. Several prominent bioethanol producers in the UK include Ensus (400 million liters from wheat), Vivergo (420 million liters from wheat), and British Sugar (70 million liters from sugar beet) [4]. However, the escalating production of bioethanol, in conjunction with population growth, has raised significant concerns in terms of the sustainability of first-generation bioethanol [5].

Numerous studies have thus investigated other alternative sources of bioethanol production from varieties of biomass. Bioethanol can be produced from a range of renewable feedstocks rich in carbohydrates, which can be readily hydrolysed to fermentable sugars to yield ethanol. A promising option is the utilization of waste biomass. This approach offers significant advantages, including efficient and cost-effective sugar

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extraction, without impacting food production [6]. There is a growing interest in utilizing food waste for the production of environmentally friendly bioethanol. Notably, bread, as one of the most consumed food items, also falls within the category of foods that are highly wasted [7,8]. According to the Waste and Resources Action Programme (WRAP) UK, bread waste (BW) constitutes 10% of the total food waste produced in the UK [9]. In fact, bread is identified as the second most wasted food in the UK with 20 million slices of bread being disposed daily, contributing to a yearly wastage reaching 292,000 tons with an estimated CO_2-eq emissions of 584,000 tons [10], which has a substantial influence on both the environment and the economy. BW characterized by its starchy composition has emerged as a highly promising and environmentally benign sustainable carbohydrate source for producing high-value biochemicals and bioenergy through fermentative routes [11–13]. Currently, bakeries serve as convenient locations for collecting BW from retailers. This practice is highly likely to ensure the proper segregation of BW from other types of food waste. The collection of BW through this method is made possible under a take-back agreement clause, a common practice in some European countries [14,15]. Unlike lignocellulosic biomass, which requires harsh physical, chemical, or enzymatic pretreatment to release the sugars, BW is notable for being readily accessible source of fermentable sugars [10]. Thus, it is perceived as a sustainable feedstock for bioenergy and biochemicals, with lower carbon emissions and a greener, cleaner profile [16,17].

Recent improvements and innovations have made bioethanol production derived from biomass more sustainable, from both environmental and economic perspectives. For instance, the nutrient rich syrup contained in the stillage of ethanol is converted into distillers dried grains and soluble (DDGS) for animal feed application [18,19] and energy recovery of waste materials resulted through an integrated anaerobic digestion (AD) and combined heat power (CHP) [20–23]. In addition, other remaining useful fractions of biomass such as lignin and C5-sugars are also utilized to produce added value products [24]. Giwa et al. [25] conducted a study to assess the economic feasibility of a multi-product biorefinery that generates bioethanol, hydrogen, and bio-oil by fast pyrolysis of by-products. Their findings indicated that the most profitable scenario involved producing ethanol from biochar and hydrogen from hydrogen-rich non-condensable gas. The economic feasibility of this approach was largely dependent on variables such as the prices of feedstock and products, as well as the initial capital investment. Pinto et al. [26] explored the potential of 2G ethanol production in biorefineries to contribute to a low-carbon bioeconomy. The findings showed that the most profitable scenario involved biomass washing but came with higher environmental impacts (2–18%) compared to conventional 1G to 2G approaches. However, the addition of protein had the lowest environmental impact, and a combination of mitigation strategies became economically viable due to increased commercialization of decarbonization strategies prices and affordable soybean protein costs. Demichelis et al. [27] presented a three-step method for assessing the technical, environmental, and economic dimensions of bioethanol derived from various waste biomasses (sugarcane, potatoes, rice straw, cattle manure, and organic municipal solid waste). Their study determined the bioethanol yields, economic feasibility, and environmental effects, revealing that cattle manure had the most substantial reduction in environmental impacts, while sugarcane exhibited the lowest energy consumption. Vaskan et al. [28] examined the potential economic viability and environmental implications of utilizing empty fruit bunches to create multiple products, including ethanol, heat, power, and C5 syrup. The scenario showed a better environmental performance than relevant reference systems. The study of Ali Mandegari et al. [21] explored various biorefinery scenarios for producing bioethanol and electricity using lignocellulose from a sugar mill, with different combinations of bagasse, brown leaves, and coal. Among the scenarios, using a centralized CHP unit without coal co-combustion was found to be the most environmentally favourable, while scenario where bagasse and brown leaves were utilized for coal burning, stood out as the economically viable choice due to lower capital costs and economies of scale. Sadhukhan et al. [29] performed a study to examine both the economic and environmental dimensions of bioethanol production using diverse lignocellulosic biomasses found in Mexico. The results showed that rubber wood, woody biomass, and wheat straw were the most economically viable choices. Additionally, in terms of environmental impact, coffee pulp, husks, and woody residues were identified as the most eco-friendly feedstock options [29]. However, it is important to note that CO_2 that emerges from the fermentation and CHP unit is ultimately discharged into the atmosphere, raising environmental burdens. Decarbonization has become a critical issue for reducing or eliminating emissions associated with bioethanol production technologies.

During bioethanol production, one-third of carbon is lost as CO_2 [$C_6H_{12}O_6 \rightarrow C_2H_5OH + 2CO_2$] which restricts ethanol yields to 0.51 g/g [30]. The recovery and utilization of this carbon can significantly improve the overall efficiency of the process [30]. Capturing the CO_2 generated during bioconversion of sugars from fermentation and the flue gas in the CHP unit has attractive potential to achieve net-zero CO_2 emissions [31]. One potential co-product that can be derived from CO_2 capture and use is succinic acid (SA), recognized as a valuable product with an anticipated market value of 282.8 million USD [32]. The fermentative SA production via reductive TCA (tricarboxylic acid) pathway requires CO_2 as co-substrate [$C_6H_{12}O_6 + 2CO_2 + 2NADH + 2H^+ \rightarrow 2C_4H_6O_4 + 2H_2O$] [33]. Thus, the CO_2 produced during fermentation and flue gas combined with glucose produced from bio-feedstock sources can be utilized for SA production. SA can function as a fundamental component for generating a diverse array of applications within the pharmaceutical, food, cosmetics, and chemical sectors [34]. By adding CC unit to extract CO_2 from the flue gas, additional SA can be generated, augmenting the overall value of the production process compared to solely focusing on bioethanol. Hence, combining food waste conversion of bioethanol production with sustainable conversion of CO_2 to value-added chemical marks a shift from a linear 'cradle to grave' chemicals manufacturing mode to a circular economy [35]. It is imperative to formulate a low-carbon manufacturing strategy, especially concerning food waste valorisation, to establish a sustainable approach to bioethanol production. To the best of our knowledge, existing literature has yet to report on the economic and environmental evaluation of decarbonization strategies for bioethanol derived from BW. The novelty of this study lies in its exploration of a techno-economic and life cycle assessment, aimed at decarbonizing the production of bioethanol from BW. Furthermore, this study explores the use of CO_2 derived from ethanol fermentation and flue gas emissions for bio-SA production. Our methodology is thoroughly comprehensive, encompassing advanced process modeling using pinch technology, coupled with a robust evaluation of economic feasibility and life cycle assessment. These approaches allow us to assess the efficiency of various operational strategies, facilitating an in-depth understanding of yield efficiencies, heat, and power consumption across different scenarios. Such insights are pivotal in determining the cost-effectiveness and analyzing the environmental implications of the proposed strategies. A distinctive feature of our research is the exploration of a carbon-negative approach for a BW-centric multi-product bioethanol biorefinery, championing it as both sustainable and environmentally conscientious.

2. Methodology

This research aims to assess the economic and environmental aspects of producing bioethanol from BW. The entire process is modelled using Aspen Plus V11 to estimate material and energy flows in the bioethanol production process. Fig. 1 provides a simplified overview of the BW valorisation strategy, encompassing a spectrum of sub-processes such as enzymatic hydrolysis, bioethanol fermentation, downstream separation processing, AD, CHP, and CC for SA production. The comprehensive analysis involves techno-economic assessment (TEA) and life cycle

