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An environmental and economic sustainability assessment of novel rhamnogalacturonan-I pectin production from agricultural residues

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ABSTRACT

This study presents for the first time an environmental and economic assessment of large-scale process configurations employing novel technologies for rhamnogalacturonan-I pectin production from sugar beet pulp (SBP) and discarded red beetroot (DRB). Three alternatives to conventional acidic extraction (CE) were compared. Alternatives 1 and 2 focused on pectin derived from SBP, combining acid-free microwave assisted extraction (MAE) or hydrothermal extraction (HE), respectively, whilst Alternative 3 considered pectin extraction from DRB using MAE. All alternatives shared a membrane filtration and freeze drying downstream processing. Prospective life cycle assessment (LCA) and techno-economic analysis (TEA) were performed using scaled-up data from own experimental extraction and purification results using process simulation. The assessment was from cradle-to-gate for the "production of 1 kg of pectin", considering eight environmental impacts and six economic indicators including climate change (CC) and minimum selling price (MSP). The CC impacts were reduced by 89.3 – 94.4 % in the alternatives compared to the Reference case because of introducing solvent-free downstream processing. The MSP ranged from 16.2 to 56.0 €/kg, with the alternative processes showing better economic feasibility, which can be further increased through the co-production of DRB juice based on the findings from a sensitivity analysis. The processes were ranked based on overall eco-efficiency scores as follows: A3 (0.91) > A1 (0.73) > A2 (0.62) > > Reference case (0.05), highlighting the environmental and economic advantage of MAE and valorization of DRB, emphasizing the promise of emerging technologies and alternative feedstocks for developing more sustainable and competitive pectin production processes.

1. Introduction

The global agricultural sector and food supply chain generate large quantities of byproducts and discarded produce, causing significant environmental issues and a resource management challenge. For instance, in Europe, sugar beet is the second largest agricultural crop (157 million tons per year), which generates 13 million tons per year of sugar beet pulp (SBP) resulting from sugar extraction [1,2]. According to reports from the United Nations Development Programme, around 1.3 billion tons of food are wasted annually, of which 42 % originates from discarded fruits and vegetables [3]. Discarded products such as carrots [4], red beetroots [3,5] and oranges [6], and waste streams like SBP, are rich in cellulose, hemicellulose and bioactive compounds such as pectin [1,3,5].

The efficient valorization of agricultural byproducts and discarded products not only minimizes waste and reduces environmental impacts but creates opportunities for resource circularity and generates economic value through the production of high-value added products. Among the bioactive compounds, rhamnogalacturonan-I (RG-I) pectin and its derived pectooligosaccharides (POS), stands out due to its wide range of applications in the food, pharmaceutical, and cosmetic industries, where it is valued for its gelling, stabilizing, and thickening

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properties [7]. Moreover, POS, which are rich in neutral sugars, also exhibits prebiotic properties, making it a promising ingredient in functional foods [8–10]. The strong market growth of pectin (6.9 % compound annual growth rate (CAGR) by 2028) and prebiotics (11.4 % CAGR by 2029) highlights the promising potential of sustainable production pathways [11,12].

However, conventional pectin production processes, which rely on acidic extraction and alcohol precipitation, present several challenges, including high energy consumption, significant solvent use, and important environmental issues [13]. These limitations have led to the development of emerging technologies such as hydrothermal extraction (HE) and microwave-assisted extraction (MAE), which offer several advantages. For example, HE employs water under high temperature and pressure, reducing the need for chemical solvents, while MAE enhances extraction efficiency and reduces processing time [14,15]. Several studies have reported higher pectin yields using both HE and MAE [3,4,14], as well as lower environmental impacts [16,17] compared to traditional methods. However, commercial pectin production pathways still rely on traditional downstream purification methods that use alcohol for precipitation, thus a significant bottleneck in terms of environmental impacts and operating costs [13]. Membrane filtration has been proposed as an alternative solvent-free purification solution due to several advantages including cost-effectiveness, scalability, and low energy consumption [4,18].

Whilst the literature has demonstrated the technical viability of emerging extraction and purification technologies for producing pectin from agricultural waste and discarded products, there is no clear direction on the best process configurations, based on economic feasibility and environmental performance metrics, due to a lack of prospective techno-economic analysis (TEA) and environmental sustainability studies evaluating promising combination of technologies and feedstocks. Only a handful of environmental sustainability studies have been reported in the literature, covering both conventional [6,13] and emerging technologies [16,19,20] to produce pectin derived primarily from citrus (orange peel) and food waste (SBP), and a single TEA study [6,13] comparing the selling price of pectin produced via different modifications of the conventional route, including the use of organic acids, reduced ethanol consumption, bioethanol and renewable electricity. Although these studies offer insights into the hotspots of the systems assessed, i.e. the use of alcohol-based solvents and energy consumption in drying steps, inconsistencies across system boundaries, e.g. exclusion of downstream processing stages, and the limiting range of options compared are insufficient to steer relevant stakeholders towards more sustainable and innovative industry.

The authors have done extensive experimental research on pectin extraction from agricultural waste (other than citrus) using MAE coupled with membrane purification and freeze-drying [3,18,21,22]. Building on this foundation, the aim of the present study is two-fold. First, to generate experimental results using HE and DRB as additional extraction technology and feedstock for pectin production, respectively, thus extending the scope of data previously established. Secondly, to conduct, for the first time, a comparative and prospective economic and environmental sustainability assessment of different large-scale process configurations for RG-I pectin production from SBP and DRB (a currently underutilized waste stream), to identify the most promising pathways in terms of value added and environmental impacts. The analysis employs TEA, LCA and eco-efficiency, well-established sustainability assessment methods, using optimized experimental data scaled up with process simulation tools and advanced chemical engineering calculations, thus providing a robust basis for prospective evaluations and comparison with the conventional route. This integrated assessment provides stakeholders with quantitative evidence at the early stages of process development, supporting sustainable innovation and decision-making towards economically viable pectin production, while diversifying the supply chain beyond traditional sources.

2. Materials and methods

The following RG-I pectin production processes are being proposed for evaluating and comparing their environmental and economic sustainability at a larger scale. This is to determine their advantages and potential limitations with respect to the valorization techniques currently in place. In addition, two different sources of biomass are also being examined: SBP and discarded red beetroot (DRB), to evaluate the potential advantages of using waste as raw material in a multiproduct biorefinery approach.

- Reference case (SBP CE): pectin derived from SBP using conventional extraction and downstream processing (see Fig. 1a);
- Alternative 1 (SBP MAE): pectin derived from SBP using acid-free microwave assisted extraction and novel downstream processing (see Fig. 1b);
- Alternative 2 (SBP HE): pectin derived from SBP using hydrothermal extraction and novel downstream processing (see Fig. 1c); and
- Alternative 3 (DRB MAE): pectin derived from DRB using acid-free microwave assisted extraction and novel downstream processing (see Fig. 1d).

The MAE technique was chosen for the feedstock comparison because of the higher pectin yields reported and lower operation time [23]. The process conditions for simulating each option, and therefore, to obtain the material and energy balances to conduct the prospective LCA and TEA at a larger scale, were taken from different sources and will be discussed in detail in Section 2.2. For the Reference case and Alternatives 1 and 3, these were sourced from literature including the authors' previous work on MAE and membrane purification processes [3,18,22,24]. For pectin extraction in Alternative 2, on the other hand, this is based on new work carried out at the Technical University of Denmark and presented in Section 2.1 as part of this study. The methodologies employed to carry out the process simulations, LCA and TEA are presented in Sections 2.3 – 2.5.

2.1. Optimization of RG-I pectin recovery from SBP by hydrothermal extraction

2.1.1. Raw material preparation

SBP was supplied by AB Azucarera Iberia. SBP was prepared as in our previous study [14]. It was washed, dried at 60 °C, ground using a hammer mill (Polymix, PX-MFC 90 D, Kinematica AG, Switzerland) and passed through a sieve (particle size < 1 mm) before use.

2.1.2. Hydrothermal extraction of RG-I pectin from SBP

HE was conducted in a 600 mL stainless-steel vessel Parr Series 4760 (Parr Instruments Company, IL, USA) immersed in an oil bath with silicone oil using an Eco Gold E40G (Lauda, Germany).

SBP was mixed with water in a solid-to-liquid ratio of 10 % (w/v) (10 g of dry SBP) using the equipment vessels. These were introduced in the oil bath once the set temperature was reached (see Table 1). The extraction time measurements began once the temperature stabilized again. At the end of the extraction, the vessels were cooled down in an ice bath for 30 min. The solid and liquid fractions were separated using a vacuum filter. The solid was dried at 60 °C and weighed to calculate the solid recovery on a dry basis (g spent solid/g SBP). The liquid was stored at 4 °C for further analysis to quantify the monomeric and oligomeric sugar and degradation compounds contents.

