



Potato peel integrated biorefinery design for succinic acid and bioethanol production: Technoeconomic and environmental analysis

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ABSTRACT

The increasing demand for energy and chemical products, coupled with the depletion of fossil resources, underscores the need for sustainable biorefinery alternatives. Potato peel (PP), an abundant agro-industrial residue, represents a promising feedstock due to its valuable biochemical composition. This study evaluates two PP valorization strategies: (i) Scenario 1, involving the microbial fermentation for the production of succinic acid (SA), and (ii) Scenario 2, a coupled system integrating bioethanol production where CO₂ emitted during bioethanol fermentation is reutilized as a substrate for SA synthesis. A techno-economic analysis (TEA) and life-cycle assessment (LCA) were conducted using SuperPRO designer process simulation, experimentally validated yields, literature-derived kinetic data, and cost estimations based on a 30-year plant lifetime. Scenario 1 yielded a minimum selling price (MSP) of \$3.77 per kg of SA, along with a net present value (NPV) of \$81.35 million and an internal rate of return (IRR) of 11.21 %, confirming its economic feasibility. In contrast, Scenario 2 resulted in negative NPV values and significantly higher MSPs for both SA and bioethanol, indicating that economic feasibility was not achieved under the current assumptions. However, Scenario 2 exhibited superior environmental performance, achieving more than a 20 % reduction in global warming potential (GWP) and fossil resource use per metric ton of PP processed. Sensitivity analysis showed that SA market price and fixed capital investment were the dominant parameters influencing profitability in both scenarios. Overall, while Scenario 1 is the economically preferable option, Scenario 2 highlights the environmental benefits of incorporating carbon-reuse strategies into emerging biorefineries.

1. Introduction

Over the past century, crude oil has been essential for industrial development, enabling large-scale production of fuels, plastics,

additives, and high-value products such as coatings, resins, and dyes [1]. However, the rise in global energy and chemical demand, driven by population growth and industrialization, has accelerated the depletion of fossil resources causing higher costs and worsening environmental

Abbreviations: Compound Annual Growth Rate, CAGR.; Cumulative Energy Demand, CED.; Individual Free-On-Board Equipment Cost, Ceq.fob.; Chemical Engineering Plant Cost Index, CEPCI.; Cost Of Manufacturing, COM.; Discounted Cash Flow, DCF.; Department Of Energy, DOE.; Discounted Payback Period, DPP.; Fixed Capital Investment, FCI.; Freshwater Eutrophication, FE.; Freshwater Ecotoxicity, FET.; Fossil Resource Scarcity, FRS.; Functional Unit, FU.; Greenhouse gas, GHG.; Global Warming, GW.; High-value, Low-Volume, HVLV.; Internal Rate of Return, IRR.; Life Cycle Assessment, LCA.; Liquid-liquid extraction, L-L.; Liquid-to-solid ratio, LSR.; Low-Value, High-Volume, LVHV.; Modified Accelerated Cost Recovery System, MACRS.; Marine Eutrophication, ME.; Marine Ecotoxicity, MET.; Minimum Selling Price, MSP.; Middle-value, middle-volume, MVMV.; Net Present Value, NPV.; Potato Peels, PP.; Presaccharification and Simultaneous Saccharification and Fermentation, PSSF.; Succinic Acid, SA.; Stratospheric Ozone Depletion, SOD.; Terrestrial Acidification, TA.; Total Capital Investment, TCI.; Techno-economic Analysis, TEA.; Terrestrial Ecotoxicity, TET.; Working Capital, WC.; Wastewater Treatment, WWT.

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impacts [2,3].

In response, research and industry have focused on renewable, sustainable alternatives to meet the demand and reduce the environmental impact of fossil-based systems [2,3]. In this context, biomass, as a carbon-neutral resource, offers a versatile platform for producing materials and chemicals traditionally derived from petroleum [1,3]. Significant efforts have been dedicated to the industrialization of biomass-based technologies, where residual biomass replaces crude oil as the primary resource [1,4]. The use of residual biomass offers several benefits, such as the efficient and cost-effective extraction of sugars without affecting food production [2]. Biorefineries, designed as integrated systems, convert biomass into multiple value-added products through advanced processes and technologies. Depending on technological maturity, these biorefineries can produce high-value, low-volume (HVLV), middle-value, middle-volume (MVMV), or low-value, high-volume (LVHV) products [3].

In this context, potato peels (PP) are a promising feedstock for biorefineries, as they are a widely available agricultural residue [5]. Potato (*Solanum tuberosum* L.) is the fourth most important food crop globally, after rice, wheat, and corn, and plays a key role in the food industry. In 2023, global potato production was estimated at around 380 million MT [5–9]. The demand for processed potato products has been steadily increasing, as these are considered convenience foods requiring minimal preparation time. In fact, the processed potato market was valued at \$40.08 billion in 2024 [8,10]. Potatoes processing generates substantial amounts of PP waste, ranging from 70 to 140 million MT annually, accounting for 15–40 % of the original mass of the processed product [5–8]. This waste, if not managed properly, decomposes and releases methane, a potent greenhouse gas contributing to ecosystem degradation [8].

Therefore, developing sustainable strategies to valorize this waste is crucial, both to mitigate environmental impacts and to reduce economic losses in the potato processing industry [7]. To date, most PP are either discarded or converted into low-value products like animal feed and fertilizers. However, due to their composition rich in carbon-based compounds such as starch, cellulose, hemicellulose, proteins, and lipids, as well as dietary fiber and essential minerals like calcium and potassium, PP could be an attractive precursor for the production of carbon-based materials within the framework of the biorefinery concept [5,6,11]. Recent research has focused on producing biogas, extracting polyphenols, and obtaining chemicals such as succinic acid (SA) and bioethanol [8].

SA is classified as an MVMV product and ranks among the top ten platform chemicals [3,12,13]. In fact, the U.S. Department of Energy (DOE) has recognized SA as a compound of strategic importance in the renewable-based chemical industry [3]. Also, SA can be obtained by microbial fermentation of sugars from plant biomass [13] and its production from renewable raw materials offers a sustainable alternative to petroleum-based feedstocks like n-butane [12]. This interest stems from the potential use of SA in a wide range of applications, including plasticizers, food additives, polyurethanes, cement additives, detergents, resins, cosmetics, coatings, antibiotics, vitamins, and other high-value industrial chemicals such as 1,4-butanediol, γ -butyrolactone, and 2-pyrrolidone [3,12,14].

On the other hand, bioethanol is a high-octane, biodegradable biofuel, that emits less CO₂ than fossil fuels and reduces the risk of ozone formation [15]. In addition, owing to its role as an alternative energy source, global bioethanol production is expected to grow by about 28 % by 2026 [16]. It can be produced from various renewable, carbohydrate-rich feedstocks, which can be hydrolyzed into fermentable sugars [4]. However, first-generation bioethanol production, which relies on food crops, raises concerns about long-term sustainability [2]. Therefore, sustainable biofuels must provide a net positive energy balance and be produced without compromising food supplies or human resources [4]. In this context, the use of non-food biomass, such as PP, emerges as a promising solution [2]. Nevertheless, in bioethanol

production, approximately one-third of carbon is converted by microorganisms in CO₂, thus limiting the theoretical bioethanol yield to 0.51 g/g. Interestingly, the CO₂ generated during this process could be used as a residual feedstock for SA biosynthesis, improving overall system efficiency [17].

Considering all the above, it is worth recalling that although CO₂ is not strictly classified as a pollutant [18], its increasing concentration contributes significantly to global warming and ocean acidification. Since the late 19th century, the global average temperature has risen by approximately 1.18 °C [4,19], and the associated economic costs are estimated to exceed \$1.2 trillion annually [4]. These facts highlight the need to integrate innovative processes that reduce environmental impacts and improve overall system efficiency. In this context, coupling bioethanol and SA production represents an opportunity to valorize carbon emissions while reducing processing costs.

To assess the feasibility and environmental benefits of emerging processes, such as integrated biorefinery, it is essential to apply tools like techno-economic analysis (TEA) and life cycle assessment (LCA) during the early stages of development. TEA evaluates economic performance, scalability, and key parameters through sensitivity analyses [20,21], while LCA quantifies environmental impacts across the entire product life cycle, identifying hotspots and guiding improvements [20–22]. Their combined application provides a holistic framework to balance economic and environmental trade-offs, thereby supporting the design of optimized valorization pathways [20].

Within this framework, the main objective of this study is to compare two scenarios designed to maximize the valorization of PP into high-value products. Scenario 1 focuses on the production of only bio-succinic acid, whereas Scenario 2 integrates bioethanol production with the reutilization of the released CO₂ for SA biosynthesis, effectively coupling both conversion routes

2. Materials and methods

Fermentative process modelling to produce SA and bioethanol from PP of the Kennebec variety, and the energy and mass balances of these processes were conducted using the SuperPro designer® software version 13.0 [23]. SuperPro Designer has been widely employed for similar biorefinery simulations to obtain diverse products using different raw materials [24–26]. The approach of this work aimed to establish mass and energy balances to support the economic and environmental evaluation of the processes. Simulation inputs were primarily sourced from prior research conducted by the authors at a laboratory and pilot plant scale [27,28] and supplemented with information from technical literature, applying realistic assumptions, when necessary [13,29–37].