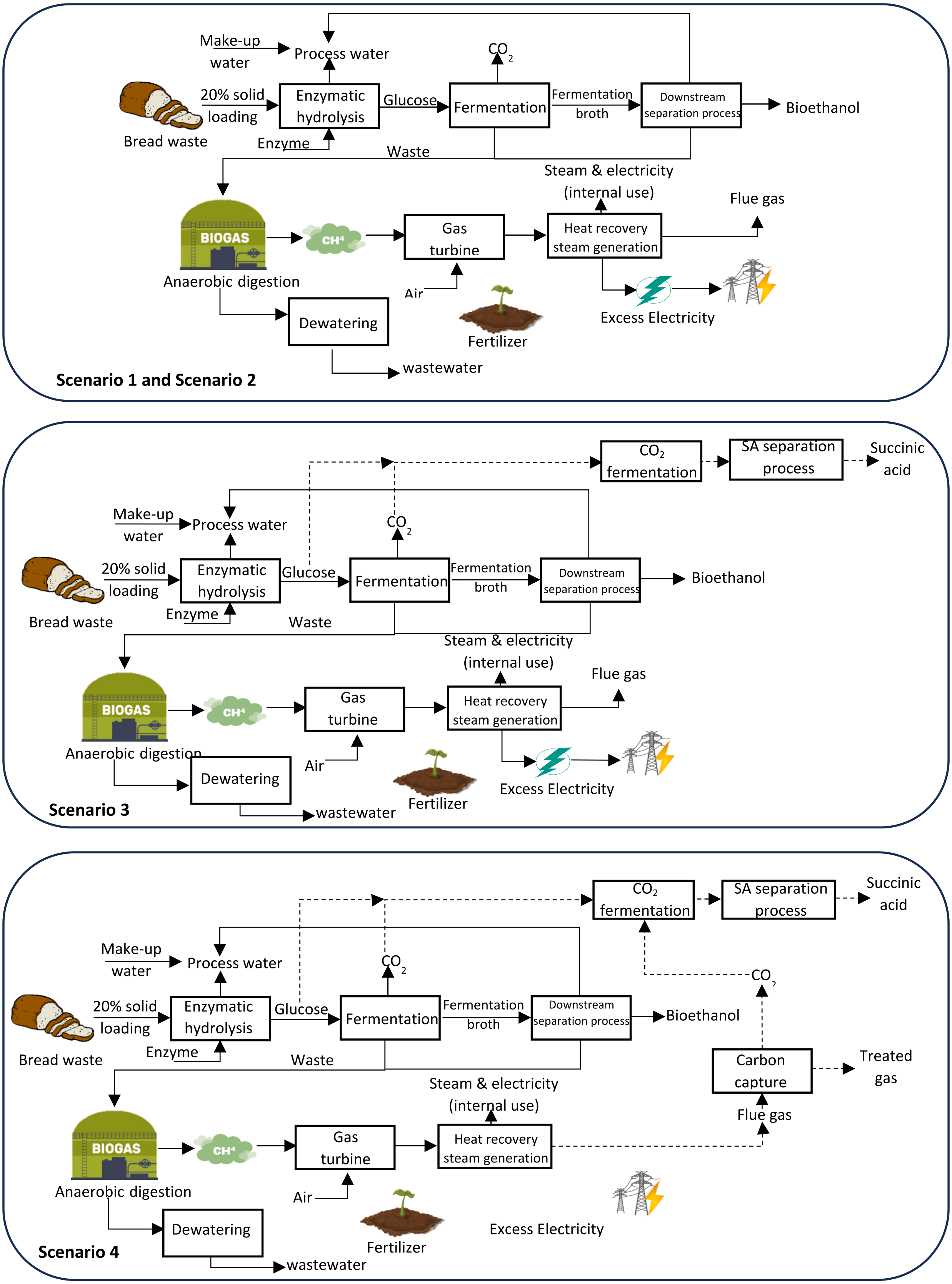

Fig. 1. Simplified schematic representation of an integrated bread waste valorisation into multiple products including ethanol, biogas and SA for all scenarios studied.

assessment (LCA), which together provide insights into the sustainability of this process. The TEA entails conceptual process design for mass and energy balance, cost estimations for initial investment and ongoing operations, profitability analysis, and sensitivity assessments with respect to key economic factors. Concurrently, the LCA, executed using SimaPro software v9.4.0.2 with the Ecoinvent 3.8 database, quantifies the environmental ramifications of the entire process, offering a holistic perspective of its sustainability. In this study, the geographical location of the proposed biorefinery is assumed to be in the UK region. Consequently, the assumptions for the TEA and LCA are specifically customized to align with this region's specific context.

The following four biorefinery scenarios are designed for integration into a system aimed at extracting value from the BW, with their primary distinctions centred on co-product generation and efforts to reduce carbon emissions.

- **Scenario 1**: The process begins with BW that undergoes sequential hydrolysis and fermentation. During this phase, fermentable sugars are consumed by *S. cerevisiae KL17 microorganism* to produce bioethanol and CO_2. Bioethanol is subsequently recovered from the fermentation mixture in the downstream section. The remaining waste fraction, primarily protein and fat, undergoes digestion in AD, leading to biogas production. The biogas is utilized as an energy source for a CHP unit, effectively supplying the necessary energy for the entire process, encompassing both heat and electricity demands. Dewatered digestate from AD is directly used as a fertilizer or soil conditioner, contributing to sustainable waste management.
- **Scenario 2**: It shares similarities with Scenario 1 in terms of the products and process involved. However, in this scenario, pinch technology is employed to provide a systematic and structured approach to improving energy efficiency, reduce operating costs, and optimize process design. The biogas produced from the AD is utilized for CHP generation.
- **Scenario 3**: This scenario considers CO_2 valorisation to produce SA. The CO_2 emitted from bioethanol fermentation is directed into the SA fermentation process. A portion of the fermentable sugar needed for this process is sent to the fermentation reactor, where glucose and CO_2 are converted into SA using *Actinobacillus succinogenes*. The amount of glucose diverted towards SA fermentation is determined based on the quantity of CO_2 generated, which acts as a limiting factor. Thus, to convert all CO_2 produced during bioethanol fermentation to SA product, the ratio of glucose diverted to CO_2 fermentation is 0.53 glucose needed for SA/total glucose. After fermentation, the resulting broth is filtered, and the filtrate undergoes downstream processing to achieve pure SA. Notably, all the residues generated from these two processes are efficiently employed to generate steam and electricity, contributing to resource optimization. Similar to the previous two scenarios, considerations for residue utilization and a heat integration network are applied in Scenario 3, optimizing energy efficiency.
- **Scenario 4**: It considers the utilization of CO_2 emitted from both bioethanol fermentation and flue gas of CHP to produce more SA. The additional CO_2 from CC unit is sent to CO_2 fermentation unit to generate more SA product, however, the capacity of the biorefinery is kept constant at 100 MT BW/day. The same considerations of pinch analysis, AD, and CHP are applied in this scenario.

2.1. Process description

2.1.1. Enzymatic hydrolysis

In this work, the selected feedstock is wet BW with a composition of 46 wt% carbohydrates, 7.9 wt% protein, 2 wt% fat, 2.5 wt% fibre, 1% salt, and 40.6% water. This composition and key process details are based on the research conducted by Narisetty et al. [36]. To prepare the BW for further processing, the BW suspension is subjected to autoclaving, which eliminates the need for gelatinization and liquefaction steps. The autoclaving process involves loading the BW into the autoclave with a suitable quantity of water to create a slurry. The autoclave is sealed, and it is heated to a high temperature, typically ranging from 121 °C to 150 °C, with a pressure of 1–3 bar. These elevated temperature and pressure conditions break down the complex carbohydrates and other organic compounds in the BW into simpler molecules, which re easier to digest in the subsequent enzymatic hydrolysis process. Additionally, autoclaving serves to sterilize the waste material, eliminating any potentially harmful microorganisms. The produced supernatant is then cooled to 60 °C and transferred to enzymatic hydrolysis with the addition of *Dextrozyme Peak enzyme* at a rate of 0.6 mg per gram of BW for a period of 48 h. During enzymatic hydrolysis, the carbohydrate is transformed into individual glucose molecules, facilitated by the combined action of the enzyme amylase and amyloglucosidase. Once this saccharification process is complete, the resulting sugar-rich liquid, is carefully separated from other components that have undergone enzymatic hydrolysis.

2.1.2. Fermentation and downstream separation

Prior to fermentation, the saccharified slurry is cooled to 32 °C. The process conditions, 1 bar, 32 °C, are based on experimental data from the work of Narisetty et al. [36]. In this phase, the hydrolysed slurry is mixed with the fermenting yeast *Saccharomyces cerevisiae*. After fermentation, the resulting culture broth, containing approximately 7–10% ethanol by weight, undergoes separation and purification through two distillation columns. Before entering the first distillation column, the culture broth is heated to 100 °C to minimize heating duty in the reboiler of the first distillation column. In the first column, which consists of 16 stages, ethanol is concentrated to approximately 21% by weight as distillate, while the water-soluble organics are separated in the bottom stream. Subsequently, in the second distillation column, the liquor undergoes further concentration until it reaches the azeotropic concentration of around 93 wt% ethanol. This azeotropic mixture is then directed to the dehydration zone, where a molecular sieve is employed to attain an ethanol concentration of 99.7% by weight [37].