2.1.3. Experimental design for RG-I pectin extraction

In order to optimize the conditions of RG-I pectin recovery by HE, a response surface methodology – face-centered design (RSM-FCD) was planned (Table 1). The parameters selected were temperature (X_1, 135 – 175 °C) and extraction time (X_2, 30 – 90 min). The response variable was the concentration of POS, calculated as the sum of oligogalacturonides

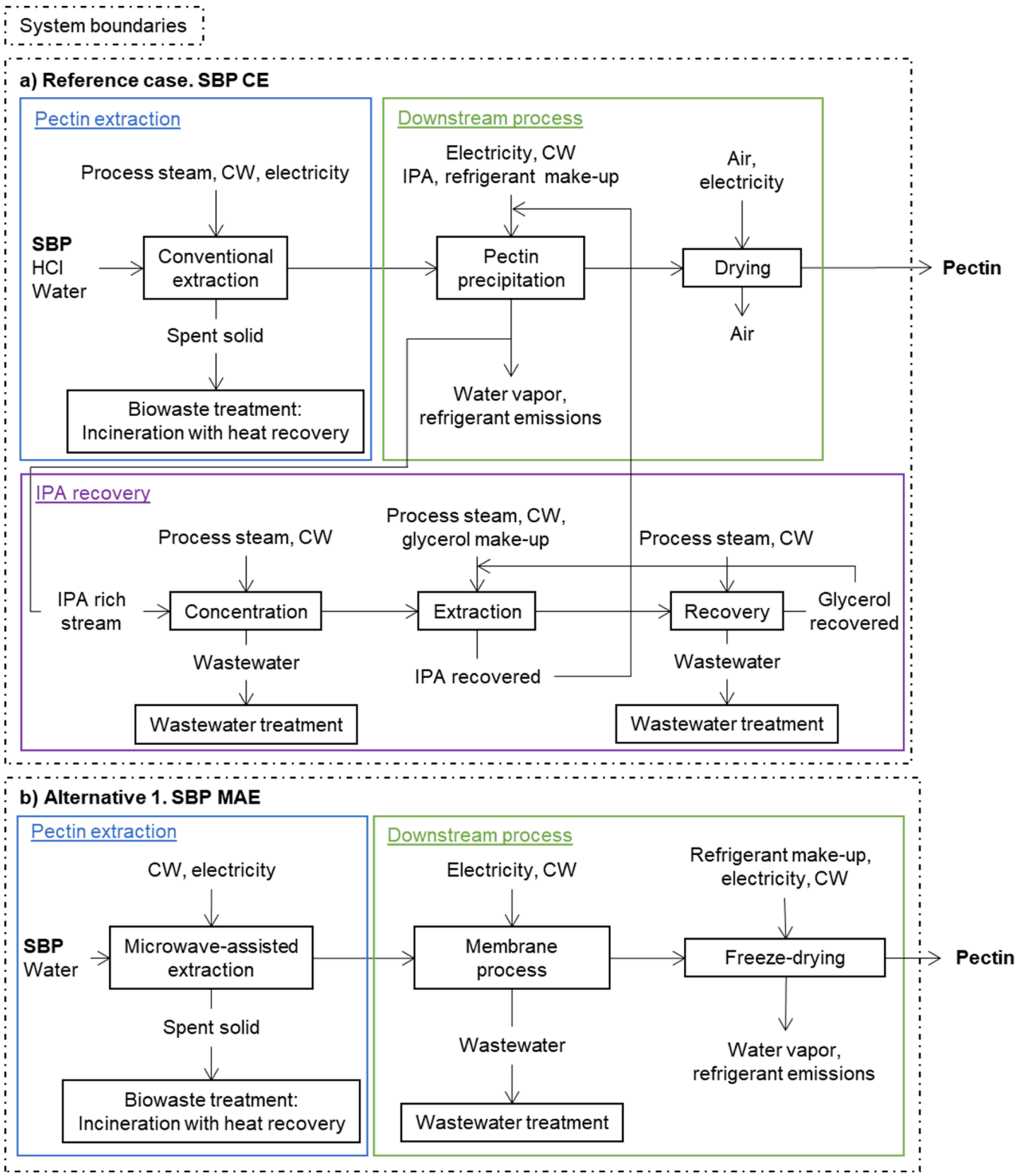

Fig. 1. LCA system boundaries considered for the production of RG-I pectin in the reference case (RC. SBP CE) and the three alternative scenarios evaluated (A1. SBP MAE, A2. SBP HE and A3. DRB MAE). SBP: sugar beet pulp; CW: cooling water; IPA: isopropanol; DRB: discarded red beetroot.

(OGalA), galactooligosaccharides (GalOS), rhamnooligosaccharides (RhaOS), and arabinooligosaccharides (AraOS). Thus, 11 experiments (Table 1) were carried out in duplicate, including three replicates at the center of the design to estimate the pure error sum of squares. The experimental results were analyzed using Statgraphics Centurion XVIII software. The severity factor (SF) was calculated according to MacAskill et al. [25] (Eq. 1), where t is the extraction time (min) and T is the temperature (°C).

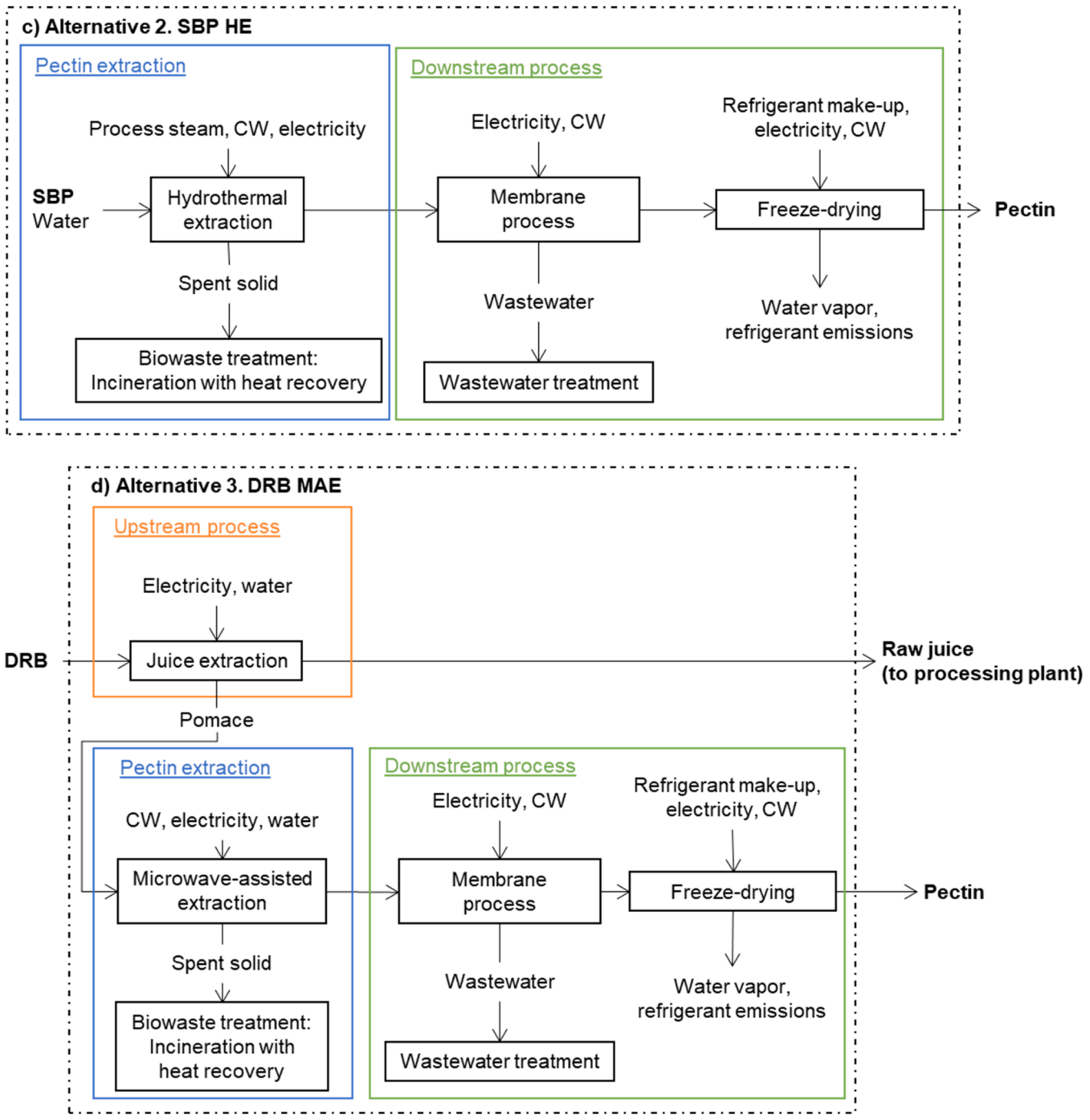

Fig. 1. (*continued*).

$$Severity\ Factor(SF) = Log\cdot\left[t\cdot\exp\left(\frac{T-100}{14.75}\right)\right] \quad (1)$$

Finally, a confirmatory run was done in triplicate under the optimal conditions to verify the results obtained by the statistical software.

2.1.4. Statistical analysis

A statistical analysis and RSM-FCD were carried out using the Statgraphics Centurion XVIII software. An ANOVA test was performed to conclude the significant difference or relationship at a confidence level of 95 % ($p < 0.05$).

2.1.5. Analytical methods

The concentration of soluble sugars (glucose, galactose, rhamnose, and arabinose), galacturonic acid (GalA), and degradation compounds (acetic and formic acids, HMF, and furfural) in the extracts after HE were determined by High-Performance Liquid Chromatography (HPLC), using a Dionex Ultimate 3000 high-performance liquid chromatography UHPLC+ Focused system (Dionex Softron GmbH, Germany) with a Bio-Rad Aminex column HPX-87H at 60 °C, a Shodez RI-101 refractive index detector, and 5 mM H_2SO_4 as mobile phase at a flow rate of 0.6 mL/min.

2.2. RG-I pectin production processes

The production of RG-I pectin from SBP and DRB was compared

Table 1
Response surface method face centered design (RSM-FCD): experimental conditions, severity factor (SF), pH of extracts and solid recovery in each run and under optimal conditions (Opt, 153 °C, 59 min).