The modelling encompassed two biorefinery integration scenarios, each designed for a processing capacity of 2000 MT of PP per batch (composition detailed in Table 1), with an assumed total of 100 batches per year, representing an annual throughput of 200 kilometric tons of potato peel. The number of batches was determined by considering both the duration of the processing stages involved and the annual availability of potato peel in the market. Specifically in Spain, where a potential plant location is envisaged, annual potato production reached

Table 1
Potato peel composition.

Component	Mass fraction (%)
Total glucan*	56.10
Moisture	7.74
Ash	7.27
Hemicelluloses**	6.48
Lignin	2.05
Other solids***	20.35

* 92 % is starch and the rest are other glucans.

** Xylan, Arabinan, Acetyl Groups and Uronic Acids.

*** Proteins, extracts and minerals.

1.97 million tonnes in 2024, and at the global level, approximately 50 % of harvested potatoes are directed to industrial processing, generating millions of tonnes of potato peel residues each year [38,39].

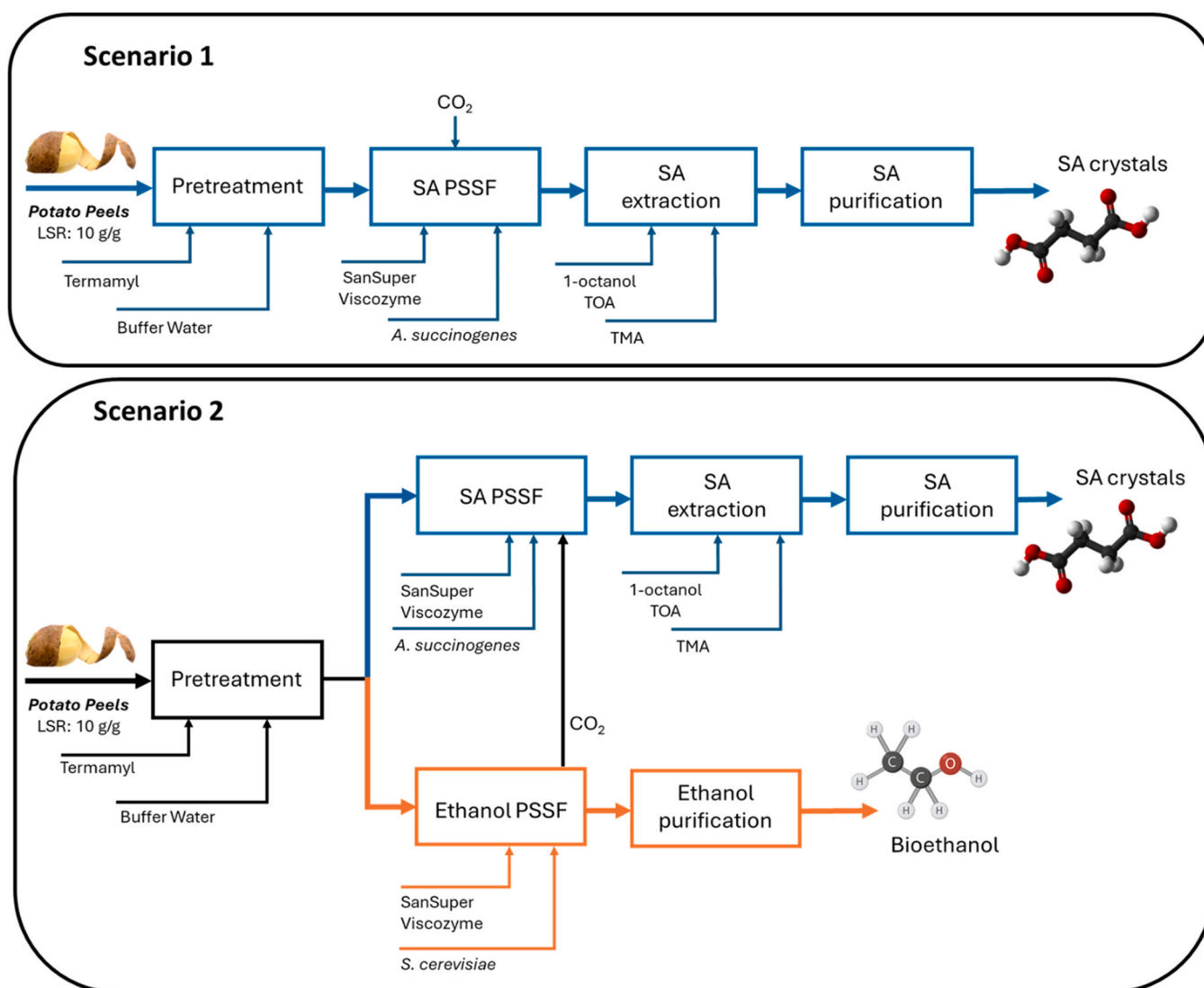
These scenarios were designed to maximize the integral valorization of PP while minimizing carbon emissions and maintaining economic feasibility. Fig. 1 provides a simplified schematic of the proposed process flow diagrams for Scenario 1 and Scenario 2, illustrating the main stages: pretreatment (including physical and enzymatic conditioning), presaccharification and simultaneous saccharification and fermentation (PSSF), liquid-liquid extraction (L-L), and final separation and purification stages:

- **Scenario 1:** In this scenario, PP undergoes an initial pretreatment consisting of mechanical grinding to reduce particle size, followed by a liquefaction stage with enzymatic treatment. The process then proceeds to the PSSF stage, where fermentable sugars are consumed by *Actinobacillus succinogenes* to produce SA. Subsequently, SA is recovered from the fermentation broth via L-L and further subjected to separation, crystallization, and purification steps to obtain high-purity SA crystals (99.5 %).
- **Scenario 2:** This scenario follows the same general scheme for obtaining SA crystals. However, the 40 % of the output stream from

the liquefaction tank is diverted to PSSF bioethanol section, where glucose is metabolized by *Saccharomyces cerevisiae* to produce this final product. The selection of a 40 % bypass of the liquefaction stream was made to ensure the optimal utilization of the available CO₂, as shown in Table 2. The resulting bioethanol-enriched stream is then directed to a final recovery and purification stage to obtain high-purity bioethanol (99.5 %). The amount of glucose routed to the SA fermentation process is determined based on the quantity of CO₂ generated during bioethanol fermentation, which serves as the limiting factor for the process. The main objective of this scenario is

Table 2
CO₂ mass balance in MT for 2000 MT of PP.

Fermentation process individually			
	Input	Recirculated	Output
Bioethanol PSSF	0.00	-	437.6
SA PSSF	281.6	0.00	0.00
Scenario 2 (40 % for bioethanol)			
	Input	Recirculated	Output
Bioethanol PSSF	0.00	-	175.04
SA PSSF	0.00	175.04	6.08



LSR: Liquid-to-solid ratio, PSSF: Presaccharification and Simultaneous Saccharification and Fermentation, SA: Succinic Acid, TOA: Tri-n-octylamine, TMA: Trimethylamine.

Fig. 1. Simplified diagram of the biorefinery scenarios 1 and 2 proposed for the potato peel valorization.

to supply the CO₂ gas generated in bioethanol production to the SA fermentation process.

2.1. Process description

The following sections provide a detailed description of the main stages illustrated in Fig. 1 for both scenarios. Additionally, Figures S.1 and S.2 in the supplementary material show the flowsheets designed for Scenario 1 and Scenario 2, respectively.

2.1.1. Pretreatment

The feedstock for the process consists of PP waste (composition detailed in Table 1). Initially, the particle size is reduced to approximately 2 mm to increase the surface area for subsequent enzymatic action. This size reduction is performed at 25 °C and 1 bar in a grinder operation (GR-101). The milled PP is then mixed with a citric acid buffered (pH 4.85) aqueous stream at a liquid-to-solid ratio (LSR) of 10 g/g, heated to 90 °C (HX-103), and pumped by a centrifugal pump (PM-101) into a liquefaction tank (R-101), where Termamyl® SC 4X alpha amylase (with an amylolytic activity of 78405 EU/mL) catalyzes complete starch liquefaction under the set conditions, rendering the starch fully accessible for the subsequent enzymatic hydrolysis stage [27,28]. The enzyme loading of the commercial Termamyl® SC 4X used in this study was 75 EU/g PP. This stage is identical for both scenarios.

2.1.2. Presaccharification and simultaneous saccharification and fermentation of SA

Prior to the PSSF stage, the liquefied stream is cooled to 50 °C (HX-104) and pumped (PM-103) into the PSSF SA reactor (R-102). In this equipment, saccharification of starch occurs first, producing fermentable sugars at 50 °C over a period of 28 h. To achieve this, the amyloglucosidase enzyme SAN Super® 360 L (amylolytic activity of 4079 UE/mL) with an enzyme load of 7.5 UE/g PP and mixture of beta-glucanases, pectinases, hemicellulases and xylanases Viscozyme® L from *Aspergillus* sp. (amylolytic activity of 4348 U/mL and 150 U/g PP of load) are introduced, resulting in an enzymatic hydrolysis yield of 98 % [27,28]. Subsequently, the temperature of the same reactor is reduced to 37 °C, and to begin the fermentation a gaseous CO₂ stream (sourced either commercially or from another process, depending on the scenario) along with the bacterium *A. succinogenes* are introduced (5 % v/v). In addition to SA, acetic acid and formic acid as main byproducts are also produced. The total fermentation time was 48 h, yielding a final productivity of 0.88 g of SA per gram of glucose consumed [27]. Following fermentation, a flash separator (V-102) operating at 90 °C was employed to maximize water evaporation. Subsequently, centrifugation (DC-101) was performed to remove the spent solids, classified as wastewater treatment (WWT) waste due to their elevated moisture content.