2.1.3. AD and CHP

The water-soluble organic streams from fermentation and downstream units are sent to AD for biogas production. Here the wastes are digested at 35 °C for a period of 10 days. The biogas generated is then directed to the CHP unit, where heat and electricity are produced. The resulting digestate from the AD is considered a useful by-product and can be utilized as a substitute for fertilizers [38]. The gaseous emission data for one kg of digestate used in the land application is sourced from Tiwari et al. [39]. Meanwhile, the generated biogas is directed to the CHP, characterized by an electrical efficiency of 40% and a thermal efficiency of 45% [40]. The subsystem comprising a combustor, boiler, and turbo generator is responsible for burning the biogas derived from the AD process. The mid-pressure steam (MPS) is generated in the boiler to fulfil process heating requirements. Additionally, electricity generated is used to satisfy the plant's electricity demand. Excess electrical energy generated in the CHP unit is sold to the public grid at the prevailing UK market rate, generating additional revenue.

2.1.4. Pinch analysis

Energy conversion is a critical aspect of process design, particularly in facilities with high energy consumption. Heat integration plays a crucial role in enhancing both the economic viability and environmental sustainability of a process by reducing energy consumption and utility costs [24]. This integration is achieved through the incorporation of heat exchangers to utilize heat from various hot streams for cold streams within the system. These streams are strategically arranged to maximize heat exchange. Utilizing pinch technology, an initial estimation of the minimum energy needs (a target) is determined by plotting the composite hot and cold curves, assuming a 10 °C approach temperature (ΔT

min). In this study, the energy requirements of all proposed process configurations were assessed using Aspen Energy Analyzer® software. Inlet and outlet temperatures are used to calculate the heat energy necessary for each piece of equipment. When considering heating requirements, heat is primarily allocated to streams with higher outlet temperatures. Conversely, for cooling needs, heat is allocated to streams with lower outlet temperatures.

2.1.5. CO_2 fermentation to succinic acid

The off-gas generated during ethanol fermentation contains predominantly biogenic CO_2 [41]. Notably, in this study, there is no need for a CO_2 purification system in the subsequent upgrading process. Instead, gas fermentation technology is employed to transform the gaseous carbon waste streams into SA. In this process, all CO_2 is reacted with glucose, which is then fermented using *A. succinogenes* to produce SA at 37 °C for 44 h of fermentation time. The downstream procedure is adapted from a previous study by Thanahiranya, et al. [42], with the main challenge being water removal during SA purification. SA fermentation from glucose yields approximately 0.55 g of SA per gram of BW [43]. The fermentation broth is directed to a flash drum to eliminate gas, and the resulting bottom stream is preheated prior to entering the evaporator. Within the evaporation process, the dissolved product steam is concentrated at an evaporation temperature of 101.5 °C. Following this step, a centrifuge is employed to separate SA into a solid form, facilitating further SA purification. Eventually, a product with a purity of 99.5% is achieved through the process of drying with hot air [42,44].

2.1.6. CC

The CC unit involves an absorber/stripper process, where CO_2 is absorbed within the absorption tower and then released or stripped from the absorbent in another tower known as the stripping tower. This process simulation is based on the flue gas emitted from the CHP unit, with a flow rate of 156.2 kmol/h and gas composition of 15 mol% H_2O, 13 mol% CO_2, 1 mol% CO, 3 mol% O_2, and 68 mol% N_2. The CC unit employs chemical absorption, utilizing an aqueous solution of methyldiethano-lamine (MDEA, 45% w/w) and piperazine (PZ, 5% w/w), following similar principles as established in previous studies by Malekli et al. [45] and Mudhasakul et al. [46]. Post-combustion, the flue gas is compressed to a high pressure of 43.2 bar and cooled to 25 °C using a direct cooler before entering the bottom absorber. Simultaneously, the amine solution is introduced from the top tower at 52 °C and 42.1 bar, allowing it to come into contact with the lean amine inside the absorber. The treated gas, primarily composed of 94 mol% N_2 and other gases, is released into the atmosphere from the column's top. Meanwhile, the CO2-rich amine stream exits the bottom of the tower via a pressure relief valve, reaching a pressure of 3.2 bar, with a temperature at this pressure of 72.82 °C. The rich amine with small quantities of dissolved off-gases, primarily consisting of N_2 and O_2, is directed to a flash drum to remove these gases. Subsequently, the rich amine solvent stream from the bottom of the flash drum enters a stripping column, releasing CO_2 at the top of the column. To minimize the thermal energy required for solvent regeneration within the CO_2 stripper's reboiler, the CO_2-rich MDEA/PZ solution is pre-heated before entering the stripper. This pre-heating is achieved by recovering heat from the hot CO_2-lean amine solution exiting the reboiler, with both streams flowing counter-currently. Thus, the feed stream is fed to the stripping column at 98.9 °C. The resulting CO_2 product, boasting a purity of 99 mol%, is obtained from the top of the tower. Conversely, the lean amine solvent at the distillation column's bottom is cooled by exchanging heat with the bottom stream of the flash drum. Subsequently, the regenerated amine solvent is pressurized by a pump before being recycled back to the top of the absorber column.

2.2. Techno-economic evaluation

2.2.1. Total capital investment

The equipment procurement expenses for the developed process are calculated using the Aspen Plus Economic Evaluator. However, for estimating the procurement cost of particular equipment (e.g., fermenters), data from existing literature is employed from Tiwari et al. [39]. In the case of equipment costs obtained from literature, adjustments are made in accordance with standard engineering scaling factors using the following equation:

$$new\ \ cost = base\ \ cost \times \left(\frac{new\ \ size}{base\ \ size}\right)^{0.6} \tag{1}$$

The overall capital expenditure (Capex) is determined by adding up the fixed capital investment (FCI), land utilization, and working capital. The FCI itself includes both the total direct costs (TDC) and total indirect costs (TIC). TDC covers a spectrum of expenses such as the procurement and installation of equipment, instrumentation, control systems, as well as expenses related to piping and electrical components. Additional details concerning unit costing, equipment procurement, and the capex calculation can be found in the Supplementary materials. The parameters used for discounted cash flow analysis in this study are outlined in Table 1. The discounted cash flow analysis parameter and total capital investment factor are taken from Humbird et al. [47] and Peters et al. [48].

2.2.2. Total operating cost

The operational expenses consist of specific categories, including direct manufacturing expenses, fixed manufacturing costs, and general manufacturing costs. These cost components are calculated using the equations provided by Turton, et al. [49]. Detailed information on the costs associated with raw materials and utilities for all the processes employed in this study is presented in Table 2.

2.2.3. Profitability analysis

Total revenue of the biorefinery for all scenarios is calculated based on the main product as well as all the co-products generated from the process. The selling price used for the different products involved are as follows: bioethanol 1.3 $/L [51], SA 3 $/kg [42] and fertilizer 0.46 $/kg [52]. On top of that, the remaining purified CO_2, which is not utilized for SA production, represents economic value that improves the overall revenue of the process. Thus, the surplus CO_2 is recognized as a by-product, at price of 0.035 $/kg [42]. Financial feasibility of the plant is evaluated by employing diverse profitability metrics, encompassing net present value (NPV), internal rate of return (IRR), return on investment (ROI), payback period (PBP), and minimum selling price (MSP). The MSP is ascertained via discounted cash flow (DCF) analysis, signifying the market price necessary to attain the NPV of zero at the culmination of the facility's operational life. To account for economic uncertainties, a sensitivity analysis is carried out, focusing on key parameters such as BW price, steam price, chemicals price, bioethanol price, and SA price. This analysis involved varying one parameter while

Table 1
Discounted cash flow analysis parameters.

Parameter	Value
Plant life	30 years
Tax rate	34% [6]
Discount rate	10%
Equity financing	100%
Depreciation	MACRS and 10 years recovery period
Plant construction period	3 years
First 12 months' expenditures	8%
Next 12 month's expenditures	60%
Last 12 month's expenditures	32%
Working capital	10%

Table 2
Cost of raw materials and utilities.

Parameter	Value	Unit cost	Reference
BW	100	$/MT	[6]
Enzymes	3.08	$/kg	[6]
Yeast extract	0.0052	$/kg	[38]
Peptone	0.0052	$/kg	[38]
Ammonium sulphate	0.1	$/kg	[50]
Magnesium sulphate	0.18	$/kg	[50]
Potassium phosphate	1.2	$/kg	[50]
Process water	0.177	$/1000 L	[42]
$MgCO_3$	0.5	$/kg	[42]
NaOH	0.4	$/kg	[42]
Cooling water	0.032	$/L	[6]
Low pressure steam	0.018	$/kg	[6]
Medium pressure steam	4.77	$/GJ	[42]
Refrigerant	4.77	$/GJ	[42]
Electricity	0.077	$/kWh	[6]

keeping the others constant to understand their impact on the NPV of each scenario.

2.3. Environmental assessment

LCA is carried out in accordance with ISO 14040 standards, encompassing the establishment of objectives and boundaries, execution of an inventory analysis, assessment of environmental impacts, and the subsequent interpretation of findings.