RUN	Independent variables				SF	Extract pH	Solid recovery (%)
	x_1	X_1 (Temperature, °C)	x_2	X_2 (Time, min)			
1	-1	135	-1	30	2.5	4.4 ± 0.1	93.7 ± 2.4
2	-1	135	1	90	3.0	4.0 ± 0.0	83.1 ± 3.7
3	1	175	-1	30	3.7	3.8 ± 0.0	71.6 ± 1.8
4	1	175	1	90	4.2	3.8 ± 0.1	58.6 ± 0.7
5	-1	135	0	60	2.8	4.2 ± 0.0	85.6 ± 0.5
6	1	175	0	60	4.0	3.8 ± 0.0	62.3 ± 1.9
7	0	155	-1	30	3.1	4.1 ± 0.1	81.0 ± 0.0
8	0	155	1	90	3.6	3.8 ± 0.0	69.4 ± 0.1
9	0	155	0	60	3.4	3.9 ± 0.0	74.1 ± 7.0
10	0	155	0	60	3.4	3.9 ± 0.0	69.9 ± 2.2
11	0	155	0	60	3.4	3.9 ± 0.0	74.6 ± 2.0
Opt		153		59	3.3	3.9 ± 0.1	74.0 ± 2.2

using different technologies. SBP and DRB were used as raw materials. Four different processes were evaluated to compare conventional pectin production with the emerging technologies combination proposed here, based on the process conditions described next.

Reference case (SBP CE): The process conditions for the conventional extraction of RG-I pectin were sourced from the work reported in Mao et al. [24]. According to the study, pectin is extracted from SBP by conventional acidic extraction (CE) using hydrochloric acid. The optimum extraction conditions reported were pH = 1 at 90 °C over 120 min, with a solid to liquid ratio of 2 g/43 mL. Following the extraction, the mixture was cooled down to 55 °C using cooling water (CW). The solids and liquids were separated via filtration, with the spent solids assumed to be sent to incineration with heat recovery. The liquid fraction or extract was concentrated under vacuum (concentration factor (CF) = 2) and IPA was added for the precipitation of pectin in an extract/IPA volume ratio 1:1 [22,24]. Pectin was filtered, washed with IPA and dried using hot air at 45 °C until moisture content was less than 1 % (wt.). The process flow diagram is provided in Fig. S1.

Alternative 1 (SBP MAE): The process conditions for this alternative were obtained from our previous work reported in Del Amo-Mateos et al. [14]. The optimum extraction conditions found were at 165 °C over 12 min and 10 % (w/v) solid to liquid ratio. The extracted mixture was cooled down to 55 °C using CW. The solids and liquids were separated via filtration, with the spent solids assumed to be sent to incineration with heat recovery. The liquid fraction or extract was purified and concentrated using a membrane process, including diafiltration (DF) and ultrafiltration (UF). First, the extract was cooled down to 30 °C using CW as coolant and diluted to improve the removal of impurities in the diafiltration step. Then, it was concentrated (CF = 2) in the UF step [18]. Part of the permeate was recirculated for the initial dilution of the extract and for the diafiltration process. The pectin solid product was obtained from the retentate by freeze-drying, which included a cascade refrigeration system and two sublimation chambers [26,27]. The process flow diagram of the proposed scaled-up process is provided in Fig. S2.

Alternative 2 (SBP HE): The process conditions for the extraction of RG-I pectin using hydrothermal extraction are presented as part of this study in Section 2.1.2. The optimal extraction time and temperature are discussed in Section 3.1. The solids and liquids were separated via filtration, with the spent solids assumed to be sent to incineration with heat recovery. The downstream processing for the liquid fraction or extract is assumed to be carried out in the same way as for Alternative 1. The process flow diagram of the proposed scaled-up process is provided in Fig. S3.

Alternative 3 (DRB MAE): This alternative included an upstream process, where a mechanical juice extractor was used to separate the pomace from the beetroot juice. The MAE was used to extract the pectin from the pomace using water as the solvent. The extraction conditions reported in Del Amo-Mateos et al. [3] were used: 160 °C, 5.3 min, and 10 % (w/v) solid to liquid ratio. The mixture was cooled down to 55 °C after extraction, followed by a solid-liquid filtration. The downstream processing for the extract is assumed to be carried out in the same way as for Alternative 1. The process flow diagram of the proposed scaled-up process is provided in Fig. S4.

2.3. Process simulation

The process simulation of the four pectin production processes was carried out using Aspen Plus ® V.14 [28] using the conditions described in Section 2.2 to obtain mass and energy balances at larger scales. The results were used to construct the life cycle inventories for the LCA and the basis for the TEA studies. The components were defined using native Aspen components and items not included were defined according to Humbird et al. [29]. The supplementary information provides the list of components defined (Table S1) and unit operation models used (Table S2) in the process simulation using Aspen Plus.

The capacity of the plants was assumed at 2000 t/day on a wet basis, the minimum scale of a biorefinery to be profitable [1]. For all scenarios, a plant lifetime of 25 years and 8000 working hours per year were established. The process flow diagrams, and corresponding mass balances of all proposed Alternatives can be found in Figs. S1 – S4 and Tables S3 – S6.

A more detailed description of the simulation is provided in the following subsections.

2.3.1. RG-I pectin extraction

Pectin was extracted from SBP in the Reference case and Alternative 1 and 2. In the case of Alternative 3, pectin was extracted from DRB pomace which corresponds to an upstream step involving a mechanical juice separation.

The pectin extraction process is represented in the flow diagrams in Figs. S1 – S4 for all cases. For the Reference case (Fig. S1), HCl and water are fed into the solvent tank to produce an HCl solution, which is then mixed with the SBP. The mixture of SBP and HCl solution is then fed into the CE extractor, which was modeled using experimentally determined conversions [3,22,24] of specific reactions (e.g., arabinan to AraOS) [30]. In Alternative 1 – 3 (Fig. S2 – S4), pectin extraction is carried out at pressures above the atmospheric one to avoid water evaporation (vapor fraction set to 0 %), thus requiring the addition of a pump before the extractors. The HE and MAE extractors were simulated following the same procedure as in the CE extractor.

Following the extraction, the resulting mixture is cooled to 55 °C prior to solid to liquid filtration.

2.3.2. Downstream process and solvent recovery: reference case

2.3.2.1. Pectin precipitation and air drying. As indicated in Fig. S1b, prior to the precipitation step, the extract is concentrated twice in an evaporator at vacuum conditions and cooled to 30 °C. The concentrated extract is then mixed with IPA and cooled to 8 °C using an ammonia refrigerator cycle. Details of the refrigerator cycle can be found in

Table S2. The precipitation was modeled using experimentally determined conversions [14] of specific reactions (e.g., AraOS to arabinan). The pectin is then separated from the IPA-rich stream in a filter and washed with two volumes of IPA [24]. The wet pectin is then dried using hot air at 45 °C.

2.3.2.2. Isopropanol recovery. The IPA recovery was carried out via extractive distillation according to the process reported in Hartanto et al. [31]. Briefly, the IPA-rich stream obtained after the pectin precipitation process (see Fig. S1) was subjected to distillation, reaching approximately the azeotrope composition in the distillate at 1 bar and 80.2 °C (63.5 % molar basis). Then, the distillate was conducted to an extractive column where glycerol was used as a carrier to separate the IPA from water. A distillate with a composition of 99 % of IPA (weight basis) was obtained and recirculated to the pectin precipitation unit. Finally, a recovery column was used to separate the glycerol from the water and the glycerol recovered in the bottom was recirculated to the extractive column.

A detailed description of all unit operations simulated in Aspen Plus and described here can be found in Table S2.

2.3.3. Downstream process: alternatives 1, 2 and 3

2.3.3.1. Membrane process. In Alternatives 1 – 3, the extract was purified in a membrane process using a 3 kDa molecular weight cut-off (MWCO) membrane. First, the extract was cooled to 30 °C. Then, the membrane purification system was simulated in Aspen Plus using pumps and component separators introducing the experimental separation yields [8,18] (see Fig. S2 – S4; for more information, see Table S2). The inlet pressure and flow rate per membrane were based on the conditions reported for the commercial membrane VT MAX PES 3000 Da (Synder Filtration, USA) at 5 bar and 68 m^3/h, respectively. After the membrane process, the retentate is transferred to the freeze-drying system and the permeate is recirculated to make the initial extract dilution and to the diafiltration unit [18]. The excess of permeate is assumed as wastewater.

2.3.3.2. Freeze-drying. The freeze-drying process was simulated according to Gosalvitr et al. [26] with a three-level cascade refrigeration system using NH_3 and CO_2 as refrigerants (see Fig. S2 - S4). A preliminary optimization of the freeze-drying process was carried out to select the outlet temperature of the retentate in the NH_3 cycle step to minimize the total power consumption in the compressors (results not shown). First, the retentate from the membrane process was cooled to −5 °C in the upper NH_3 refrigeration cycle. Then, the temperature decreased to −23 °C (lower NH_3 refrigeration cycle) and finally to −45 °C in the CO_2 refrigeration cycle. The drying process was carried out in two stages, following the basic principles of freeze-drying [27]. The primary drying was carried out at −45 °C under vacuum conditions (0.07 mbar), followed by increasing the temperature to 25 °C in the second drying step whilst maintaining the vacuum condition.

A detailed description of all unit operations simulated in Aspen Plus and described here can be found in Table S2.

2.4. Life cycle assessment of the RG-I pectin production processes

The environmental impacts of the pectin production processes were evaluated using the LCA methodology in accordance with the ISO 14040/44 standards [32,33]. The LCA for Experts V8.6 software [1,22,34–36]re was used to model the system and calculate the environmental impacts. The goal and scope, inventory data and assumptions, and impact assessment method are described over the next sections.