2.1.3. SA extraction

To separate the SA obtained from the fermentation broth, a reactive extraction process was implemented, designed as a cascade system comprising an extraction reactor (R-103) followed by a settler (V-103) [33,40]. The extractant used was Tri-*n*-octylamine (TOA) diluted in 1-octanol, with mass fractions of 8 % and 92 %, respectively [13,31], due to its high affinity for SA. The extraction was carried out at 25 °C using a 1:1 vol ratio of solvent to fermentation broth [32]. Then, in the extraction reactor, a succinic acid-tri-*n*-octylamine (SA-TOA) complex is formed and remains dissolved in 1-octanol.

2.1.4. SA crystallization and purification

After the extraction, during the decantation step, the organic complex formed (SA-TOA) is separated from the aqueous phase corresponding to the fermentation broth. The organic stream, containing the SA-TOA complex, is subsequently directed to a back extraction stage in

a secondary reactor (R-104), where trimethylamine (TMA) is added, leading to the formation of a new SA-TMA complex. In the associated settler (V-105), the organic phase, composed of 1-octanol and TOA, is recirculated at 95 % to the initial reactive extraction stage in order to reduce its economic and environmental impact [33,41].

The outlet stream from the decanter, containing the SA-TMA complex, is fed into a crystallizer (CR-101), where SA crystals are obtained at 20 °C (HX-101) over 24 h, with a yield of 90 % [33,42]. The resulting stream is subsequently directed to an evaporator (V-104), where the 95 % of TMA is recovered and recycled, in the same way as implemented for TOA and 1-octanol, while the SA crystals are sent to a dryer (RDR-101) at 100 °C for 5 h [31,32], resulting in a crystal purity of 99.5 wt% [13,30,33].

2.1.5. PSSF of bioethanol

As described in Section 2.1.2 for SA fermentation, 40 % of the liquefied stream, previously cooled to 50 °C, is pumped into the bioethanol PSSF reactor (R-105). The operating conditions and enzymes used during the pre-saccharification step in this unit are the same as those described in the corresponding section for SA production [27,28]. After a pre-saccharification period of 10 h, fermentation begins in the same reactor by lowering the temperature to 34 °C and inoculating *S. cerevisiae* at a concentration of 5 mg fresh yeast per gram of total solution. In addition to bioethanol, CO₂ was also produced as by-product. The total fermentation time was 54 h, and the final PSSF yield reached 72 % [28].

2.1.6. Bioethanol downstream purification

The stream obtained after the PSSF operation is centrifuged to separate the exhausted solids (DC-102). The resulting bioethanol-rich stream is then purified using two distillation columns in series (C-101 and C-102) until its azeotropic concentration of 92.5 % (w/w) is reached. Subsequently, bioethanol dehydration is carried out by pressure swing adsorption using molecular sieves (GAC-101), achieving a fuel-grade bioethanol purity of 99.5 % (w/w) [34,35].

2.2. Techno-economic evaluation

2.2.1. Estimation of Total capital investment (TCI)

Material and energy balances were performed using the previously mentioned SuperPro Designer software, considering a feedstock input of 2000 MT/batch of PP and a producing capacity of 100 batches/year. The equipment sizing was also estimated using this software, and the individual free-on-board equipment cost (Ceq.fob) was determined based on parameters reported in established textbooks and bibliographic references [43–46], and a Chemical Engineering Plant Cost Index (CEPCI) of 798.8 (June 2024) was used [47]. Given the technological novelty of the biorefinery and the use of costly materials that introduce a higher level of risk, the Fixed Capital Investment (FCI) was calculated by multiplying the total sum of Ceq.fob by a Lang factor of 3.2 [48]. The working capital (WC) was assumed to be 5 % of the FCI [48,49], and the final TCI was calculated as the sum of the FCI and WC.

2.2.2. Estimation of manufacturing cost (COM)

The calculation of the COM was carried out considering the previously calculated FCI, the manufacturing labor expenses (C_{OL}), the utility expenses (C_{UT}), the raw material expenses (C_{RM}) and the waste treatment expenses (C_{WT}), according to the following Eq. 1:

$$COM = 0.18 \cdot FCI + 2.73 \cdot C_{OL} + 1.23 \cdot (C_{UT} + C_{RM} + C_{WT}) \quad (1)$$

The C_{OL} values were obtained following the methodology reported by Turton et al., (2012), where the operating labor requirement for chemical processing plants was estimated using the following Eq. 2:

$$N_{OL} = (6.29 + 31.7 \cdot P^2 + 0.23 \cdot N_{wp})^{0.5} \quad (2)$$

where N_{OL} represents the number of operators per shift, P denotes the number of processing steps involving the handling of particulate solids, and N_{np} corresponds to the number of nonparticulate processing steps. For labour cost calculations, it was assumed that the plant operates 365 days per year, including maintenance activities, with three shifts per day. Each employee is considered to work five shifts per week for 49 weeks per year. The annual salary per operator was assumed to be \$50,000/year, as reported by [25,50]. Additional details are provided in Table S1 of the Supplementary Material.

On the other hand, C_{UT} , C_{RM} y C_{WT} were calculated by multiplying the utility, raw material, and waste streams obtained from the material and energy balances by their respective unit costs. In this study, PP, the primary raw material, was assumed to be obtained at no cost, based on the premise that the biorefineries are co-located with potato-processing industries [51,52]. Additional details regarding these costs are provided in Tables S2, S3 and S4 of the Supplementary Material.

2.2.3. Economic analysis and indicators

To assess the economic feasibility of the proposed scenarios a discounted cash flow (DCF) analysis was performed. The parameters selected for this analysis were derived from biorefinery-focused literature [48,53], and included a 10 % discount rate, a 30-year plant lifetime, and 100 % equity financing. The depreciation period was set at 7 years, following the Modified Accelerated Cost Recovery System (MACRS). The plant construction was assumed to span 3 years, with capital costs distributed as 8 % in the first year, 60 % in the second, and 32 % in the final year. Additionally, the WC was estimated at 5 % of the FCI. The total revenue of the biorefinery, for both scenarios, is calculated based on the two main products generated in the process: bioethanol and SA. The selling price considered for SA is \$4.00/kg [54], while that for bioethanol is \$1.30/kg [55].

Additionally, the techno-economic indicators used to evaluate the profitability of the two proposed scenarios include the net present value (NPV), the minimum selling price (MSP), the discounted payback period (DPP), and the internal rate of return (IRR). The MSP was determined as the market price required for the NPV to reach zero at the end of the plant's lifetime (30 years). The DPP is defined as the time, after plant startup, required for the project to recover the initial FCI, considering the discounted cash flows to time zero [45].

To evaluate the influence of various parameters on the profitability of the scenarios, a sensitivity analysis was conducted. The variables analyzed included products prices, FCI, raw material costs, and utility costs. In this analysis, all other parameters were kept constant while only the variable of interest was varied.

2.3. Life cycle assessment methodology

To assess the potential environmental impacts of the two proposed biorefinery designs, the four phases established by the LCA methodology according to ISO 14040 and 14044 standards [56,57] were followed: (i) goal and scope definition, (ii) life cycle inventory, (iii) life cycle impact assessment and (iv) interpretation.

2.3.1. Definition of goal and scope

The aim of the LCA is to explore the potential environmental impacts of the valorization of potato peels through the design of two biorefineries: i) 100 % of the feedstock is used for SA production (Scenario 1), and ii) 60 % of the feedstock is used for SA production, while the remaining 40 % is directed towards bioethanol production (Scenario 2). In the latter, the CO_2 generated during bioethanol fermentation is recirculated to the SA production section to reduce external CO_2 consumption, as an improvement and circular strategy. To capture the environmental trade-off between emitting and recirculating this stream, the CO_2 is fully included in the global warming indicator and is not treated as biogenic.

An attributional LCA was conducted using a cradle-to-gate approach, which is commonly applied in biorefinery assessments [58]. This approach considers all environmental impacts associated with a product's life cycle, from the extraction of raw materials ("cradle") to the point at which the product leaves the manufacturing facility ("gate") (see Fig. 2). This approach evaluates the environmental impacts associated with the production of the biorefinery products while its main limitations are that it does not capture consequential effects and excludes impacts associated with product use and end-of-life, which are beyond the scope of this study. Additionally, to report the environmental performance of these biorefinery designs, two functional units (FU) were selected: i) a feedstock-based FU (FU1), defined as 1 t of treated potato peels, considering waste valorization as the purpose of the system; and ii) a product-based FU (FU2), defined as 1 kg of product: SA (99.5 % purity) and bioethanol (99.5 % purity), as the function of the system can also be the manufacture of both components. This allows comparison with other studies in literature using different feedstocks to obtain these products. Moreover, in Scenario 2, where SA and bioethanol are produced, both mass and economic allocation were applied to distribute the environmental burdens. For the economic allocation, a selling price of \$1.30/L for bioethanol [55] and \$4.00/kg for SA [54] were considered.

Economic allocation is inherently sensitive to market price fluctuations; therefore, the resulting environmental impacts should be interpreted with caution. Nevertheless, economic allocation was selected as the primary interpretation method because it better represents the functional behaviour of the system, in which the co-products have different market roles and exhibit different levels of economic relevance. Mass allocation, while based on a physical property, does not capture these functional differences and may therefore misrepresent the relative environmental contributions of the co-products when their market values vary substantially.