2.3.1. Goal and scope

The primary aim of this study is to evaluate the environmental performance of a biorefinery that utilizes BW as feedstock to produce bioethanol and SA. The key objective is to compare various scenarios for the combined production of bioethanol and SA from an environmental perspective. To facilitate this comparison, the functional unit chosen for assessment is 1 kg of BW, enabling the evaluation of different multi-product scenarios using the same feedstock. In terms of system boundaries, our assessment adopts a "cradle-to-gate" approach. This methodology encompasses a segment of the product's life cycle, extending from the initial resource extraction phase ("cradle") to the moment it exits the factory premises ("gate"). The LCA system boundaries encompass several processes, including enzymatic hydrolysis, ethanol fermentation, AD, CHP, SA production, and the CC unit, which are depicted in supplementary materials (see Fig. A8).

2.3.2. Life cycle inventory (LCI)

During the inventory analysis phase, the input and output data for all sub-processes are described, and potential allocation methodologies are discussed. Mass and energy inputs and outputs (inventories) are estimated through the process design simulations. Table 3 presents the inventories for all the four scenarios. In this study, a 'system expansion' approach is employed, where electricity generated in the CHP unit receives an 'avoided emissions' credit, as it substitutes for an equivalent amount of electricity from the average UK national grid production. Similarly, the environmental benefit of SA production through CO_2 utilization is also considered.

2.3.3. Life Cycle Impact Assessment (LCIA)

Life Cycle Impact Assessment (LCIA) is carried out using the CML-IA method, employing SimaPro v9.4.0.2 LCA software. The CML-IA baseline method estimates the potential environmental impacts at midpoint level, showing results through 8 impact categories, including Abiotic depletion (ADP), Abiotic depletion (fossil fuels), Global warming (GWP), Ozone layer depletion (ODP), Human toxicity (HTP), Photochemical oxidation (PCOP), Acidification (AP), and Eutrophication (EP). The CML-IA baseline is recognized as one of the most consistent methods for life cycle impact assessment in this research area [53,54]. In addition, this approach organizes the results into midpoint categories based on common mechanisms (e.g., climate change) or widely accepted groupings (e.g., ecotoxicity) [55].

Table 3
LCI of the foreground system of bioethanol production from 1 kg of BW for all scenarios.

Chemical/utility	Unit	Scenario			
		1	2	3	4
Chemicals					
Enzyme, Glucoamylase	kg	0.0006	0.0023	0.0006	0.0061
Yeast extract	kg	0.0290	0.1131	0.0137	0.0316
Peptone	kg	0.0290	0.1131	0.0137	0.0316
Ammonium sulphate	kg	0.0232	0.0905	0.0110	0.0253
Magnesium sulphate	kg	0.0194	0.0754	0.0091	0.0211
Potassium phosphate	kg	0.0116	0.0453	0.0055	0.0127
Magnesium carbonate	kg	–	–	0.0740	0.7020
Sodium hydroxide	kg	–	–	0.0533	0.5057
Utilities					
Process water	kg	1.0398	4.0490	1.1806	2.1526
Cooling water	kg	41.7554	111.6422	26.4241	362.1592
Steam	kg	6.1544	10.1884	2.1053	25.9262
Electricity	kWh	–	–	–	2.4741
Chiller	kJ	–	–	17.8733	156.5443
Main product					
Bioethanol	kg	0.2416	0.2416	0.1138	
Co-products					
Fertilizer	kg	0.05	0.1965	0.0473	1.8430
Excess electricity	kWh	0.0451	0.1866	0.0377	–
Succinic acid	kg	–	–	0.2707	19.5079
Emissions					
Water	kg	0.0831	0.3236	0.0994	0.0019
Carbon dioxide	kg	0.1714	0.6676	0.2113	0.0003
Carbon dioxide biogenic	kg	0.2301	0.2301	–	–
Carbon monoxide	kg	0.0127	2.3409	0.0127	0.0127
Nitrogen	kg	0.6011	0.0495	0.7162	0.7160
Oxygen	kg	0.0305	0.1186	0.0364	0.0364
NH_3	g	0.0575	0.2238	0.0572	0.0578
N_2O	g	0.0071	0.0071	0.0071	0.0072
NOx	g	0.0071	0.0071	0.0071	0.0072

3. Results and discussion

3.1. Process simulation

The key findings of process simulation concerning mass and energy balances have been methodically organized in Table 4.

In the context of Scenario 1 and Scenario 2, a substantial yield of 1007 kg/h of bioethanol is achieved, corresponding to a notable mass yield of 241 kg/tonne originating from BW. The ethanol yield is found to be similar to the results obtained in the mentioned study of Narisetty et al. [36]. However, the ethanol yield calculated is higher than those obtained from other food waste-based feedstocks, for instance, 156 kg/ton citrus peel [56], 170 kg/ton mango peel [57], and 200 kg/ton cassava bagasse [58].Notably, the ethanol yield from BW is comparatively higher (40.6 %, dry basis) than those from lignocellulosic biomasses involving crop residues, wood and grass species, and bagasse (22.14 %, dry basis) [29]. This can be explained by variances in the levels of carbohydrates or cellulose found in these feedstocks, where BW is found to have comparatively higher carbohydrate content around 46%. In comparison, Scenario 3 and Scenario 4 generate lower yields of bioethanol 474 kg/h and 108 kg/h, respectively. This disparity can be attributed to the fact that in Scenario 1 and Scenario 2, all fermentable sugars are completely converted into ethanol, with no concurrent production of SA. The allocation of sugars for SA production in scenarios 3 and 4 is influenced by the amount of CO_2 generated in the bioethanol fermentation process and CO_2 captured in the CC unit. In this context, CO_2 assumes the role of a limiting agent, whereby the entirety of the CO_2 produced is channelled into the conversion process that yields SA.

Regarding energy consumption, the steam demand for bioethanol

Table 4
Mass and energy balance of the integration bread waste biorefinery.

		Unit	Scenario-1	Scenario-2	Scenario-3	Scenario-4
Feedstock	Bread waste	MT/day	100	100	100	100
Energy demands	Heating	MW	15.3	6.8	5.8	6.9
	Cooling	MW	26.4	18.2	16.7	23.5
	Power	MW	0.454	0.454	0.609	1.782
Product	Bioethanol	kg/hr	1007	1007	474	108
	Succinic acid	kg/hr	–	–	1128	2107
	Power generated	MW	0.641	0.641	0.767	0.767
	Net power	MW	0.188	0.188	0.158	–
	CO_2	kg/hr	–	–	–	183
	Fertilizer	kg/hr	198	198	199	199

production in Scenario 1 (54 MJ/kg bioethanol) is found to be the highest compared to other scenarios. This amount is also higher than the lignocellulosic biomass-based bioethanol production, which is reported to be 15 MJ/kg of net bioethanol derived from lignocellulosic biomass [29]. This high demand could be attributed to the pre-heating condition and downstream processing of bioethanol in the scenarios presented in this work. On the other hand, the power consumption for all scenarios translates to 0.18, 0.18, 0.25, and 0.72 kWh/kg dry BW, respectively. The energy recovery and utilization are carried out by utilizing the organic waste produced during the process.

In our study, a methane yield of 0.278 g CH_4/g volatile solid (VS) is achieved for all scenarios, which aligns closely with the findings reported by Narisetty et al. [36]. The biogas combustion in CHP generates 554.2 kWh/ton BW electricity, which is relatively higher than the average lignocellulosic biomass-based feedstock of 514.25 kWh/ton [29]. When analysing electricity generation, it becomes apparent that Scenarios 1, 2, and 3, have the potential to meet their internal requirements. In contrast, for Scenario 4, there is no surplus electricity due to the significant power demand of the CC unit, which amounts to 1.31 MW/ton of CO_2. The additional requirement of power consumption for Scenario 4 accounts for 1.02 MW. The primary contributor to the substantial power demand is the necessity for pressurizing CO_2 during the operation of the compressor and pump, which precedes the absorption process.

Redirecting the sugar hydrolysate for SA production in Scenarios 3 and 4 results in a decrease in bioethanol production. But this improves the overall efficiency of carbon utilization of the process. In Scenario 4, the SA output nearly doubles when compared to Scenario 3. This substantial increase in SA production is attributed to the integration of 717.1 kg/h of CO_2 fraction from CC unit, which can effectively consume more glucose. Here, it is important to mention that for scenario 4, not all the captured CO_2 will be used for SA production; only 80% will be directed towards SA production, while the remaining 183 kg/h of CO_2 is regarded as an additional value-added co-product.