2.4.1. Goal and scope

The main goal of this study was to quantify and compare the environmental impacts of three Alternative pectin production processes from different feedstocks based on emerging extraction and downstream technologies, and with respect to the conventional process. These Alternative processes were proposed following green chemical engineering principles around solvents use, e.g. water as the extractive solvent, solvent-free downstream processes, and the extent of these improvements must be quantified in terms of environmental impacts as well as identifying opportunities for improving their design. The assessment followed a cradle-to-gate approach and included the extraction and processing of the raw materials, and the extraction and purification of pectin and recovery of solvents where applicable. In the Reference case and Alternatives 1 and 2, the production of SBP, a byproduct in the sugar industry with commercial value, was considered. For Alternative 3, DRB was assumed to be burden-free as this is currently treated as a waste [16]. However, the extraction of the pomace was included in the study. The functional unit was defined as the "production of 1 kg of pectin", and the system boundaries of the Reference and Alternative cases are presented in Fig. 1.

2.4.2. Inventory data and assumptions

The inventory data for the Reference case and three Alternatives to produce pectin are presented in Tables S8 – S11. The mass and energy flows of the scaled-up processes were obtained from the process simulations in Aspen Plus and the use of advanced calculations. The latter were employed to estimate the electricity consumption of the MAE extractor and filtration, mechanical juice separation in Alternative 3, and cooling water losses and refrigerant leakage as described below.

The electricity required by the MAE extractor in Alternatives 1 and 3 was calculated following the scale-up power and energy-based parameters described by García-Cubero et al. [23]. Briefly, this method is based on the absorbed power density (APD) and the absorbed energy density (AED). The APD (W/L), AED (J/L) and the energy required (E, kW) can be calculated following Eqs. 2,3 and 4:

$$APD = Q/(V \cdot t_H) \quad (2)$$

$$AED = APD \cdot t \quad (3)$$

$$E = AED \cdot V \quad (4)$$

where Q (kW) is the heat absorbed by the solvent to reach the extraction temperature obtained from the Aspen model, V (L/s) is the solvent flow, t_H (s) is the heating time considering a heating rate of 6 °C/min [3,22], and t is the extraction time.

The electricity consumption required by the filters was assumed at 5.5 kWh/t according to Piccinno et al. [37], and the Multifruit ZMF3000 (Zumex, Spain) [38], a commercially available technology, was selected as a reference for purchase cost and electricity and water consumption calculations.

Cooling water was assumed to be produced in a cooling tower. The water losses through evaporation, blowdown and windage were calculated according to the literature [39]. Briefly, 1 % of water loss due to evaporation for every temperature increment of 5.6 °C was estimated. Water losses caused by the blowdown were calculated using Eq. 5.

$$BL = EL/(CoC - 1) \quad (5)$$

where BL is the blowdown loss (m^3/h), EL is the evaporation loss (m^3/h), and CoC is the cycle of concentration, which represents the accumulation of the total dissolved solids and has a typical value between 3 and 7. The average value was used. Finally, water losses by windage were calculated as 0.2 % of the water circulation rate.

For the refrigeration cycles, the initial amount of refrigerant is obtained from the Aspen simulations, while the refrigerant make-up required due to leakage is estimated based on an annual leakage rate of 15 % [26].

Secondary data was sourced from the Ecoinvent database V3.3 [40]. The study is based in Spain and therefore, the electricity grid mix of

Spain was modelled based on the national data available for 2023 [41]. In the absence of Spanish specific relevant processes in the database, European and global data was used instead as indicated in Table S7. Spent solid, rich in glucan [14], was considered as biowaste and assumed to be treated in municipal incineration with heat recovery. The heat produced was calculated considering the low heating value (LHV) of the spent solid. LHV was analyzed in the Regional Fuel Laboratory, Larecom (Léon, Spain). The heat recovered from incineration was assumed to be exported, displacing heat from natural gas.

Alternative 3 produces raw beetroot juice in addition to pectin, thus treated as a multifunctional system in accordance with the guidelines outlined in Hauschild et al. [42]. The raw beetroot juice is assumed to be sold to a juice processing company for further conditioning such as the removal of small solids and addition of stabilizers. Mass allocation was used here given the uncertainties around juice processing options and the price of raw beetroot juice.

2.4.3. Life cycle impact assessment

The LCA modelling was carried out using the LCA for Experts V8.6 software [43], and the midpoint environmental impacts were estimated using the ReCiPe 2016 V1.1 [44] assessment method, assuming a hierarchist perspective. Whilst all 18 midpoint environmental impact categories where estimated, the following 8 environmental impact categories were presented and discussed here: climate change (CC), fine particulate matter formation (FPMF), fossil depletion (FD), freshwater consumption (FC), human toxicity – cancer (HT-C), human toxicity – non-cancer (HT-NC), land use (LU), and terrestrial acidification (TA). This selection was made according to previous research [1,45], which defined these as key categories to evaluate a biobased chemical process.

2.5. Techno-economic assessment of the RG-I pectin production processes

To evaluate the economic viability of the processes, some economic indicators were calculated including the purchase cost of the equipment (PCE), the total investment cost TIC, the production cost and the minimum selling price (MSP).

2.5.1. Production costs

The production costs of RG-I pectin were estimated using the Lang factors method described by Sinnott [46]. This method calculates, based on the PCE, the total physical plant cost (PPC), which includes equipment erection, piping and instrumentation, among others, the fixed capital cost (FCC), including design and engineering, contractor's fee and contingency, and the TIC using Eqs. 6 – 8.

$$PPC = PCE \cdot 3.15 \tag{6}$$

$$FCC = PPC \cdot 1.40 \tag{7}$$

$$TIC = FCC \cdot 1.05 \tag{8}$$

where 3.15, 1.40 and 1.05 are the Lang factors for fluid-solid processes.

The PCE was estimated from the Aspen Plus Economic Analyzer package [47], except for the membranes (Alternatives 1 – 3), MAE extractor (Alternatives 2 and 3) and juice extractor (Alternative 3). For the membrane process, the price of the commercial membrane VT MAX (PES 3000 Da) (Synder filtration, USA) was used. The microwave equipment price was estimated based on the data available in the literature, which established that the industrial equipment price per kW of power installed varies between 1000 and 5000 € [48]. An average price of 3000 €/kW was selected. The commercial price of the Multifruit ZMF3000 (Zumex, Spain) was used in Alternative 3 for the red beetroot juice extraction (upstream process).

The variable costs were calculated using the input data from the Aspen Plus simulation and the market price of the chemicals and utilities: 89 €/t SBP [49], 91.56 €/t HCl [50], 1503 €/t IPA [51], 832 €/t glycerol [52], 19.00 €/t low-pressure steam [1], 21.00 €/t medium-pressure steam [1], 25.00 €/t high-pressure steam [1], 0.10 €/kW·h [53], 0.06 €/m^3 cooling water [1], 0.07 €/m^3 process water [1], 2.16 €/kg NH_3 [26] and 0.25 €kg CO_2 [26]. The DRB price was assumed to be 0 €/kg, since it is a waste produced in the supply chain and needs to be managed. As mentioned in Section 2.4.3, the DRB juice obtained in Alternative 3 was assumed to be sold to a juice processing company for its treatment before being sold to consumers. The market price of red beetroot juice is around 5 €/kg [54]. A selling price for the processing factory of 10 % of the market price (0.50 €/kg) was assumed.

Labor operating costs were estimated using Wessel's correlation (Eq. 9), as described by García-Velásquez & van der Meer [1].

$$\log_{10} Y = -0.783 \cdot \log_{10} X + 1.252 \cdot B \tag{9}$$

where Y is the operating labor in man-hours per ton per processing step, X is the plant capacity (t/day), and B is a constant depending upon the type of process. For average labor requirements, the value of B is established as 0.

RG-I pectin does not have an established market price. Therefore, the MSP for the plant to be profitable was calculated for all scenarios. Once all the costs were estimated, the cash flows were projected over the plant lifetime (25 years), incorporating the revenues for product sales and subtracting the operating expenses. To ensure profitability, the MSP was obtained considering a net present value (NPV) of the plant of 0 € (break-even point) and an internal rate of return (IRR) of 10 % [55].

2.5.2. Sensitivity analysis

Sensitivity analyses were carried out in Alternatives 1 and 3 to study the effect of the assumptions considered.

Microwave equipment for process extraction is an emerging technology that is not available yet on a large scale. To determine its influence on pectin MSP, a sensitivity analysis of the microwave equipment price per kW of installed power was carried out, considering the range observed in the literature of 1000 – 5000 €/kW [48].

The price of the DRB assumed was 0 €/ton, since it is considered a waste. However, if DRB starts to be used as a raw material for pectin production, a high-value-added product, processing factories are expected to sell the discarded products to pectin producer companies. The influence of DRB prices was studied in order to consider this future scenario. The sensitivity analysis was made from 0 €/ton to the selling price of SBP (89 €/ton).

Finally, in Alternative 3, the sale price of DRB juice was established as 10 % of the market price. Since this value might vary depending on the juice processing company that buys the byproduct, a sensitivity analysis of this price on the pectin MSP was done from 0 €/kg to 1 €/kg.

2.6. Eco-efficiency assessment

The eco-efficiency of the different Alternatives studied has been calculated to establish a correlation between the LCA and TEA results obtained, and to identify the most promising pathway for pectin production. The eco-efficiency methodology outlined in the ISO 14045 standard [56] was used here, which defines eco-efficiency as the measurement of the economic value of a system relative to the environmental efficiency, expressed as a ratio of an economic and environmental indicator [57]. In this study, the value added (VA) of the pectin was selected as the economic indicator because VA reflects the net economic benefit generated by a product after accounting for input costs. In this study, the VA has been estimated using Eq. 10, assuming the difference between the selling price of fructooligosaccharides (FOS), a commercial prebiotic used as a proxy as it has similar properties to the pectin obtained in this study, and the MSP of pectin estimated here as described in Section 2.5.1:

$$VA_i = P_{FOS} - MSP_i \tag{10}$$

where P_{FOS} is the selling price of FOS (78.2 €/kg [58]) and MSP_i is the minimum selling price of pectin produced via process i, and calculated in Section 2.5.1.