2.3.2. Life cycle inventory

The life cycle inventory collects all input and output data of the proposed biorefineries. Energy requirements for equipment, as well as chemical and process water consumption, were obtained from the energy and mass balances of the process modelling using the SuperPro designer® software version 13.0 [23]. These simulation results are part of the life cycle inventory presented in Table 3. In addition, the transport of feedstock to the biorefinery was omitted, assuming that the facility is located adjacent to the processing factory of potatoes [59].

In the environmental analysis, background processes for chemicals, process water, electricity and heating transfer agents were sourced from the Ecoinvent® database version 3.8 [60]. The computational implementation of the life cycle inventory data was performed using SimaPro v10.1.0.6 software [61]. Environmental burdens associated with potatoes cultivation were based on the data provided by [62], with the addition of a specific peeling stage. For this, data from an industrial potato peeler were considered, including an average water consumption of 5.6 L per kg of potatoes and an estimated equipment lifespan of 28 years for the machine [63,64]. The transport of the potatoes from the agricultural to the processing stage is excluded from the system boundaries, as it falls outside the primary focus of this study.

2.3.3. Life cycle impact assessment (LCIA)

To transform the inventory data to environmental impact indicators, the ReCiPe 2016 (H) v1.07/World (2010) (H) method was applied [65]. This method was selected due to increased transparency, less uncertainty and increased comparability across studies. The characterization factors provided by ReCiPe were used to assess environmental impacts across nine different impact categories, which are widely accepted and commonly applied in other biorefinery studies [66], namely: Global Warming (GW), Stratospheric Ozone Depletion (SOD), Terrestrial Acidification (TA), Freshwater Eutrophication (FE), Marine Eutrophication (ME), Terrestrial Ecotoxicity (TET), Freshwater Ecotoxicity (FET), Marine Ecotoxicity (MET) and Fossil Resource Scarcity (FRS).

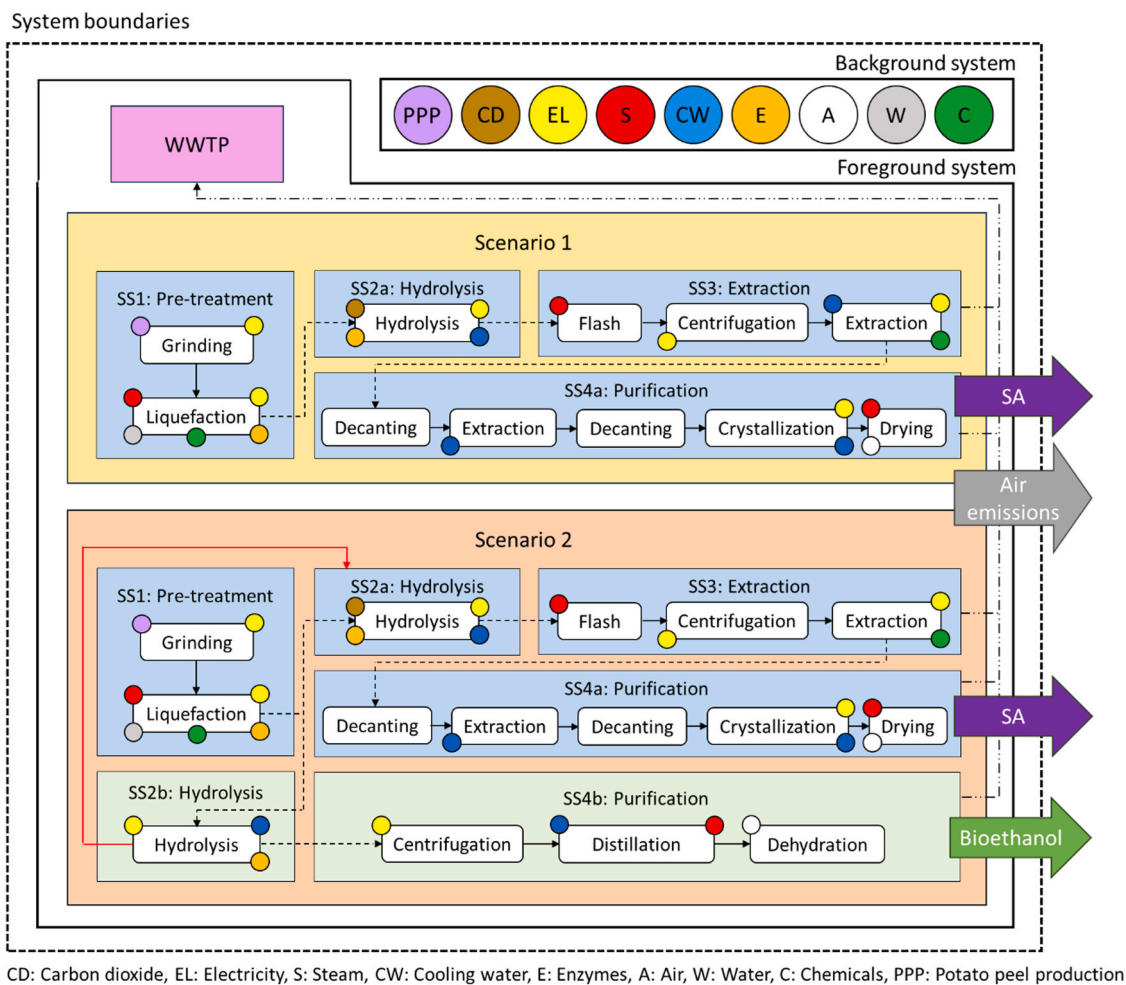


Fig. 2. System boundaries of the scenarios under study. The red arrow represents the recirculation of the CO₂ flow from SS2b to SS2a. SA: Succinic Acid.

Table 3
Life cycle inventory data per metric ton of treated potato peels.

Pretreatment (SS1)							
Inputs	A	B	Unit	Outputs	A	B	Unit
Potato peels	1		MT				
Citric acid	110.2		kg				
Enzymes	0.84		kg				
Water	8.72		MT				
Electricity	15.04		kWh				
Steam	1.15		MT				
Fermentation (SS2)							
Inputs	A	B	Unit	Outputs	A	B	Unit
Enzymes	17.2		kg	Emissions			
Carbon dioxide	218.8	44.3	kg	Carbon dioxide	78	46.8	kg
Electricity	367.2	343.9	kWh				
Cooling water	157.1	155.5	m ³				
Extraction (SS3)							
Inputs	A	B	Unit	Outputs	A	B	Unit
tri-n-octylamine	3.32	1.99	kg	Emissions			
1-octanol	36.8	22.9	kg	To WWTP	9.18	5.51	MT
Electricity	0.58	0.47	kWh				
Cooling water	8.58	5.75	m ³				
Steam	11.34	6.81	MT				
Purification (SS4)							
Inputs	A	B	Unit	Outputs	A	B	Unit
Air	1.88	1.13	MT	Products			
Water	10.75	6.45	kg	SA	385.1	230.6	kg
Trimethylamine	2.58	1.55	kg	Bioethanol	-	84.5	kg
Electricity	0.18	1.09	kWh	Emissions			
Cooling water	0.47	90.98	m ³	To air	84.35	58.7	kg
Steam	95.1	1522	kg	To WWTP	0.41	4.01	MT

Furthermore, the Cumulative Energy Demand (CED) method (Low Heating Values) v1.11 [67] was applied, as it is considered one of the most suitable LCIA methods for evaluating energy requirements [59].

3. Results and discussion

3.1. Tecno-economic results

3.1.1. Equipment, TCI and manufacturing costs

The total final cost associated with the purchase and installation of the equipment involved in Scenarios 1 and 2 amounts to \$MM115 and \$MM119, respectively. A detailed breakdown of the equipment costs for each scenario is provided in Table S5 of the Supplementary Material. Additionally, Fig. 3 illustrates the percentage contribution of each process section to the overall cost.

As shown in Fig. 3.A, in Scenario 1, the highest proportion of equipment costs is attributed to the fermentation and extraction sections. This trend is further supported by the individual equipment costs detailed in Table S5, which reveal that these sections include the reactors, identified as the most economically significant units within the fermentation-based SA production process, accounting for 62 % of the total equipment cost in Scenario 1 and 56 % in Scenario 2. These reactors are designed to operate at a maximum working volume of 45 m³ with mechanical agitation [45]. The substantial economic contribution of the fermenters observed in this study is consistent with findings reported in other works. For instance, Hafyan et al. [2] estimated that, in the design of a biorefinery for SA and bioethanol production from bread waste, 43–53 % of the total process cost was attributed to the reactors. Similarly, Thanahiranya et al. [33] reported an even higher contribution of fermentation tanks, ranging from 76 % to 90 % of the total equipment cost in the production of SA from glycerol, substantially exceeding the values obtained in the present study. In contrast, Fig. 3.B illustrates a clear decrease in the extraction-related costs in Scenario 2, accompanied by a notable increase in the fermentation and purification sections. This shift is mainly driven by the addition of a fermentative line for bioethanol production, which requires the installation of costly equipment such as an additional fermentation reactor (\$24,246,000) and a molecular sieve unit (\$2543,054). The substantial economic contribution associated with the purification stage is consistent with findings by Mupondwa et al. [68], who reported that, in bioethanol production from triticale grain and lignocellulosic straw, the largest portion of process costs was linked to purification, particularly distillation and product recovery.