3.2. Process integration by pinch technology

To improve the energy efficiency of the process, energy integration is implemented by recovering heat from various hot process streams to heat cold streams using heat exchangers. This assessment of energy conservation covers the entire process and identifies specific areas where it can optimize energy recovery to meet the heating and cooling goals effectively. A detailed target heating and cooling temperature for scenarios studied can be seen in supplementary materials (See Table S2, S4, and S6). Fig. 2 represents the heating and cooling consumption reduction after heat integration implementation. As can be seen, heating requirement in scenario 1 is 15.3 MW. The two most significant contributors to this value are: the pre-heating of BW slurry prior to autoclave and the purification of fermentation broth downstream, accounting for 7.2 MW and 4.3 MW, respectively. To optimize energy utilization, heat integration is implemented across scenarios 2, 3 and 4 based on pinch technology. Scenario 2, for example, achieves a notable 56% reduction in heating needs through these measures. However, the most remarkable reduction in heating requirements is observed in Scenario 3, with an impressive 62% decrease (compared to scenario 1). This improvement can be primarily attributed to the reduced bioethanol production capacity in the downstream process. Additionally, heat integration shows a substantial decrease in heating requirement in the evaporation process prior to the crystallization of SA. Conversely, the heating consumption of Scenario 4 is comparatively similar to that of Scenario 2. Despite the increased energy demands associated with the CC unit and evaporator in the SA downstream process, it is important to note that the substantially reduced energy consumption in bioethanol purification plays a significant role in achieving demand reduction.

When looking at the cooling requirements, scenario 1 leads with 26.4 MW. This cooling demand primarily arises from the necessity to lower the temperature before the enzymatic hydrolysis process and the heat duty of the condenser in bioethanol purification. On the other hand, when considering reductions in cooling consumption, Scenario 3 stands

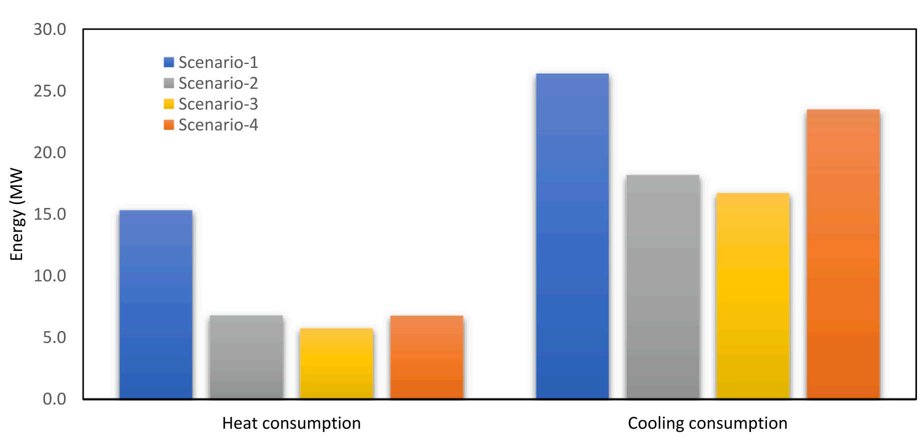

Fig. 2. Energy (heat and cooling) requirements before and after integration.

out with an impressive 37% decrease. Meanwhile, Scenario 2 exhibits significant reductions in cooling needs, registering a reduction of 31%. However, the cooling reduction in Scenario 4 accounts for 11%, which is a relatively lower reduction compared to Scenarios 2 and 3. This is due to the additional cooling energy needed in the CC process. In CC units, especially those involving gas compression, intercoolers can reduce the energy required for compression. By cooling the gas between compression stages, the intercooler decreases the temperature and, consequently, the volume of the gas. This leads to a reduction in the energy needed for subsequent compression stages. This approach has been successfully implemented by Malekli et al. [45]. Utilizing the organic Rankin cycle (ORC) and combining with air blower can effectively recover waste heat from the CC process. This recovered heat can then be used to generate additional power or assist in other process heating requirements, thus reducing the overall energy demand [41]. Zhao et al. [59] also found that incorporating an intercooler into the system leads to substantial energy savings during the regeneration process.

3.3. TEA

3.3.1. Total capital investment

A breakdown of the equipment cost contribution for all investigated scenarios is presented in Fig. 3. As can be seen, reactors represent the largest share of the total equipment cost (more than 40% in all scenarios). The percentage contributions of reactors are based on the cost of the fermenters, enzymatic hydrolysis reactors, autoclave, and anaerobic digestors. The fermenters, enzymatic hydrolysis reactors, and anaerobic digestor costs were estimated based on integer numbers of 500 m^3 reactors [6]. Heat exchangers represent the second largest contributor to the equipment cost, representing 12.64% in Scenario 1 to approximately 14% in Scenario 4. Another major contributor in all scenarios is the distillation column, contributing 13.46%, 13.22%, 8.84%, and 9.31% in scenarios 1, 2, 3 and 4 respectively. The high-cost contribution of the distillation column in Scenario 1 and 2 originates from the higher capacity of bioethanol.

Table 5 provides a breakdown of the total capital investment (TCI) for all the scenarios under study. The TCI figures stand at $41.01 million for Scenario 1, $41.77 million for Scenario 2, $53.89 million for Scenario 3, and $59.38 million for Scenario 4, respectively. As expected, Scenario 4 incurs the highest TCI compared to the other scenarios, primarily due to the extra capital investment required for the additional equipment in the SA and CC unit. On the contrary, Scenario 1 has the lowest TCI among all the scenarios. This difference can be attributed to the absence of heat integration technology in Scenario 1, which results in a smaller and less extensive set of equipment compared to the other scenarios. However, Scenario 2 has slightly higher TCI compared to Scenario 1. It is attributed to similar processes being employed, with the distinction lying in the incorporation of additional equipment for heat integration.

3.3.2. Total operating cost

A breakdown of the annual operational expenditure (OPEX) is presented in Table 6. The cumulative operating costs encompass the direct manufacturing cost, fixed manufacturing cost, and general manufacturing cost. OPEX across all scenarios spans a range from 18 M$/year to 23.7 M$/year. Scenario 4 has the highest yearly operating expenses at 25.9 M$ compared to other scenarios. This is primarily due to the increased costs associated with obtaining a larger SA capacity,

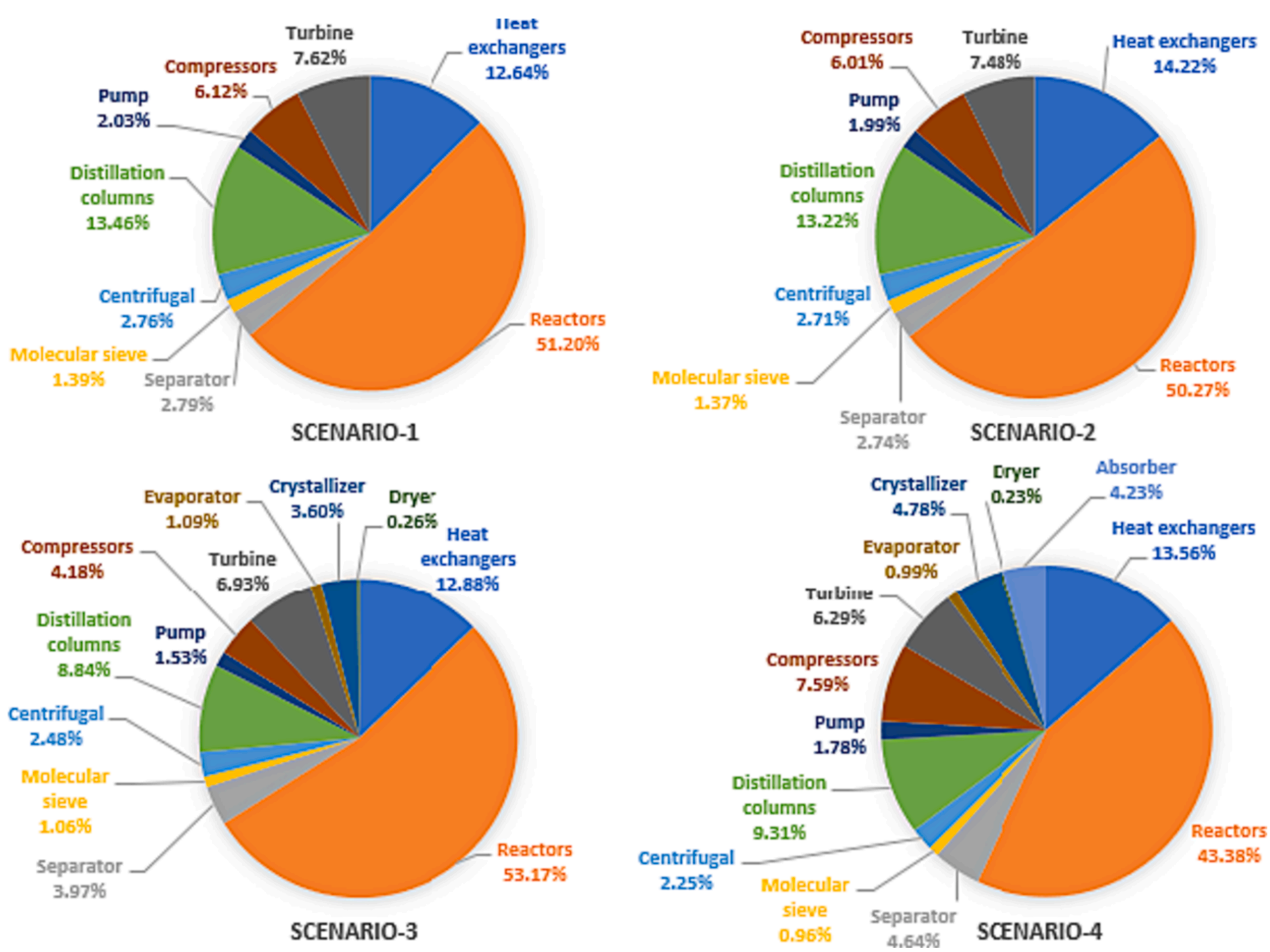

Fig. 3. Contribution of each equipment cost for all scenarios.