The environmental impact categories included in the LCA are used to quantify the environmental indicators. Thus, the eco-efficiency is calculated as shown in Eq. 11 [57].

$$EE_{i,j} = VA_i / EnvI_j \quad (11)$$

where EE_{ij} is the eco-efficiency (€/impact unit) of pectin produced via process i for environmental indicator j, VA_i is the value added of pectin produced via process i (€/kg pectin), and $EnvI_j$ is the environmental indicator j (impact unit/kg pectin). The unit of environmental indicator j depends on the environmental impact category, e.g. kg CO_2 eq. for CC, kg PM2.5 eq. for FPMF.

The results of the eco-efficiency indicators for each pectin production process (Eq. 11) can be aggregated into a single overall eco-efficiency (EEC) score to aid comparison of different processes. The calculation of overall eco-efficiency includes normalization and weighting of the eco-efficiency scores as indicated by Gosalvitr et al. [59] and summarized in Eqs. 12 – 13:

$$EEN_{i,j} = EE_{i,j} / EE_{max,j} \quad (12)$$

where $EEN_{i,j}$ is the normalized eco-efficiency of pectin production process i for environmental indicator j, and $EE_{max,j}$ is the maximum eco-efficiency for environmental indicator j (impact unit/€).

$$EEC_i = \sum_{j=1}^{J} (w_j \cdot EEN_{i,j}) \quad 0 \le EEC_i \le 1 \quad (13)$$

where EEC_i is the overall eco-efficiency of pectin production process i, w_j is the weight of importance of impact j, and J is the total number of environmental impacts. Equal importance of all environmental impacts has been considered to determine the outcomes assuming a "level playing field", thus $w_j = 0.125$ for each impact category.

As for the interpretation of the results, and based on the approach adopted in this work, the highest overall eco-efficiency score corresponds to the best option and vice versa.

3. Results and discussion

3.1. RG-I pectin recovery and optimization from SBP by HE

Firstly, the effect of the temperature and the time on POS extraction from SBP by HE was appraised by an RSM-FCD (Table 1) to maximize the POS concentration of the extract. Three experiments at the central point (155 °C, 60 min) were conducted to estimate the experimental error and evaluate the reproducibility of the extraction process.

The extract pH and solid recoveries can be found in Table 1. The pH

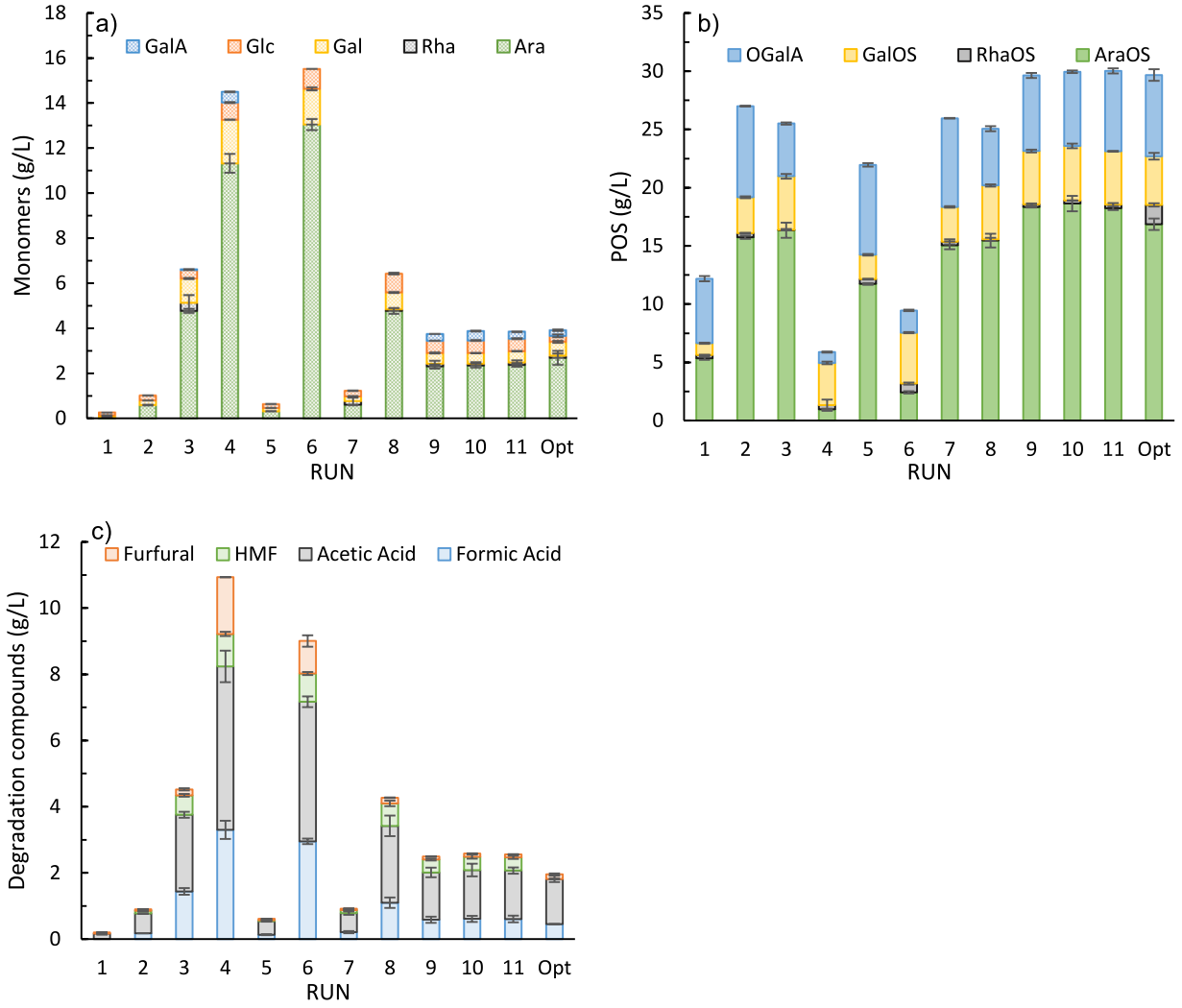

Fig. 2. RSM-FCD results: composition of extracts in each run and under optimal conditions (Opt, 153 °C, 59 min). Monomer (a), oligomer (b), and degradation compounds concentrations (c).

ranged from 3.8 to 4.4, showing no dependence on either temperature or time extraction. In the case of the solid recovery, the highest solid recovery (93.7 %) was obtained under the less severe conditions (Run 1, SF=2.5); while the lowest (58.6 %) was obtained under the most severe conditions (Run 4, SF=4.2). The results showed the high dependency between the severity of the extraction and the biomass solubility. A higher temperature and time extraction led to higher biomass solubilization, which is in accordance with previous studies using SBP as raw material for pectin recovery by MAE [14], hydrothermal processing [60] or sequential MAE and ultrasound-assisted extraction [18].

The composition in terms of monomer, oligomer, and degradation compounds content in the extracts can be found in Fig. 2. As can be seen, monomers (Fig. 2a) and degradation compounds (Fig. 2c) showed a similar behavior. The highest concentration of monomers and degradation compounds were found at the highest SF (Runs 4 and 6, SF=4.2 and 4.0, respectively). This indicates that when oligomers are subjected to high temperatures for long periods, they can degrade to their monomers and furfural, HMF, or formic acid. Due to this degradation, the lowest concentration of POS was obtained under these conditions (Fig. 2b, Runs 4 and 6).

The statistical analysis of POS concentration resulted in a second-order polynomial equation to relate POS with the independent variables (Eq. 14).

$$POS(g/L) = -614.3 + 7.4\cdot T + 2.7\cdot t - 1.4\cdot10^{-2}\cdot T\cdot t - 2.1\cdot10^{-2}\cdot T^2 - 4.1\cdot10^{-3}\cdot t^2$$

$$R^2 = 0.996; \quad R^2_{adj} = 0.990 \tag{14}$$

where T is the HE temperature (°C) and t is the extraction time (min).

The ANOVA results indicated that the regression model is highly significant (p-value < 0.0001). This suggests that the model accurately represents the relationship between the parameters studied (temperature and time) and the POS concentration.