Additionally, based on the equipment values, a Lang factor of 3.2, and a WC requirement of 5 %, the TCI for Scenarios 1 and 2 was estimated at \$MM385 and \$MM401, respectively. These values are comparable with those reported for other biorefineries of similar scale. For instance, Qian et al. [69], proposed a wood-based bioethanol biorefinery

with TCI values ranging from \$MM541.1 to \$MM627.6, depending on the selected processing route, slightly higher than the values obtained in this study. In contrast, for smaller-scale operations, lower TCI values have also been reported. Melitos et al. [1], for example, estimated a TCI of approximately \$MM110 for a biorefinery producing SA from commercial glycerol or glucose. Additionally, the results were expressed in terms of FCI per kilogram of product, with Scenario 1 yielding \$4.76/kg of SA and Scenario 2 reaching \$5.65/kg of total products. When compared with previously reported biorefinery studies for SA production, the study by Ladakis et al. [53], which employed the organic fraction of municipal solid waste as feedstock and assumed an annual production capacity of 60 kilometric tons, reported a value of \$6.45/kg SA. This result is higher than those obtained in both scenarios of the present work, indicating a comparatively lower investment cost for SA production in the current process. On the other hand, Kim et al. [70] reported a value of \$5.20/kg SA using sugar-rich wastewater as feedstock, which is much closer to the value obtained in both scenarios of this study.

The COM results are presented in Fig. 4 and further detailed in Table S6 of the supplementary material. The final estimated cost for Scenario 1 was \$MM150, while that for Scenario 2 was \$MM143.

As illustrated in Fig. 4, the cost contributions from raw materials and utilities represent the most significant portions of overall COM, compared to other contributions such as waste treatment or labor expenses. This is primarily due to the use of high-priced reagents such as TMA (\$10000/MT), TOA (\$1000/MT), and even octanol (\$210/MT), which are employed during the extraction stage. For this reason, solvent recirculation through a Reactive Extraction–Back Extraction Solvent Recycle Loop is of paramount importance from both an economic and an environmental perspective. Although alternative SA recovery technologies have been reported in the literature, they were not considered in the present study due to several inherent limitations. For example, electrodialysis based on ion-exchange membranes is constrained by the limited chemical compatibility of conventional membranes with strong organic solvents such as 1-octanol, which can lead to severe membrane degradation [71]. Indeed, numerous recent studies have focused on enhancing the chemical resistance of ion-exchange membranes to aggressive organic media [72]. Another commonly reported approach for SA recovery is precipitation, which involves the formation of succinate salts using an alkaline reagent such as calcium hydroxide. However, the study conducted by Alexandri et al. [73] demonstrated significantly lower recovery yields compared with those obtained using reactive extraction, consistent with the findings of the present work.

Utilities, on the other hand, also typically represent a major expense in industrial process design. For instance, in the study by Melitos et al. [1] on a SA biorefinery, utility costs alone accounted for 14 % of the total manufacturing costs. A similar trend is observed in the study by Kim et al. [70] about SA production from sugar-rich wastewater, where

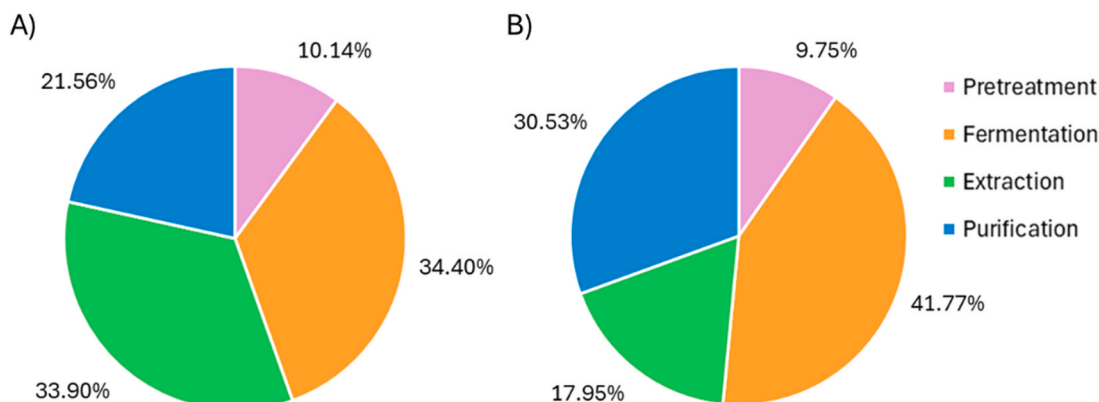


Fig. 3. Percentage of area cost in scenario 1 (A) and 2 (B).

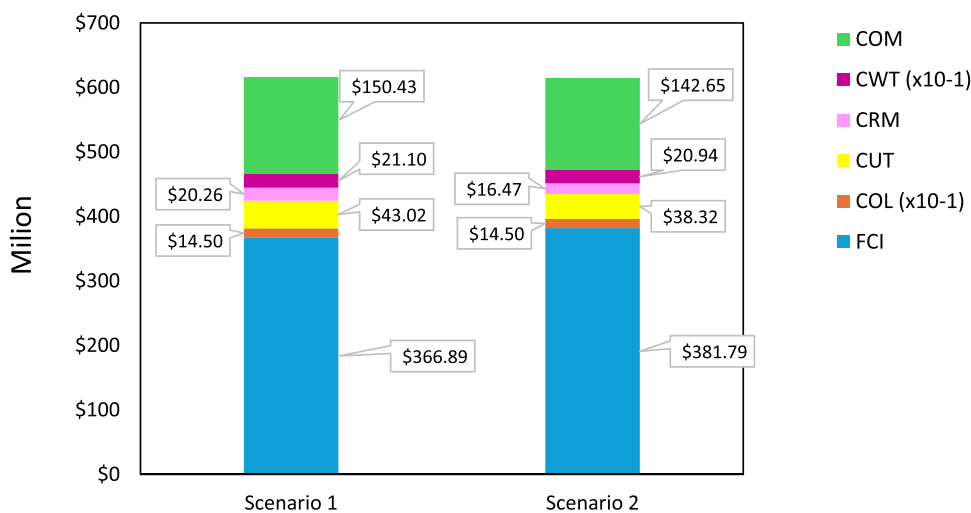


Fig. 4. Annual cost of manufacturing for each scenario.

raw material and utility expenditures are also identified as key contributors to the overall COM.

Finally, although the relative cost associated with the Cwt stream is low compared with other process contributors (\$2109,800 for Scenario 1 and \$2094,400 for Scenario 2, as reported in [Supplementary Table S6](#)). Within an integrated biorefinery framework and considering the composition of the initial solid, despite its high moisture content, this fraction could potentially be valorized, for instance, as animal feed for pigs, ruminants, or even aquaculture species [74–76], among other applications. Nevertheless, the detailed evaluation of such valorization routes falls beyond the scope of the present study, which was primarily focused on the core processing stages.

3.1.2. Economic indicators

During the TEA, the economic indicators MSP, DPP, IRR, and NPV were evaluated. In Scenario 1, the MSP of SA was estimated at \$3.77/kg. Although this value is slightly below the assumed market price of \$4.00/kg, which is a positive indicator, it remains very close to the selling price, potentially compromising the economic viability of the process. Nevertheless, SA is considered a product with strong market potential, having experienced a historical market growth rate of 12.1 % from 2019 to 2024. Projections indicate a compound annual growth rate (CAGR) of 12.3 % from 2025 to 2032 [54], suggesting that the current MSP should

not be a major concern. In fact, several scientific studies on bio-based SA production have considered significantly higher market prices. For instance, Tomczyk et al. [14] reported a price of \$9.00/kg, while Melitos et al. [1] considered values up to \$5.00/kg. The opposite situation was observed in Scenario 2, where an MSP of \$5.63/kg for SA and \$4.82/kg for bioethanol was observed. The fact that these values exceed current market prices confirms that this scenario is not economically feasible. In the case of SA, the MSP is not far from those reported by other authors, as previously discussed. However, for bioethanol, the selling price is significantly higher than the \$1.3/kg assumed in this study, which severely limits its economic viability. Regarding the comparison between the price obtained in our study and the values reported in various market analyses for petrochemically derived SA, the latter is considerably more favorable, ranging between \$1.4–1.8/kg in 2024 [77]. Although the petrochemical route is currently more cost-competitive, the targets established under the 2030 Agenda to reduce the reliance on fossil resources [78] support a gradual transition toward bio-based SA production, whose costs are expected to decrease as technological maturity and process integration.