Table 5
Capital investment for all scenarios.

Summary of fixed capital estimates	Scenario-1	Scenario-2	Scenario-3	Scenario-4
A. total plant direct cost				
1. Equipment purchase cost (EPC)	$9,062,436	$9,229,336	$11,908,502	$13,120,602
2. Equipment installation	$3,534,350	$3,599,441	$4,644,316	$5,117,035
3. Instrumentation and control	$1,178,117	$1,199,814	$1,548,105	$1,705,678
4. Process piping	$2,809,355	$2,861,094	$3,691,636	$4,067,387
5. Electrical equipment	$906,244	$922,934	$1,190,850	$1,312,060
6. Buildings	$2,628,107	$2,676,508	$3,453,466	$3,804,975
7. Site development	$906,244	$922,934	$1,190,850	$1,312,060
8. Auxiliary facilities	$4,984,340	$5,076,135	$6,549,676	$7,216,331
TPDC	**$26,009,192**	**$26,488,195**	**$34,177,401**	**$37,656,128**
B. Total plant indirect cost (TPIC)				
1. Engineering	$2,899,980	$2,953,388	$3,810,721	$4,198,593
2. Construction	$3,081,228	$3,137,974	$4,048,891	$4,461,005
TPIC	**$5,981,208**	**$6,091,362**	**$7,859,611**	**$8,659,597**
C. Total plant cost (TPC)	**$31,990,400**	**$32,579,557**	**$42,037,013**	**$46,315,726**
1. Contractor's fee	$1,599,520	$1,628,978	$2,101,851	$2,315,786
2. Contingency	$3,199,040	$3,257,956	$4,203,701	$4,631,573
D. Fixed capital investment	**$36,788,960**	**$37,466,490**	**$48,342,565**	**$53,263,085**
Land use	$543,746	$553,760	$714,510	$787,236
Working capital	$3,678,896	$3,746,649	$4,834,256	$5,326,308
E. Total capital investment	**$41,011,602**	**$41,766,900**	**$53,891,331**	**$59,376,629**

Table 6
Summary of total annual operating cost for all scenarios.

	Scenario 1	Scenario 2	Scenario 3	Scenario 4
A. Direct manufacturing cost				
1. Raw material cost (C_{RM})	$4,069,631	$4,069,631	$5,726,802	$7,099,718
2. Utilities cost (C_{UT})	$5,381,090	$2,385,183	$1,951,415	$1,171,697
3. Operating labor (C_{OL})	$1,330,390	$1,330,390	$1,330,390	$1,330,390
4. Supervisory and clerical labor	$266,078	$266,078	$266,078	$266,078
5. Maintenance and repair	$2,207,338	$2,247,989	$2,900,554	$3,195,785
6. Operating supplies	$331,101	$337,198	$435,083	$479,368
7. Laboratory charges	$199,559	$199,559	$199,559	$199,559
8. Patents and royalties	$766,718	$661,860	$795,960	$861,584
Total direct manufacturing cost	**$14,551,904**	**$11,497,890**	**$13,540,732**	**$14,604,179**
B. Fixed manufacturing cost				
1. Local taxes and insurance	$1,103,669	$1,123,995	$1,450,277	$1,597,893
2. Plant overhead	$2,266,319	$2,290,710	$2,682,249	$2,859,387
Total fixed manufacturing costs	**$3,369,988**	**$3,414,705**	**$4,132,526**	**$4,457,280**
C. General manufacturing cost				
1. Administration costs	$566,580	$572,678	$670,562	$714,847
2. Distribution and selling costs	$2,811,299	$2,426,821	$2,918,521	$3,159,142
3. Research and development	$1,277,863	$1,103,100	$1,326,600	$1,435,973
Total general manufacturing costs	**$4,089,162**	**$3,529,921**	**$4,245,121**	**$4,595,115**
Total operating cost ($/year)	**$22,011,053**	**$18,442,515**	**$21,918,378**	**$23,656,574**

resulting from higher raw material expenses. On the other hand, no SA production and pinch-technology based energy savings lead to lowest operating cost (18.4 M$/year) for Scenario 2. Reducing utility expenses leads to decreased operational costs. In Scenario 1, utility cost makes up a significant portion, accounting for 24% of the overall operating expenses. For Scenario 2, 3, and 4, the primary driver of costs is the expenditure on raw materials. Among these factors, the cost associated with BW stands out as the most influential. It is essential to recognize that the pricing of BW feedstock directly impacts both the costs related to its collection and its transportation to the biorefinery facility [6]. Additionally, other cost factors such as labour, maintenance, supplies, plant overhead, and distribution and selling expenses make up a significant part of the overall operating costs. These costs are predominantly determined by the FCI calculation [60].

Table 7
Profitability analysis of all scenarios.

Parameter	Scenario 1	Scenario 2	Scenario 3	Scenario 4
NPV (M$)	$-116,611,187	$-83,777,981	$41,583,981	$163,179,389
IRR (%)	–	–	18	33
ROI (%)	−28	−19	11	32
PBP (years)	−13.9	−17.7	10.6	2.2
MSP of bioethanol ($/Liter)	2.55	2.21	0.17	–
MSP of succinic acid ($/kg)	–	–	2.40	1.72

3.3.3. Profitability analysis

The profitability analysis under the base case evaluation of the four scenarios is summarized in Table 7 and the cumulative cash flow diagram over the years is presented in Fig. 4. The profitability of the process is estimated based on IRR, ROI, PBP, NPV, and MSP. As can be seen, the NPV and PBP show a positive value in Scenario 3 (42 M$ and 10.5 years) and Scenario 4 (163 M$ and 2.2 years). This result suggests that scenarios 3 and 4 bring substantial economic value to the biorefinery and have significant potential for generating high financial returns. On the other hand, Scenarios 1 and 2 fail to achieve payback, exhibiting negative NPV values of −116 M$ and –83 M$. The NPV is calculated by aggregating the discounted cash flows over the project's lifetime. This involves discounting the projected cash flows of the project to their

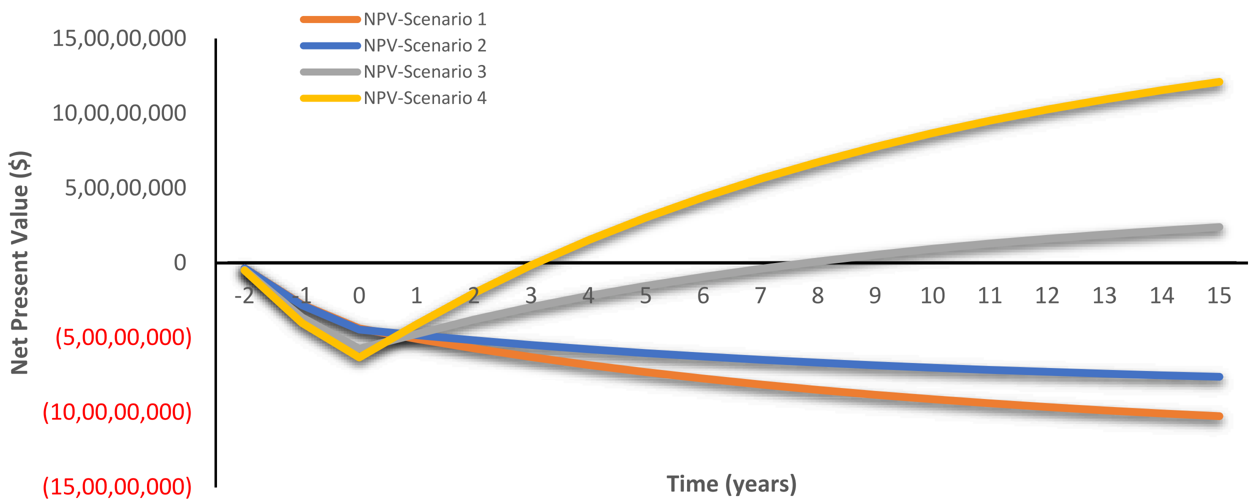

Fig. 4. Cumulative cash flow diagrams for the four scenarios.

present value using the discount rate, and then summing these values. The negative NPV for these scenarios indicates their economic infeasibility, stemming from high capital and operational expenditures that outweigh the revenues generated. In general, Scenario 4 stands out as the most profitable option, as it outperforms the others across all profitability indicators. Consequently, it is deemed an optimistic and highly competitive scenario. The impacts of other economic parameters are further discussed in the sensitivity analysis.