According to Eq. 14, the positive linear coefficient of temperature and time indicates that increasing any of the parameters also increases the POS concentration. These effects were statistically significant with a p-value of 0.01 and 0.04, respectively. However, the quadratic term of

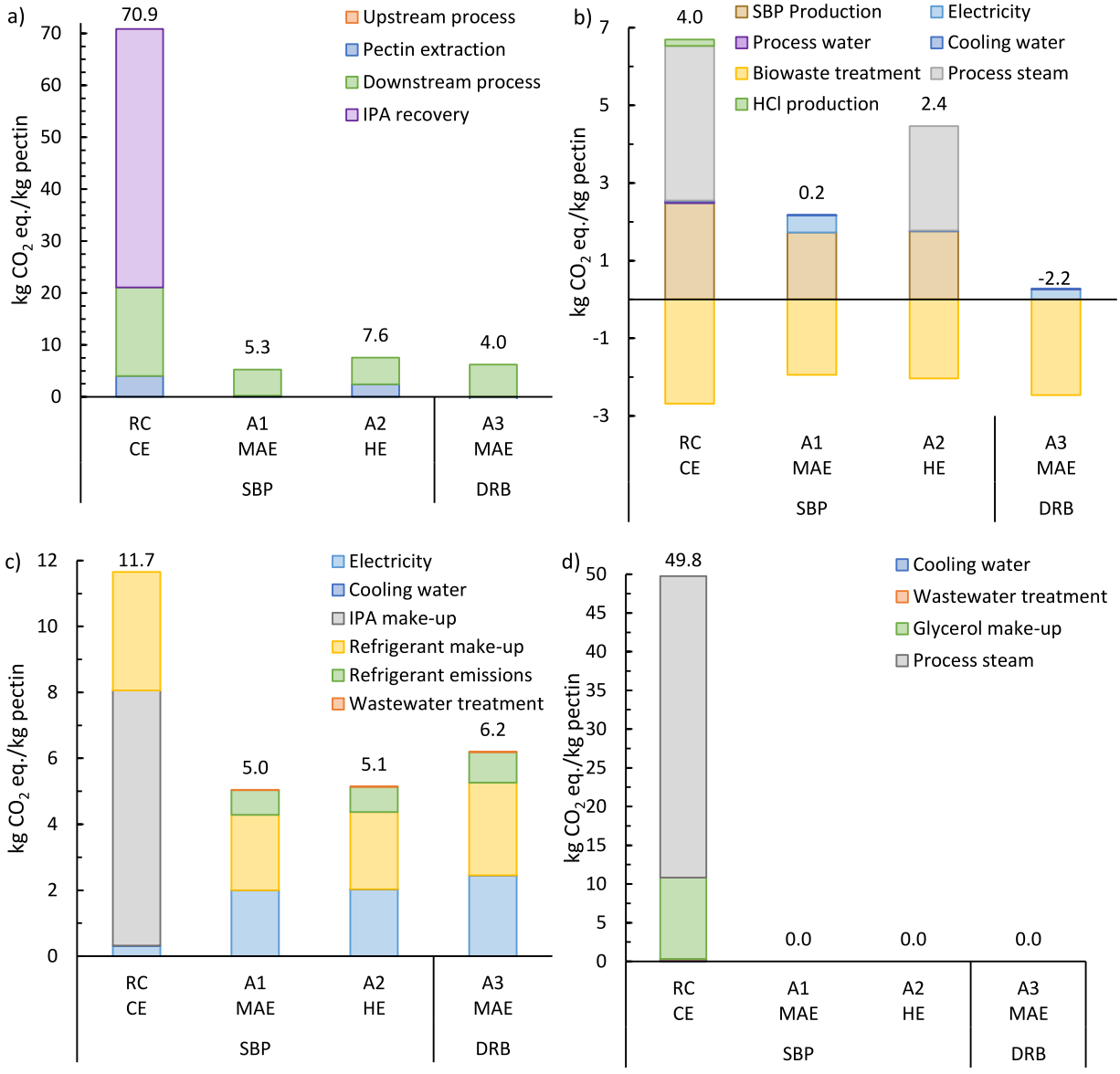

Fig. 3. Climate change impact of RG-I pectin production in the reference case (RC) and the three alternatives (A1 – A3): Impact of total process (a), pectin extraction step (b) downstream step (c), IPA recovery (d). The values over the bars show the net value of the climate change impact. SBP: sugar beet pup; DRB: discarded red beetroot; CE: conventional extraction; HE: hydrothermal extraction; MAE: microwave-assisted extraction; IPA: isopropanol.

both parameters was significant (temperature p-value = 0.0001 and time p-value = 0.003) and their negative values suggest that both variables have a peak after which the POS concentration begins to decrease. The interaction term between temperature and time had a substantial influence on POS concentration, as indicated by the significant p-value (p-value < 0.0001). The large negative coefficient implies that the combined effect of high temperature and long extraction time can lead to a decrease in the POS concentration. This statistical result follows the tendency observed in the experimental results when evaluating the effect of the extraction SF on the POS concentration described above.

Based on the model proposed, the optimal conditions for POS recovery from SBP by HE were 153 °C and 59 min. Under these conditions, the model predicted 29.7 g POS/L. Three confirmatory runs were carried out at 153 °C for 59 min to validate the model. The results can be found in Table 1 (pH extract and solid recovery) and Fig. 2 (extract composition). The POS concentration experimentally obtained was 29.7 g/L. Thus, the *t*-test concluded that there were no significant differences between the predicted and experimental concentrations, and 59.8 % of POS were recovered from SBP. This result was similar to the POS recovery yield obtained in previous work, (165 °C, 12 min), of 59.7 % indicating that HE required less temperature but a higher extraction time than MAE.

3.2. Environmental impacts of RG-I pectin production

The environmental impact results of RG-I pectin production for the Reference case and three Alternatives are presented in Fig. 3 for CC and Fig. 4 for the other environmental impact categories (for a full list of environmental impact results, see Table S12). The Reference case is discussed first to establish the benchmark against which the design improvements in the alternatives are evaluated, starting with CC followed by the other environmental impacts.

3.2.1. Climate change

The total CC impact of the Reference case (SBP CE) is estimated at 70.9 kg CO_2 eq./kg pectin, with the IPA recovery stage contributing with the large majority (70.2 %) of this impact. This is because of the steam required in the distillation columns (Fig. 3d). This is followed by the downstream process stage, accounting for 16.5 % of the total CC impact due to the IPA make-up consumption. Pectin extraction, on the other hand, contributes with less than 10 % of the total CC impact in part due to the credits obtained from the recovery of heat from the treatment of spent solids. Although this approach was consistently applied to all Alternatives, the Reference case received higher credits due to its lower pectin extraction yield compared to the more efficient extraction technologies used in the Alternative 1 – 3. Consequently, more spent solid was generated, leading to greater heat recovery.

The total CC of the Alternative processes' ranges from 4.0 to 7.6 kg CO_2 eq./kg pectin as shown in Fig. 3a, which is 89 – 94 % lower compared to the Reference case. Alternative 3 (DRB MAE) and Alternative 2 (SBP HE) have the lowest and highest CC impacts, respectively.

The downstream process stage is the dominant hotspot for all three Alternatives, contributing between 67 % and 100 % of the total CC impact, due to the electricity consumption and refrigerant make-up required in the freeze-drying step. However, the elimination of IPA in the Alternatives' downstream process allowed to reduce CC impact between 47.0 % and 57.3 % compared to the Reference case. It is observed in Fig. 3c that the CC impact of the downstream process stage in Alternative 3 is 17.7 – 19.4 % higher than the other two Alternatives due to lowest pectin yield obtained from DRB. However, this is compensated by the fact that this is a burden-free feedstock requiring minimal processing (juice extraction) as shown in Fig. 3b, thus resulting in an overall lower CC impact.

The pectin extraction stage is the second hotspot in Alternative 2 which uses an HE extractor, contributing to 32 % of the total CC impact. Process steam is required to heat the HE extractor, whilst the MAE extractor operates on electricity.

The significant reductions observed across all three Alternatives are a direct result of the elimination of organic solvents for pectin precipitation and washing, as this avoids the need of distillation for solvent recovery. Whilst a drying step is inevitable in any extraction process, the coupling of membrane filtration with freeze-drying leads to overall CC improvements in pectin production even though freeze-drying is more energy intensive compared to air drying. Freeze-drying is widely used in the food industry as it avoids the thermal degradation of bioactive compounds (such as prebiotics) [1].

A direct comparison with previous LCA studies is not possible due to differences in system boundaries, e.g. exclusion of downstream processes [16], and differences in scale, e.g. pilot-scale [20]. However, as a way of validation, these findings are consistent with trends identified in previous LCA studies. For example, Nadar et al. [13] identified the use of ethanol for pectin precipitation, pectin washing and its recovery as the main hotspot in a pectin production process employing a conventional extraction technology. García-García et al. [16] reported notable impact reductions of around 75 % when using MAE in pectin production compared with the conventional extraction technology. Gonzalez-Garcia et al. [20] assessed pectin production processes comparable to Alternative 2, employing HE and enzymatic hydrolysis, and identified the freeze-drying stage as the main hotspot in the pilot-plant.

3.2.2. Other environmental impacts

Significant reductions estimated between 48 % and 95 % were observed for all three Alternatives across the other environmental impact categories presented in Fig. 4. Similar to CC, the IPA recovery is the main hotspot in FD, FC, HT-C, HT-NC and LU for the Reference case due to the process steam required in the reboiler's columns, the glycerol make-up and the cooling water used (only for FC impact). The downstream process, on the other hand, was identified as the main hotspot in FPMF and TA because of the ammonia emissions from the refrigerator cycle, which accounted for 81.3 % and 90.8 % of the total impacts, respectively.

The other environmental impacts estimated for the three Alternatives are comparable although different hotspots were identified. For example, the downstream process stage is the main hotspot in FPMF, FD and TA for all three Alternatives, and in LU, HT-C and HT-NC for Alternative 3. This is mainly because of the refrigerant make-up and the compressors' electricity consumption in the freeze-drying step, and in the case of FPMF and TA, due to ammonia emissions. The pectin extraction stage is the main hotspot in FC and LU for Alternatives 1 and 2 due to the agricultural activities to grow the sugar beet crops needed to produce SBP. The downstream process and pectin extraction stages contribute equally to impact categories HT-C and HT-NC for Alternatives 1 and 2. SBP production, refrigerant make-up, and electricity consumption are the main contributors to these impacts, each accounting for 24 – 40 % of the total impacts. However, these results indicate that the choice of extraction technology has little influence over these impact categories whilst the choice of feedstock could potentially have some effect.