Fig. 5 presents the DCF for each of the 30 years of the plant's operational lifetime. The calculated payback periods were 6.18 years for Scenario 1 and 9.20 years for Scenario 2. Furthermore, Scenario 1 consistently yields higher annual discounted cash flows over the entire

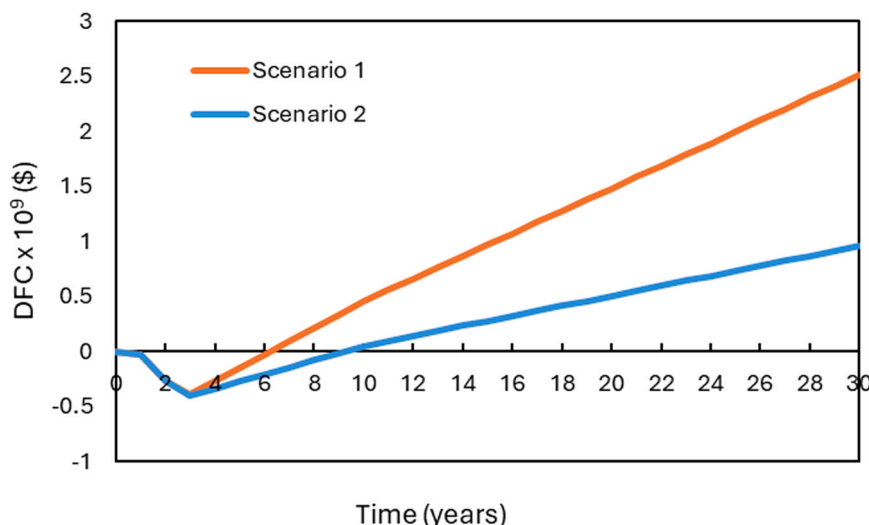


Fig. 5. DFC diagram of each scenario for a 30-year period.

operational lifetime of the plant. These findings suggest that Scenario 1, which involves the exclusive production of SA, enables a faster recovery of the initial capital investment.

The DPP obtained in the present work are slightly higher than those reported in other simulation-based studies of SA biorefineries. For instance, Melitos et al. [1] reported payback periods ranging from 4.38 to 8.46 years depending on the scenario considered. Nevertheless, the values obtained in the actual study remain competitive when compared to those reported by Kim et al. [70], who estimated payback periods between 8.56 and 10.49 years for SA production from sugar-rich wastewater. These results are consistent with IRR values obtained. For Scenario 1, an IRR of 11.21 % was calculated, above the 10 % threshold, indicating economic viability. However, as expected, the IRR for Scenario 2 was calculated at 3.94 %, which falls below the benchmark, suggesting that this scenario would not be economically feasible.

Finally, the NPV calculated for Scenario 1 was \$MM81, indicating that the proposed biorefinery scheme for SA production from potato peels would be economically viable. Scenario 2, which involves the coproduction of SA and bioethanol, resulted in a negative NPV of \$MM-339. Under the current assumptions regarding the discount rate and product selling prices, this configuration would not be considered economically feasible.

3.1.3. Sensitivity analysis

A sensitivity analysis was conducted to evaluate the influence of key economic variables on the NPV of the biorefinery for each of the proposed scenarios. The parameters considered included utility costs, raw material costs, FCI, the selling price of SA and bioethanol, all of which were previously analyzed in the preceding section. Each parameter was varied by $\pm 20\%$ from its base value to assess its impact on the economic performance of the system. The results of the sensitivity analysis for both scenarios are presented in Fig. 6.

The sensitivity analysis results presented in Fig. 6 indicate that, for both scenarios, the most influential parameter on the NPV is the selling price of SA. For instance, a 20 % increase in the SA market price (corresponding to \$4.80/kg) leads to a substantial rise in the NPV of Scenario 1 by \$MM278, significantly enhancing the economic viability of the biorefinery. In Scenario 2, although this price increase is not sufficient to render the process economically viable, given that the MSP for SA in this configuration is \$5.89/kg, it still results in a notable NPV improvement of \$MM167. As previously discussed, the positive market outlook for SA, with projected price growth, underscores the relevance of this finding. In contrast, the selling price of bioethanol exhibited low sensitivity in Scenario 2. A 20 % variation in its value led to only a 7.4 % change in the corresponding NPV, indicating a limited influence on the overall economic performance of the process.

Another economically significant factor highlighted in Fig. 6 is the FCI. In both scenarios, the calculated FCI was relatively high, and thus a $\pm 20\%$ variation had a substantial impact on the resulting NPV. Specifically, a 20 % reduction in FCI led to a remarkable 73 % increase in

NPV in Scenario 1, raising it from \$MM81.4 to \$MM140.9. In Scenario 2, the same reduction resulted in an 18 % improvement, increasing the negative NPV by \$MM62. Although the FCI is a critical factor, a 20 % increase in its value did not make Scenario 1 economically unfeasible, highlighting the robustness of this scenario from an economic standpoint. Finally, raw material and utility costs were found to be the least influential parameters in both scenarios.

3.2. LCA results

3.2.1. Environmental assessment

Table 4 presents the results of the environmental analysis carried out using a cradle-to-gate approach for each of the proposed scenarios and for both FUs selected. As shown, on a per-batch basis of feedstock treated, Scenario 2 leads to a reduction in impacts across all assessed categories compared to Scenario 1. This indicates that the addition of a section for skins aimed at bioethanol production, along with the recirculation of the CO₂ generated to the SA line, results in an environmental improvement. The reductions range from over 16 % in categories such as GW, FRS and CED, to less than 5 % in SOD and ME. This improvement is primarily due to the absence of SS3 in the bioethanol production line, which includes an intensive extraction stage and represents the main source of environmental impacts in Scenario 1.

Focusing on FU2, the environmental burdens allocated to SA and bioethanol depend strongly on the allocation method. Under mass allocation, SA receives 73.24 % of the total burdens, and although the production of SA decreases by approximately 155 kg when the bioethanol section is added, only about 55 % of that mass is recovered as bioethanol. As a result, the environmental performance of Scenario 2 slightly worsens under FU2, yet it still performs slightly better than Scenario 1 in terms of TET and CED, while differences in GW, FE, and FRS remain below 5 %. In contrast, economic allocation assigns 86.75 % of the total burdens to SA due to its higher market value, increasing the environmental impacts attributed to this product and highlighting the sensitivity of the results to the choice of allocation method. The impacts associated with bioethanol are presented in Table S8 of the Supplementary Material.

Focusing on Scenario 1, SS3 emerges as the main contributing subsystem in CED, while SS1 plays the most significant role in the rest of the analysed impact categories. The impact of SS1 is primarily driven by the production of citric acid used for the buffer solution, as it is a highly intensive process, and by feedstock production, especially in SOD and ME, since the agricultural phase of potato cultivation and the use of nitrogen-based fertilizers contribute significantly to these impact categories. GW impacts arise mainly from the production of carbon dioxide required for fermentation and from enzyme production, while the latter process is the main contributor of this subsystem in the other impact categories. Although used in small amounts, enzymes are produced through highly energy-intensive processes, leading to substantial environmental impacts. Finally, impacts related to SS3 subsystem are

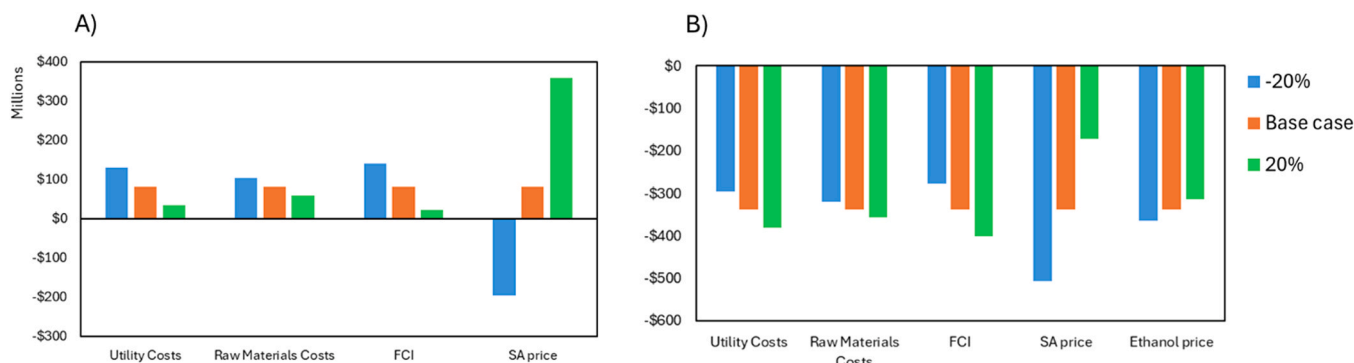


Fig. 6. Sensitivity analysis of each scenario: Scenario 1 (A) and Scenario 2 (B).