The calculated MSP of bioethanol for Scenario 1 and Scenario 2 are nearly double (2.55 $/L and 2.21 $/L) compared to the current market price of bioethanol (1.3 $/L) [51]. MSP of bioethanol from lignocellulosic biomass feedstocks has been reported involving switchgrass (1.07 $/L) [50] and sugarcane (1.23 $/L) [61]. It should be noted that a significant reduction of MSP can be achieved in Scenario 3 (0.17 $/L). This result can be correlated to the larger SA production and revenue of SA contributes significantly to the total of revenue. Furthermore, the co-production of SA and CO_2 in Scenario 4 is found economically favourable to the production of bioethanol. However, the substantial additional revenue generated by SA constitutes a significant contribution to the MSP of bioethanol, resulting in an MSP value less than 0 in Scenario

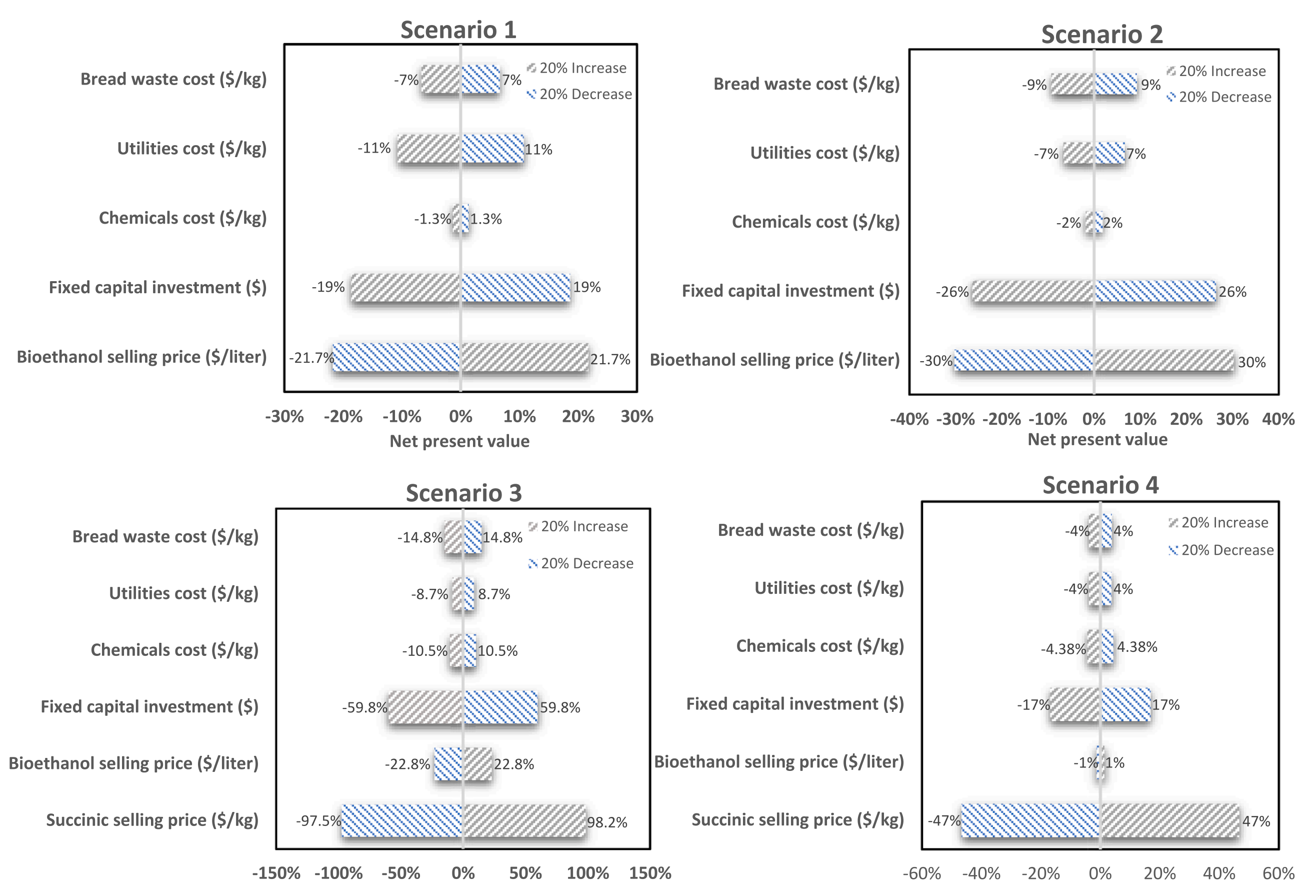

Fig. 5. Sensitivity analysis for different scenarios.

4.

3.3.4. Sensitivity analysis

A sensitivity analysis is conducted to assess the influence of critical economic variables, with a focus on identifying the most sensitive parameters. It is observed that BW cost, utilities cost, chemicals cost, FCI, bioethanol selling price and SA selling price have the biggest impact on the net present value (NPV). The sensitivity analysis involved varying these input parameters by +/−20% of their base values and studying the influence on NPV as illustrated in Fig. 5.

This range has been selected as it aligns with the preliminary estimates provided in this range [6]. In Fig. 5, it is evident that the FCI cost played a significant role in all the scenarios under consideration. Nevertheless, it was the fluctuation in bioethanol selling price that exerted the most substantial influence on the NPV in scenarios 1 and 2. In contrast, scenarios 3 and 4 were primarily shaped by changes in the SA selling price.

When the bioethanol selling price was decreased by 20%, it resulted in a 21.7% reduction in NPV for scenario 1 and a substantial 30% drop in NPV for scenario 2. Conversely, a 20% reduction in FCI cost led to a 19% NPV increase in scenario 1 and a 26% NPV increase in scenario 2. It is noteworthy that variations in chemicals costs had the least impact on NPV across all the scenarios examined.

For scenarios 3 and 4, SA price emerged as the most influential factor. A 20% decrease in SA price had a significant impact, causing NPV to plummet by 97.5% in scenario 3 and by 47% in scenario 4. Interestingly, bioethanol price did not significantly affect NPV in scenarios 3 and 4, largely due to its diminished production capacity in these particular scenarios. Conversely, the increment of SA price with 20% resulted in a substantial enhancement of NPV for both Scenarios 3 and 4, with improvement of 98.2% and 47%, respectively.

3.4. LCA

The results obtained from the LCA analysis across different impact categories are shown in Table 8. Negative values in the table signify environmental savings attributed to the avoidance of primary product and co-products' impacts generated by the systems under study. To analyze these results closely, the percentage contributions of the different components are illustrated in Fig. 6 for all scenarios.

The LCA results unveil significant variations in GWP across different scenarios. Specifically, Scenario 1 demonstrates the highest GWP, totalling 1.739 kg CO_2-eq/kg BW. This significant GWP score primarily arises from the substantial contribution of steam consumption during the pre-heating of pretreatment and purification processes, accounting for more than 95% of the impact. Additionally, emissions from flue gas generated by the CHP system and gaseous emissions resulting from fertilizer application contribute significantly, comprising 10% of the GWP impact.

Scenario 1 features a notable positive impact: the excess electricity generated results in a 0.8% decrease in overall environmental impacts and the avoided use of fertilizer represents a 0.1% reduction. Most prominently, the significant positive environmental effects are chiefly attributed to the credit from bioethanol production, capable of offsetting 17% of the total adverse impacts, thereby contributing to a meaningful reduction in the overall GWP. It is worth noting that this scenario show higher GWP scores when compared to a prior study Narisetty et al. [36], with value of 1.27 kg CO_2-eq. This disparity can be attributed to the absence of additional BW utilization to generate sufficient steam to meet the entire energy demands for bioethanol production.