3.3. Techno-economic analysis of RG-I pectin production

The TEA of the four different scenarios for RG-I pectin production process focused on the following key economic indicators: pectin production rate, PCE, TIC, the production cost, and MSP. The results are summarized in Table 2, while Figs. 5 and 6 illustrate the PCE and variable cost distributions, respectively.

Pectin yield and quality are key parameters for contextualizing the economic assessment. As shown in Table 2, yields varied significantly across the Reference case and the Alternatives. The Reference case exhibited the lowest yield among all scenarios. Among the Alternatives, SBP-based processes achieved higher yields than the DRB-based process (Alternative 3). However, in the case of Alternative 3, this is because

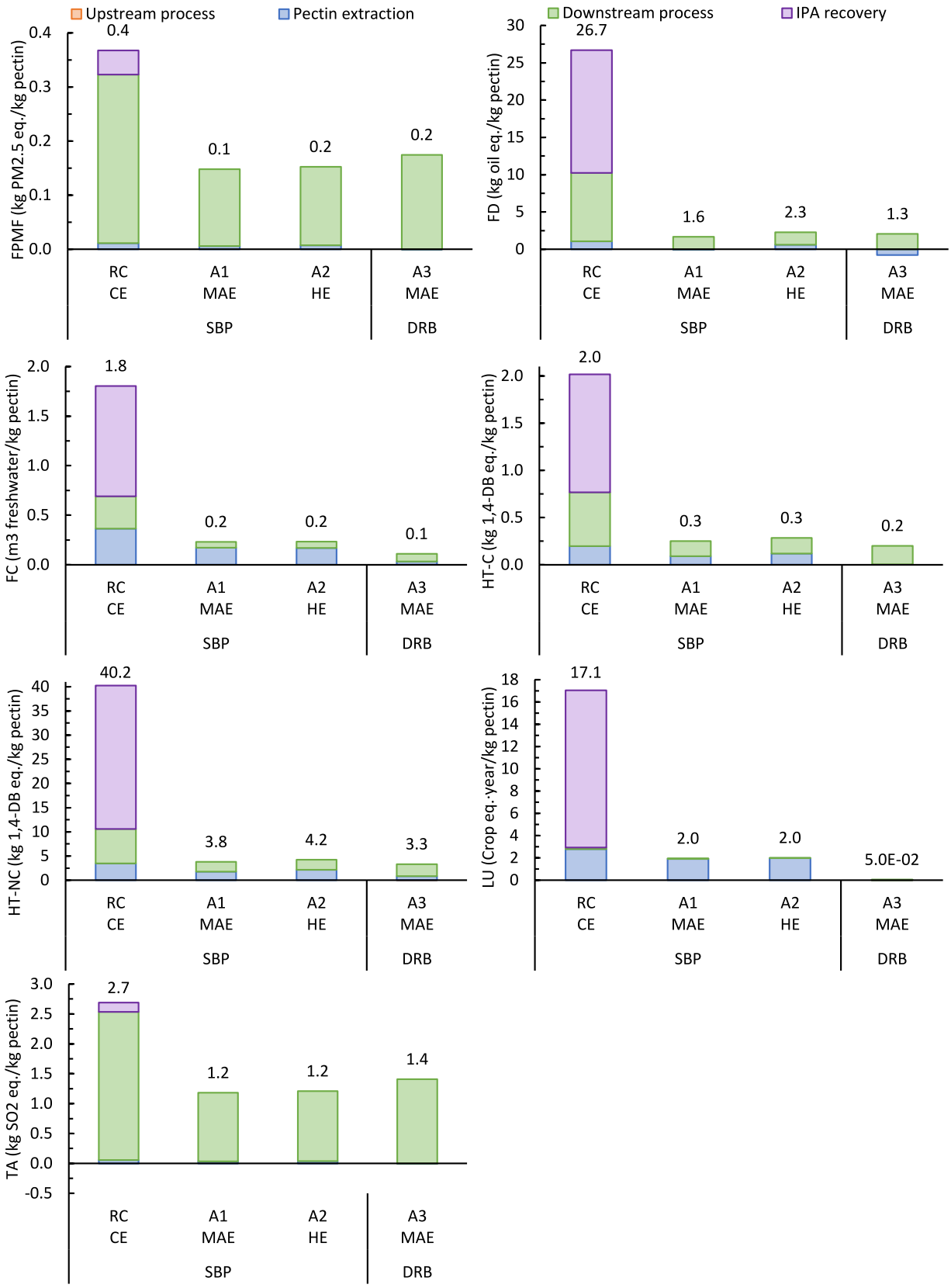

Fig. 4. Other environmental impact categories of the total RG-I pectin production process in the reference case (RC) and the alternatives (A1 – A3). FPMF: fine particulate matter formation; FD: fossil depletion; FC: freshwater consumption; HT-C: human toxicity – cancer; HT-NC: human toxicity – non-cancer; LU: land use; TA: terrestrial acidification.

Table 2
Economic evaluation of the Reference case and Alternatives for pectin production.

Scenario	Pectin yield [1] (%)	Pectin production [1] (t/year)	PCE (M€)	TIC (M€)	Production cost (M€/year)	Production cost (€/kg pectin)	Pectin MSP (€/kg pectin)
Reference case. SBP CE	21.6[2]	40278	69	318	1851	45.96	56.02
Alternative 1. SBP MAE	31.1[2]	58078	101	467	849	14.6	19
Alternative 2. SBP HE	30.5[2]	56934	48	224	742	13	16.2
Alternative 3. DRB MAE	25.5[3]	10228	21	97	323	31.5	22.4

[1] From Aspen Plus model. [2]%, g pectin/g dried SBP. [3]%, g pectin/g dried DRB pomace.
PCE: Purchase Cost of Equipment; TIC: Total Investment Cost; MSP: Minimum Selling Price.

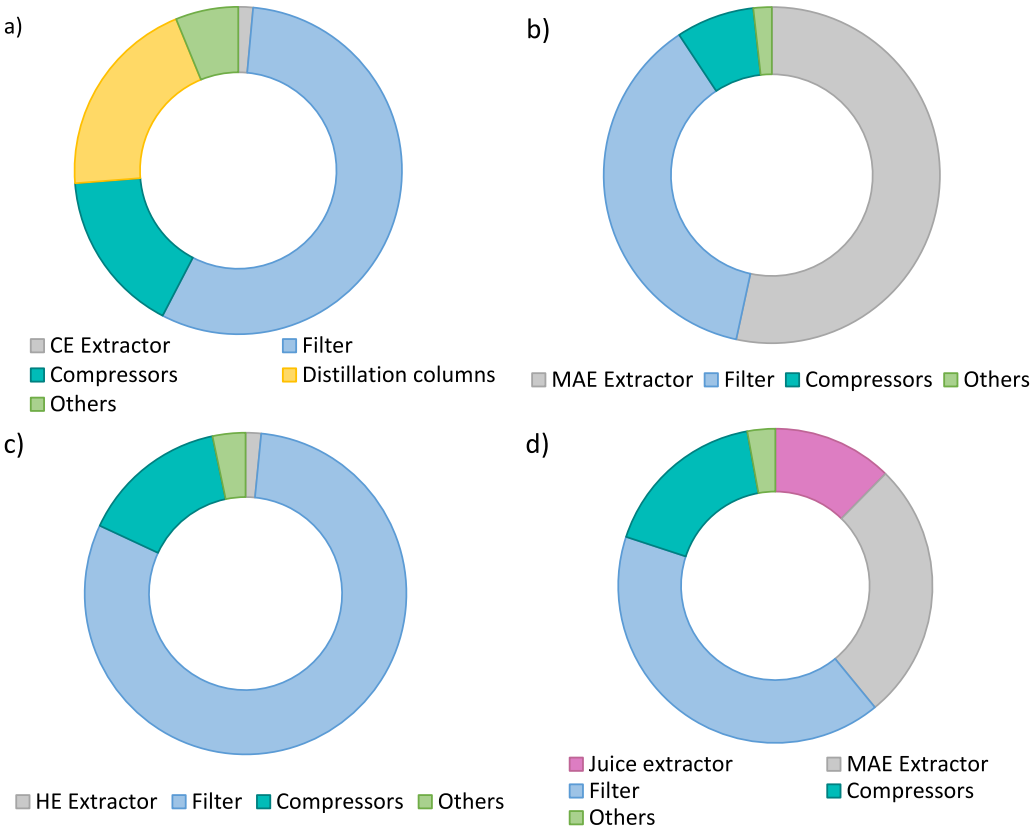

Fig. 5. Purchase cost of equipment distribution in Reference case SBP CE (a); Alternative 1 SBP MAE (b); Alternative 2 SBP HE (c) and Alternative 3 DRB MAE (d).

DRB has a higher juice content.

Pectin quality can be assessed using several parameters including composition, esterification degree, molecular weight, color or ash content, depending on the application. Since this study focuses on the use of pectin as a prebiotic, composition, and in particular the content of GalA and neutral sugars, is considered the most relevant indicator [8]. The pectin obtained from the conventional process (Reference case) was primarily rich in GalA (see supplementary material, Table S3), whereas the Alternatives produced pectins with a higher proportion of neutral sugars (see supplementary material, Table S4-S6). This difference is important because higher neutral sugar content has been associated with enhanced prebiotic bioactivity [9]. Consequently, the RG-I pectin derived from the Alternative processes can be regarded as higher-quality material, particularly suitable for functional food applications targeting prebiotic potential.

3.3.1. *Production costs*

The production cost of the processes evaluated is shown in Table 2. Alternative 1 required the highest TIC of 467 M€, followed by the Reference case at 318 M€, and Alternative 2 at 224 M€. Alternative 3 had the lowest TIC, at 97 M€, reflecting its smaller production scale. The PCE followed a similar trend, with Alternative 1 having the highest at 101 M€, while Alternative 3 having the lowest at 21 M€. The differences in capital costs can be explained by the PCE distribution shown in Fig. 5, where the equipment is named according to the process flow diagrams (Figs. S1 – S4).