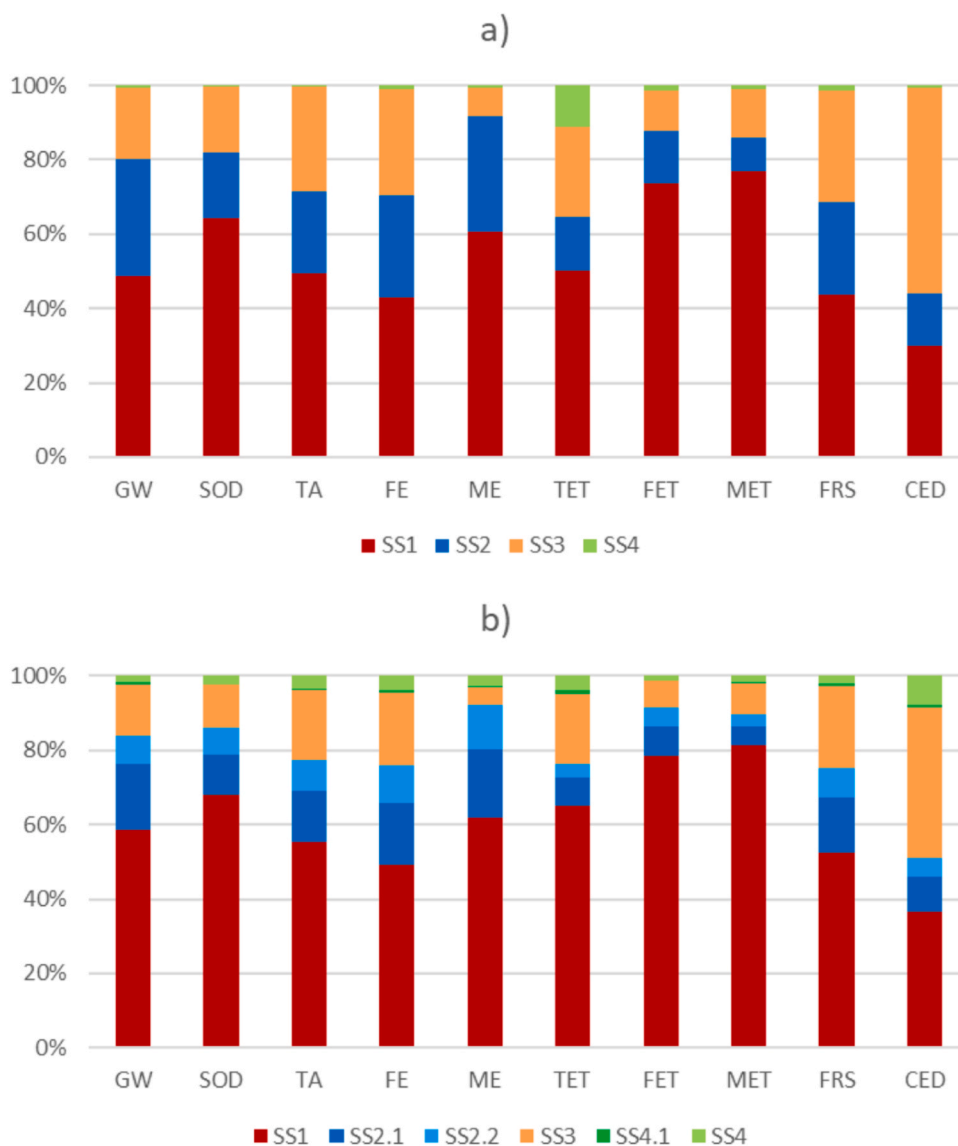


Fig. 7. Distribution of contributions to the environmental profile per subsystems involved within the SA production plant per each one of the scenarios (a: Scenario 1, b: Scenario 2). GW: Global Warming, SOD: Stratospheric Ozone Depletion, TA: Terrestrial Acidification, FE: Freshwater Eutrophication, ME: Marine Eutrophication, TET: Terrestrial Ecotoxicity, FET: Freshwater Ecotoxicity, MET: Marine Ecotoxicity, FRS: Fossil Resource Scarcity, CED: Cumulative Energy Demand.

primarily driven by heating requirements. This is due to its high fossil fuel requirements—particularly oil and natural gas—to satisfy the heat demand of the extraction processes. Therefore, this impact can be attributed not to the direct requirements of the design, but rather to the impacts associated with background processes.

Turning to Scenario 2, the distribution is quite similar to the obtained in Scenario 1, with the SS1 being the main hotspot in all impact categories except CED, where the heating requirements of SS3 has a bigger impact. As in the previous scenario, in the subsystem SS1, the citric acid production process is the primary source of impacts. However, SS2 shows a notable increase in the contribution of enzyme production, which is expected, as less CO₂ is consumed due to the recirculation of part of the CO₂ generated during fermentation. Regarding SS3, as in the previous subsystem, impacts are largely associated with energy requirements during processing. Finally, the contributions of SS4 are practically negligible; however, in Scenario 2 this subsystem becomes more relevant, as bioethanol distillation processes require a high amount of steam.

Analyzing the impact categories individually for both scenarios, emissions related to GW are predominantly caused by carbon dioxide,

followed by methane and nitrous oxide. These emissions are primarily associated with the production of citric acid required in SS1, followed by enzyme and carbon dioxide production in SS2. In SOD, the principal contributor is nitrous oxide (N₂O), originating from the transformation of nitrogen present in fertilizers applied during potato cultivation. For TA category, emissions are mainly associated with ammonia, sulfur dioxide (SO₂) and nitrogen oxides (NO_x), while for FE, the main contributors are phosphate, phosphorus, BOD₅ and COD emissions to water. In both cases, these impacts originate primarily from the most intensive background production processes, citric acid and enzyme production, as well as steam production based on the average European energy mix with wood-chip cogeneration. On the other hand, in the ME category, the profile changes completely, with the major contributors being nitrate (NO₃⁻) emissions resulting from the leaching of nitrogen-based fertilizers into water bodies. These occur both during the cultivation of the feedstock and the production of enzymes. The latter involves the cultivation of potatoes for starch extraction, which serves as a key input in the process. For TET, FET and MET, the impacts are again mainly driven by citric acid production, followed by steam production for SS3 and enzyme production for SS2. In FET and MET, the substances that

Table 4

Characterization results per batch of potato peels processed and kg of SA per each one of the scenarios proposed.

Impact category	FU1 (1 MT of potato peels)			FU2 (1 kg of SA)			
	Sc. 1	Sc. 2	Unit	Sc. 1	Sc. 2*	Sc.2**	Unit
GW	1642	1369	kg CO ₂ eq	4.26	4.34	5.14	kg CO ₂ eq
SOD	5.00	4.74	g CFC ₁₁ eq	13.01	15.01	17.78	mg CFC ₁₁ eq
TA	10.55	9.47	kg SO ₂ eq	27.40	30.03	35.57	g SO ₂ eq
FE	0.79	0.69	kg P eq	2.05	2.18	2.58	g P eq
ME	0.91	0.89	kg N eq	2.36	2.83	3.35	g N eq
TET	11.66	8.98	kg 1–4, DCB eq	30.29	28.46	33.71	kg 1–4, DCB eq
FET	69.61	65.52	kg 1–4, DCB eq	180.8	207.7	246	kg 1–4, DCB eq
MET	91.81	86.59	kg 1–4, DCB eq	238.4	274.5	325.1	kg 1–4, DCB eq
FRS	389.9	322.6	kg oil eq	1.01	1.02	1.21	kg oil eq
CED	62.85	51.2	GJ	163	162.8	192.2	MJ

GW: Global Warming, SOD: Stratospheric Ozone Depletion, TA: Terrestrial Acidification, FE: Freshwater Eutrophication, ME: Marine Eutrophication, TET: Terrestrial Ecotoxicity, FET: Freshwater Ecotoxicity, MET: Marine Ecotoxicity, FRS: Fossil Resource Scarcity, CED: Cumulative Energy Demand.

* Mass allocation.

** Economic Allocation.

contribute most significantly are copper and zinc, while in TET, emissions of the amine used, and acetic acid also account for more than 5 % of the total impacts. On the other hand, fossil fuels use for background processes of the biorefinery, particularly coal, natural gas and oil consumed during steam, octanol and enzymes generation, represent the largest contribution to FRS. Finally, in the CED indicator, both scenarios show that the main energy demand arises in SS3, especially due to the flash distillation used to reduce water content in the process.

3.2.2. Comparison with literature findings

To date, no studies examining the environmental assessment of a biorefinery focused on the valorization of potato peels have been found in the literature. Therefore, the discussion of environmental results was centered on FU2, comparing other biorefineries focused on SA production. In those studies, that produce only SA or co-products other than bioethanol, the comparison was made using the results from scenario 1, given that this scenario represents a single-product biorefinery configuration.

González-García et al. [79] investigated the valorization of apple pomace for SA production, although the feedstock production was not included within the system boundaries. The reported results, excluding enzyme production, are worse than those of the present study across all the analyzed categories. Specifically, the present study shows reductions of 20 % for GW, 4 % in SOD, 96 % in TA and 28 % in CED. This difference is mainly due to the high consumption of solvents reported in the study of González-García et al. [79] during the purification stage and the environmental burdens of their production. In this study, the main environmental impacts are related to citric acid, enzymes and steam production, with lower impacts related to chemicals due to their reutilization. However, the SA yield reported by González-García et al. [79] is higher, which further highlights the favorable environmental profile of the biorefinery design proposed in this manuscript. This is because, when using a product-based functional unit, a higher yield contributes to better environmental performance.

Baioni e Silva et al. [80] also proposed the valorization of apple pomace, but within the Brazilian context, achieving SA with a purity of 99.5 %. The reported emissions are significantly lower in all impact categories than those in the present study, ranging from an 42 %

reduction in FRS and 69 % in GW to 99 % in SOD. These differences can be attributed to the substantially lower consumption of heating agents in their proposed design. Moreover, the assumption of zero environmental burden for raw material production in that study helps to explain these discrepancies, as categories such as SOD, where raw material production has a greater influence, show the greatest differences between studies. Additionally, the SA to feedstock yield in their system was 53 %, compared to 39 % in the present study, due to the different biomass composition, which further contributes to the mitigation of impacts per unit of product.

On the other hand, Chryssikou et al. [81] proposed a biorefinery concept aimed at producing bioethanol and SA from *Phalaris aquatica* L. biomass. Their study reported only greenhouse gas (GHG) emissions, applying a mass allocation of 90.91 % to bioethanol and 9.09 % to SA. For their baseline scenario—which included fertilization and feedstock transport—the reported emissions were 1.02 kg CO₂ eq per kg of bioethanol and 193.2 g CO₂ eq per kg of SA, lower impacts than those obtained in the present study. One plausible explanation for this discrepancy is the absence of enzyme use in their SA production line, since enzyme production is a highly polluting process. In fact, raw material production accounts for 57 % of the total environmental impacts, clearly highlighting the relatively low burden associated with the subsequent processing stages.