Moving to Scenario 2, the adoption of pinch technology heralds a remarkable enhancement in energy efficiency, driving a 57% reduction in GWP to 0.749 kg CO2-eq. Additionally, replacing fossil-based ethanol with an alternative derived from BW presents a meaningful mitigation strategy, marking a 38% GWP reduction and showcasing a significant leap toward environmental sustainability. Scenario 3 and Scenario 4 uniquely benefit from the inclusion of co-products generated, resulting in smaller GWP scores of 0.142 and −0.344 kg CO_2-eq, respectively. The substantial negative value of GWP for Scenario 4 reflects the environmental benefits of co-product generation. This result aligns with findings of Brancoli, et al. [62], who reported the GWP value of −0.56 kg CO_2-eq/kg BW. This value is derived by Brancoli, et al. [53] after substituting the ethanol produced as vehicle fuel and accounting for co-product of dried distiller's grains with soluble (DDGS), which can substitute soybean meal and barley in animal feed. Similarly, another study performed by Sadhukhan, et al. [29], reported a low negative GWP value of bioethanol production from various lignocellulosic biomass, including rice husks, sawdust Chichicaxtla, and sawmill slabs Chichicaxtla, with values of −142.5, −128.2, and −120 kg CO_2-eq, respectively. The underlying reason for these significant negative GWP values was the increased net electricity generation associated with the utilization of lignin, which is absent in the case of BW.

In general, Scenarios 3 and 4 have shown the most favourable environmental performance. This indicates that the most effective configuration involves the co-production of SA. Notably, Scenario 4 outshines the rest across all impact categories and particularly stands out as the most sustainable configuration for three environmental indicators: Abiotic depletion (fossil fuels), GWP, and HTP. Its exceptional environmental outcomes are primarily attributed to the significant reduction in energy and CO_2 emissions by substituting fossil-based SA with a more sustainable alternative. This replacement, a key feature of Scenario 4, results in higher SA yields compared to Scenario 3, due to an additional 80% of CO_2 supply captured from the CC unit. Furthermore, it is observed that the surplus CO_2 from CC unit is considered a by-product, resulting in an additional positive environmental contribution.

4. Conclusions

This study explored the economic viability and environmental implications of four scenarios for the co-production of bioethanol and SA from BW. From the economic perspective, Scenario 4 emerges as the most attractive, displaying the highest NPV of 163 M$, followed by Scenario 3 with 42 M$. Scenario 1, on the other hand, is the least attractive, featuring an NPV of −116 M$. Notably, the production of byproducts like excess electricity, fertilizer, SA, and CO2 bolsters market competitiveness by reducing the MSP of bioethanol. Furthermore, the utilization of the CC unit implemented in Scenario 4 to capture CO_2 from flue gas is found to generate additional revenue by producing more SA.

Table 8
Life cycle impact assessment of the investigated scenarios.

Impact category	Unit	Scenario 1	Scenario 2	Scenario 3	Scenario 4
Abiotic depletion potential (ADP)	kg Sb eq	−1.69E−07	−1.75E−07	−1.40E−06	−2.92E−06
Abiotic depletion (fossil fuels)	MJ	15.546	2.272	−9.548	−16.2
Global warming potential (GWP)	kg CO_2 eq	1.739	0.749	0.142	−0.344
Ozone layer depletion potential (ODP)	kg CFC-11 eq	3.38E−09	3.30E−09	−8.61E−08	−1.60E−07
Human toxicity potential (HTP)	kg 1,4-DB eq	0.046	0.037	−0.122	−0.3
Photochemical oxidation potential (PCOP)	kg C_2H_4 eq	0.001	2.71E−04	2.85E−04	3.49E−04
Acidification potential (AP)	kg SO_2 eq	0.014	0.006	0.003	0.0027
Eutrophication potential (EP)	kg PO_4 eq	4.77E−04	1.51E−04	−1.43E−04	−2.96E−04

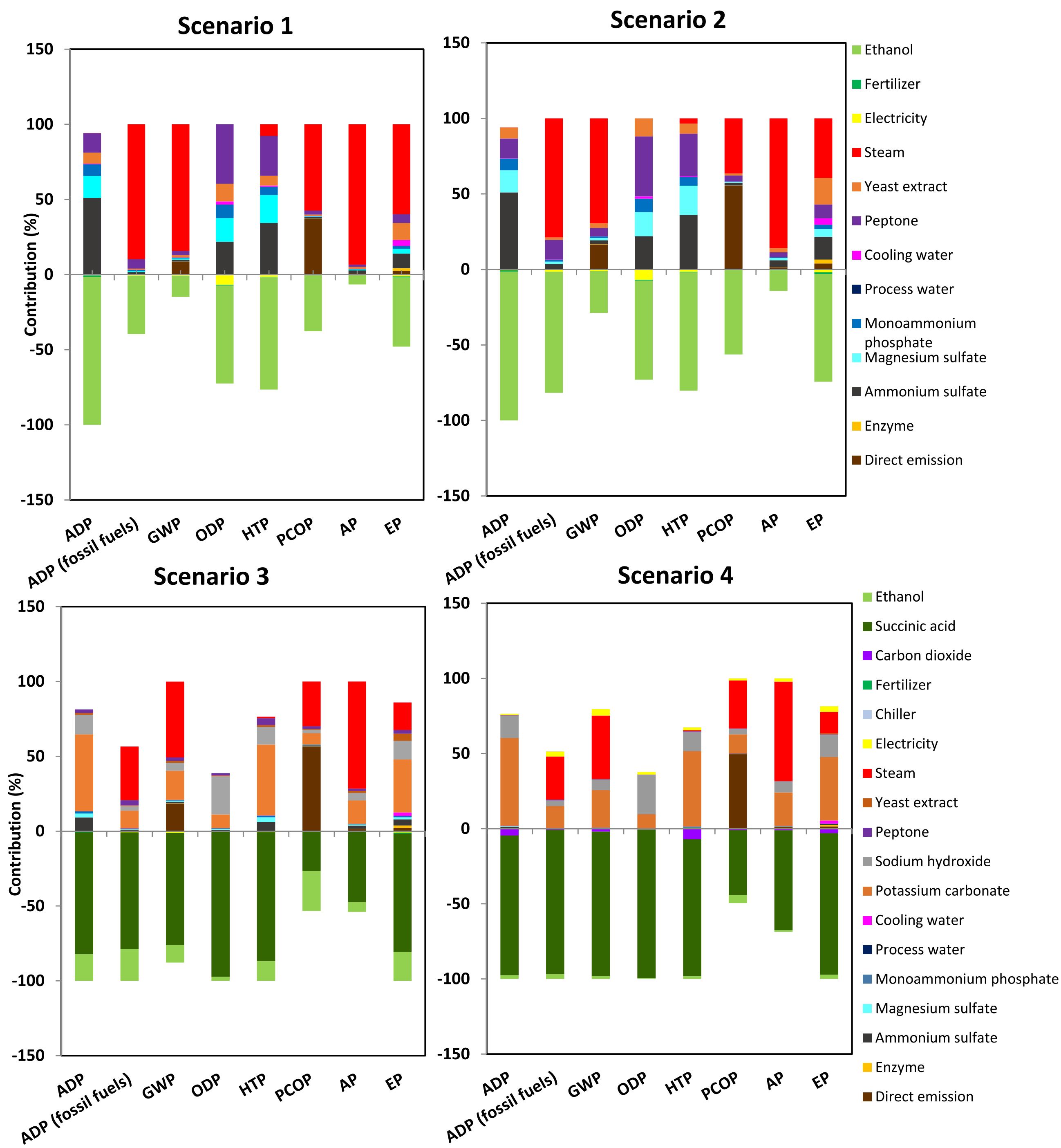

Fig. 6. Contribution analysis of different environmental impact categories for the four scenarios.

Another avenue of economic enhancement came from the application of pinch technology, which significantly curtailed both heating and cooling consumption. Specifically, Scenario 3 showcased the most substantial reduction in heat and cooling demands at 62% and 37%, respectively, resulting in considerable cost savings on utility expenses.

From an environmental perspective, the byproducts generated, markedly diminish the environmental burden. Scenario 4 reveals the lowest GWP score with −0.344 kg CO_2-eq/kg BW compared to other scenarios. LCA results identified heat consumption as a primary source of impact, underscoring that reducing heat consumption in the biorefinery could significantly curb GHG emissions. The study illustrates the advantages of integrating the co-production of SA through CO_2 utilization. Such strategies not only present the potential for achieving negative emissions and substantial economic gains but also serve to facilitate the implementation of CC technology.

CRediT authorship contribution statement

Rendra Hakim Hafyan: Methodology, Resources, Formal analysis, Data curation, Visualization, Writing – original draft, Writing – review & editing. **Jasmithaa Mohanarajan:** Writing – original draft, Data curation. **Manaal Uppal:** Writing – original draft, Data curation. **Vinod Kumar:** Writing – review & editing. **Vivek Narisetty:** Writing – review & editing. **Sunil K. Maity:** Methodology, Writing – review & editing. **Jhuma Sadhukhan:** Writing – review & editing. **Siddharth Gadkari:** Funding acquisition, Methodology, Resources, Formal analysis, Data curation, Visualization, Writing – original draft, Writing – review & editing.

Declaration of competing interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

Data availability

Data will be made available on request.

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Appendix A. Supplementary material

Supplementary data to this article can be found online at https://doi.org/10.1016/j.enconman.2023.118033.

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