Shaji et al. [82] and Gadkari et al. [83] explored the valorization of sugarcane bagasse and bread waste, respectively. These authors reported similar GW impacts per kilogram of SA (1.3–1.4 kg CO₂ eq). In Shaji et al. [82], the cultivation phase was included within the system boundaries but accounted for only 10 % of the total impacts, while the fermentation stage represented the main hotspot, contributing 60 % of the environmental burdens. Additionally, this study implemented the reuse of solid residues from the process to generate steam, reducing the impacts related to steam generation in later stages. In contrast, the present study cannot adopt this strategy due to the absence of solid residues. In Gadkari et al. [83], steam generation was also identified as the primary hotspot, responsible for 75 % of total impacts, underscoring the significance of steam demand highlighted in this study. However, their system boundaries excluded feedstock production, which may partially explain the more favorable environmental profile observed. Moreover, the enzyme usage in their study was six times lower, a critical factor given the high environmental impacts associated with enzyme production.

Finally, Kosamia et al. [84] analyzed the environmental performance of corn stover valorization, applying a system expansion approach in which all impacts were ultimately assigned to SA, the main product of the biorefinery. Among the three scenarios they proposed, two did not need external supply of steam, as the plant's needs were met by cogeneration, resulting in significantly lower impacts across all categories. The scenario with higher steam demand exhibited lower impacts in GW, TA, FE, TET and FRS; similar impacts in SOD; and higher impacts in ME, FET and MET compared to the results of the present study. This indicates that, for a biorefinery with comparable assumptions—including raw material production, pretreatment, enzyme use, and similar demand and type of chemicals—the results presented here are consistent.

To further support the environmental results, the impacts obtained in this study were compared with those reported for the production of the same products via conventional fossil-based routes (Table 5). For SA produced from naphtha oxidation (data from the Ecoinvent database), impacts in most analyzed categories are higher due to significant contributions from enzyme and citric acid production, whereas impacts in FRS are lower because of the reduced reliance on fossil feedstocks, as wood-chip cogeneration is applied for steam production. Regarding bioethanol, in categories directly related to fossil fuel consumption, such as GW and FRS, the proposed bioethanol performs better than coal-derived bioethanol. Nevertheless, in ME, which is highly influenced by NO_x emissions from feedstock processing, impacts are higher. For other

Table 5
Impacts of SA and bioethanol obtained via conventional fossil-based routes.

Impact category	SA [60]	Bioethanol [85]	Unit
GW	3.49	8.37	kg CO ₂ eq
SOD	7.9	1.16	mg CFC ₁₁ eq
TA	5.13	13	g SO ₂ eq
FE	0.65	1.83	g P eq
ME	48.48	116	mg N eq
TET	2.64	3.22	kg 1,4-DCB
FET	19.53	138	g 1,4-DCB
MET	27.64	191	g 1,4-DCB
FRS	1.4	1.65	kg oil eq
CED	67.93	-	MJ

categories, such as FE and TET, where enzyme and citric acid production make significant contributions, the fossil-based route again shows better results.

While further energy optimization of the process is recommended, the bio-based options demonstrate favorable environmental performance in categories most closely related to fossil resource consumption, such as GW and FRS, highlighting the potential benefits of the proposed biorefinery systems. Moreover, replacing the chemicals currently used in the refinery with less environmentally impactful alternatives should be investigated in order to further improve the overall environmental profile.

Overall, the environmental results obtained in the present study are consistent and fall within the range of values reported in the literature. However, future studies could focus on achieving higher SA extraction yields, as reported in some of the previously cited studies, which would help improve the environmental profile per unit of product obtained.

3.3. Comparative novelty against state-of-the-art TEA/LCA analyses

In the present study, TEA and an LCA were simultaneously conducted for two biorefinery configurations aimed at the valorization of potato peels. A comparison of the results from the two proposed scenarios reveals that while Scenario 1 demonstrates substantially better economic outcomes than Scenario 2, it is Scenario 2 that provides superior environmental performance. Considering the need to reduce environmental impacts within the context of circular bioeconomy in the coming years [86], various efforts should be directed towards increasing the economic value of Scenario 2, as it is more environmentally viable. Potential economic improvements could include enzymes reuse to reduce costs, the valorization of solid waste currently sent to the WWTP, and the commercialization of by-products derived from SA fermentation, such as acetic acid and formic acid.

To the best of the authors' knowledge, no previous work has employed PP as a feedstock within an integrated biorefinery framework that simultaneously considers TEA and LCA. Additionally, no research has yet explored the internal reutilization of CO₂ for the co-production of a high-value chemicals and a biofuels with PP. The novelty of using potato peels as a feedstock for biorefinery processes is thus a significant contribution to the field.

A study examining the co-production of bioethanol and SA with TEA and LCA has been identified, although using bread waste as the substrate [2]. Unlike the present work, that study does not include a reference scenario based on the single production of SA, which is considered the main product in this study. Additionally, it relies on substantially lower feedstock input flows, limiting the industrial applicability of its findings.

Other integrated biorefinery concepts for SA production incorporating TEA and LCA have been reported, but they rely on pure substrates such as glycerol or glucose [1]. As refined feedstocks, these materials do not require the physical-chemical pretreatment stages associated with real biomass, which artificially lowers both process costs and environmental a fair comparison with agro-industrial waste-based systems. Additionally, alternative configurations using corn stover have been

explored, typically limited to TEA [14] or combined TEA-LCA approaches targeting SA and furfural or gypsum [84]. However, these processes do not include the simultaneous production of bioethanol and therefore fail to address the challenge of securing a sustainable CO₂ supply required for SA biosynthesis. Overall, the absence of comparable studies underscores the originality and industrial relevance of the biorefinery concept proposed in this work.

4. Conclusions

This work provides a comprehensive evaluation of two biorefinery schemes for the valorization of PP, through the production of SA and bioethanol. By integrating process simulation (SuperPro Designer®), TEA, and LCA, the study offers a multidimensional understanding of the trade-offs between economic feasibility and environmental performance in emerging biorefinery systems.

The exclusive SA production scenario (Scenario 1) was found to be economically viable under current assumptions, yielding a competitive MSP of \$3.77/kg SA, a positive NPV of \$81 million, and an IRR of 11.21 % (above the threshold of 10 %). In contrast, the integrated scenario (Scenario 2), which combines SA and bioethanol production by the valorization of CO₂ emissions from bioethanol fermentation, failed to reach economic feasibility due to its higher capital costs and low ethanol market value, despite demonstrating significant environmental benefits.

From an environmental standpoint, Scenario 2 outperformed Scenario 1 across most impact categories, particularly GWP, TET, FRS, and CED. This improvement is attributed to the reduction in high-impact stages such as SA extraction and the effective reuse of internally generated CO₂. Nevertheless, when results are normalized per kilogram of product, the environmental advantages of Scenario 2 diminish due to reduced SA yields and only partial compensation by bioethanol production.

The sensitivity analysis highlighted the strong influence of SA market price and FCI on the overall economic performance. A 20 % increase in SA price drastically improved NPV in both scenarios, whereas FCI reductions were critical to Scenario 2's economic recovery. These findings underscore the importance of market dynamics, capital optimization, and cost-effective process integration in biorefinery design.

However, although SuperPro Designer is widely used in biorefinery process development, its role is most appropriate for early-stage design and preliminary techno-economic evaluations, given certain limitations and the associated uncertainties in its predictions. The software relies on moderately rigorous thermodynamic data, provides simplified kinetic models, and offers limited optimization capabilities, which may constrain its accuracy for highly complex or strongly integrated systems. Nevertheless, it remains a useful tool for generating consistent process flowsheets and supporting initial decision-making, provided that its outputs are complemented with more detailed modelling or experimental validation.

Overall, this study emphasizes the feasibility of producing bio-based SA from PP and reveals the potential of coupled biorefinery schemes to enhance carbon efficiency and circularity. However, further improvements in downstream processing, enzyme reuse strategies, and carbon valorization pathways are needed to close the economic gap in integrated scenarios. Future research should also address the scalability of these systems, their integration into existing agro-industrial infrastructure, and their performance under varying regulatory and market conditions.

CRedit authorship contribution statement

Beatriz Rodríguez-Martínez: Writing – review & editing, Writing – original draft, Visualization, Validation, Methodology, Investigation, Formal analysis, Data curation, Conceptualization. **Sara González-García:** Writing – review & editing, Visualization, Validation, Supervision, Project administration, Methodology, Conceptualization. **Adrián**

Agraso-Otero: Writing – review & editing, Writing – original draft, Validation, Methodology, Investigation, Data curation. **Ricardo Rebolledo-Leiva:** Writing – review & editing, Visualization, Validation, Supervision, Methodology, Conceptualization. **Remedios Yáñez:** Visualization, Supervision, Project administration, Conceptualization. **Gullón Beatriz:** Writing – review & editing, Writing – original draft, Visualization, Validation, Supervision, Project administration, Methodology, Investigation, Formal analysis, Conceptualization.

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.

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Appendix A. Supporting information

Supplementary data associated with this article can be found in the online version at [doi:10.1016/j.jece.2026.121288](https://doi.org/10.1016/j.jece.2026.121288).

Data Availability

Data will be made available on request.

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