In all scenarios, the filter accounted for a significant portion of the PCE, consistent with previous studies, which found that the filter used to separate spent solids from the pectin extract represented a substantial part of the investment [1]. As seen in Fig. 5a the three distillation columns required for IPA recovery accounted for 20.0 % of the total PCE. Compressors were also an important fraction of PCE in the Reference case and Alternative 2 (Figs. 5a and 5c) accounting for 16.1 % and

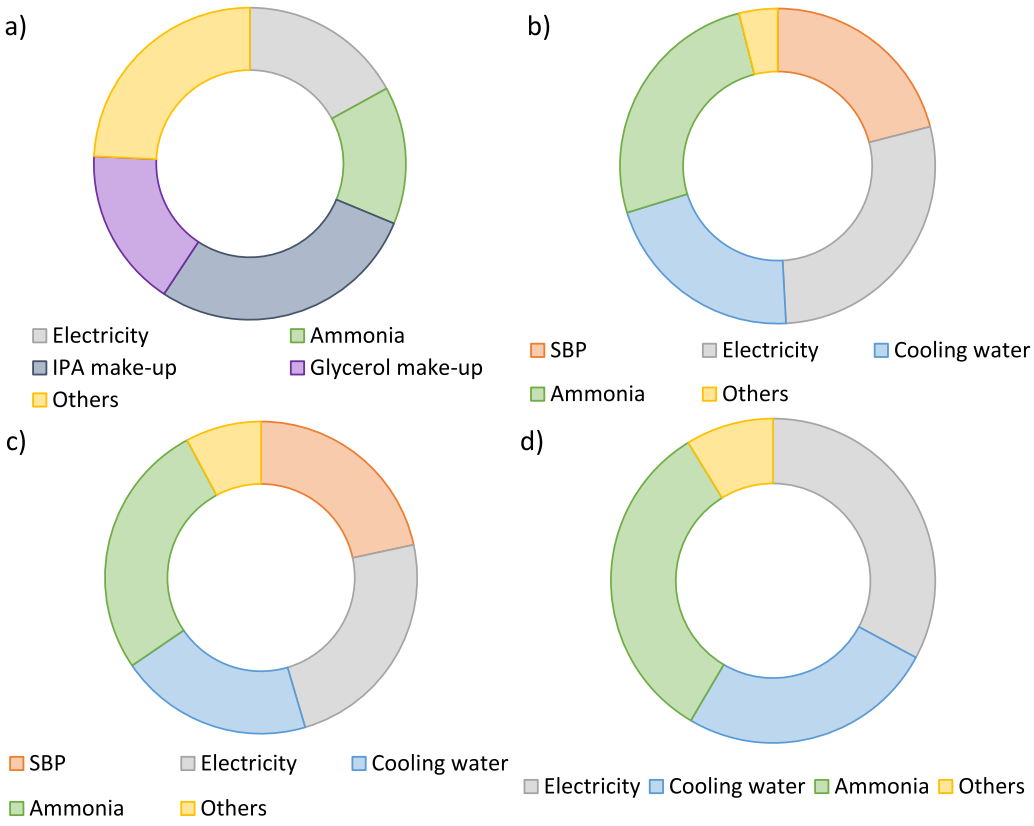

Fig. 6. Variable costs (raw materials and utilities) distribution in Reference case SBP CE (a); Alternative 1 SBP MAE (b); Alternative 2 SBP HE (c) and Alternative 3 DRB MAE (d). IPA: isopropanol; SBP: sugar beet Pulp.

14.8 %, respectively. However, when using MAE technology (Figs. 5b and 5d), the extractors accounted for 53.4 % and 26.7 % of the total PCE in Alternatives 1 and 3, respectively. This substantial contribution can be attributed to the relatively low development of technology, which resulted in higher investment costs. The higher cost of MAE technology compared to conventional extractors was also observed in other applications such as fertilizer production from organic waste [61].

The variable costs of raw materials and utilities were also analyzed for each scenario (Fig. 6). In the Reference case, the consumption of IPA for pectin precipitation significantly increased variable costs, with IPA accounting for 28.0 % of the total. The next largest contributors were electricity (17.0 %) and the glycerol required to obtain high purity IPA (16.5 %). The other variable costs shown in Fig. 6a mainly included the process steam and the SBP cost. In contrast, Alternatives 1, 2 and 3 had lower variable costs due to less solvent being required. In these scenarios, electricity, primarily consumed in the compressors of the freeze-drying and the ammonia used for refrigeration, had similar contributions. Together, these utilities accounted for 49.8 %, 55.0 %, and 65.7 % of the total variable cost in Alternative 2, 1 and 3, respectively. Lastly, the price of SBP had a notable impact on variable costs (59.3 M€/year), ranging from 7.8 % (Reference case) to 21.6 % (Alternative 1). This cost was eliminated in Alternative 3 (DRB MAE) due to the use of a discarded product.

From the costs summarized in Table 2 (PCE, TIC, and production cost), it can be concluded that emerging technologies requiring less solvent, such as IPA, can lead to lower production costs per kg of pectin, despite having higher PCE or TIC, as seen in Alternative 1.Thus, the Reference case had the highest production cost 46.0 €/kg pectin, followed by Alternative 3 (31.5 €/kg pectin), probably due to the lower amount of pectin produced, while Alternative 2 and 1 showed similar production costs, 13.0 and 14.6 €/kg pectin, respectively.

3.3.2. RG-I pectin minimum selling price and profitability

To assess the viability of the scenarios evaluated in this work, the MSP of RG-I pectin was estimated (Table 2). As described in Section 2.5, the MSP was calculated by setting an NPV of 0 € and an IRR of 10 %. The estimated MSP ranged from 16.2 €/kg pectin (Alternative 2) to 56.0 €/kg pectin (Reference case). Therefore, if the pectin is sold above the MSP, the plants would be profitable and produce benefits (NPV $>$ 0€ and IRR $>$ 10 %).

The bioactivity of pectin obtained by alternative processes was evaluated in author's previous work and showed comparable or even higher bioactivity than commercial prebiotics such as inulin and FOS [8]. Since there is no established market price for POS, they are expected to have a similar price to these commercial products.

The commercial price of inulin, at 37.5 €/kg (Energy feelings, [58]), and FOS, at 78.2 €/kg (Equisalud, [58]), are higher than the MSP estimated in Alternatives. This indicates that a potential profit margin could be achieved by selling the pectin from these scenarios, even after accounting for additional costs not included in the MSP, such as marketing, logistics, and transportation. For instance, considering the selling price of inulin, the margin would range between 15.1 €/kg and 21.3 €/kg for Alternatives. Similarly, considering FOS's market price of 78.2 €/kg, the margin could range from 55.8 €/kg to 62.0 €/kg, presenting a substantial profit opportunity. In the case of the Reference case, the estimated MSP is higher than the market price of inulin, meaning that selling the pectin at this market price would result in a negative margin. However, compared to FOS, the Reference case still offers a positive profit margin of 22.2€/kg.

3.3.3. Sensitivity analysis

Sensitivity analysis is a crucial tool to evaluate the assumptions made during TEA. Processes involving CE and HE use commercially available, well-developed technologies with established prices. However, in Alternatives 1 and 3, certain estimations were assumed, such as the MAE

extractor purchase price (3000 €/kW) and the buying and selling prices of DRB and DRB juice, respectively, in Alternative 3. The effect of these prices on the pectin MSP can be found in Fig. 7.

Firstly, according to the literature [48] the cost of MAE extractors can range from 1000 to 5000 €/kW. As can be seen in Figs. 7a and 7b, the contribution of the MAE extractor to the PCE ranged from 27.6 % to 65.6 % in Alternative 1 and from 10.8 % to 37.8 % in Alternative 3. Even at the lowest price, this contribution remains significantly higher than that of the CE or HE extractors, which accounted for less than 2 % of the PCE (Figs. 5a and 5b). In terms of pectin MSP, the price would vary from 17.0 to 21.0 €/kg in Alternative 1 and from 21.3 to 23.6 €/kg in Alternative 3. Although TIC is influenced by the MAE purchase price, this cost has no significant impact on either the MSP or on the overall viability of the plant. This finding suggests that industries requiring high-cost, less-developed technology can still achieve profitability.

In the case of Alternative 3, the results shown in Fig. 7c demonstrated the significant effect of both the DRB juice selling price and the DRB buying price on the MSP of pectin. For instance, when the DRB juice price rises from 0 to 1 €/kg, the pectin MSP drops from 40.8 to 4.0 €/kg when no DRB purchasing cost is involved, representing a reduction of 90.2 %. This reduction trend is consistent across different DRB buying prices, with lower DRB prices leading to greater reductions in the pectin MSP. Similarly, when the DRB price decreases from 89 €/ton to 0 €/ton with a DRB juice price of 1 €/kg, the pectin MSP goes from 14.4 to 4.0 €/kg, showing a 72.1 % reduction. This effect becomes more pronounced as the DRB juice selling price increases, emphasizing the impact of securing DRB at low or zero cost.

The reduction in pectin MSP ranged from 71.9 to 90.2 % when modifying the DRB juice selling price and from 20.2 to 72.1 % when varying the DRB purchase price. This indicates that, although both variables significantly impacted pectin MSP, the selling DRB juice price had a more important effect on the profitability and higher prices can lead to a more feasible plant.

These results highlight the potential of a multiproduct biorefinery model, where the revenue from selling DRB juice can offset production costs and reduce the MSP of pectin.

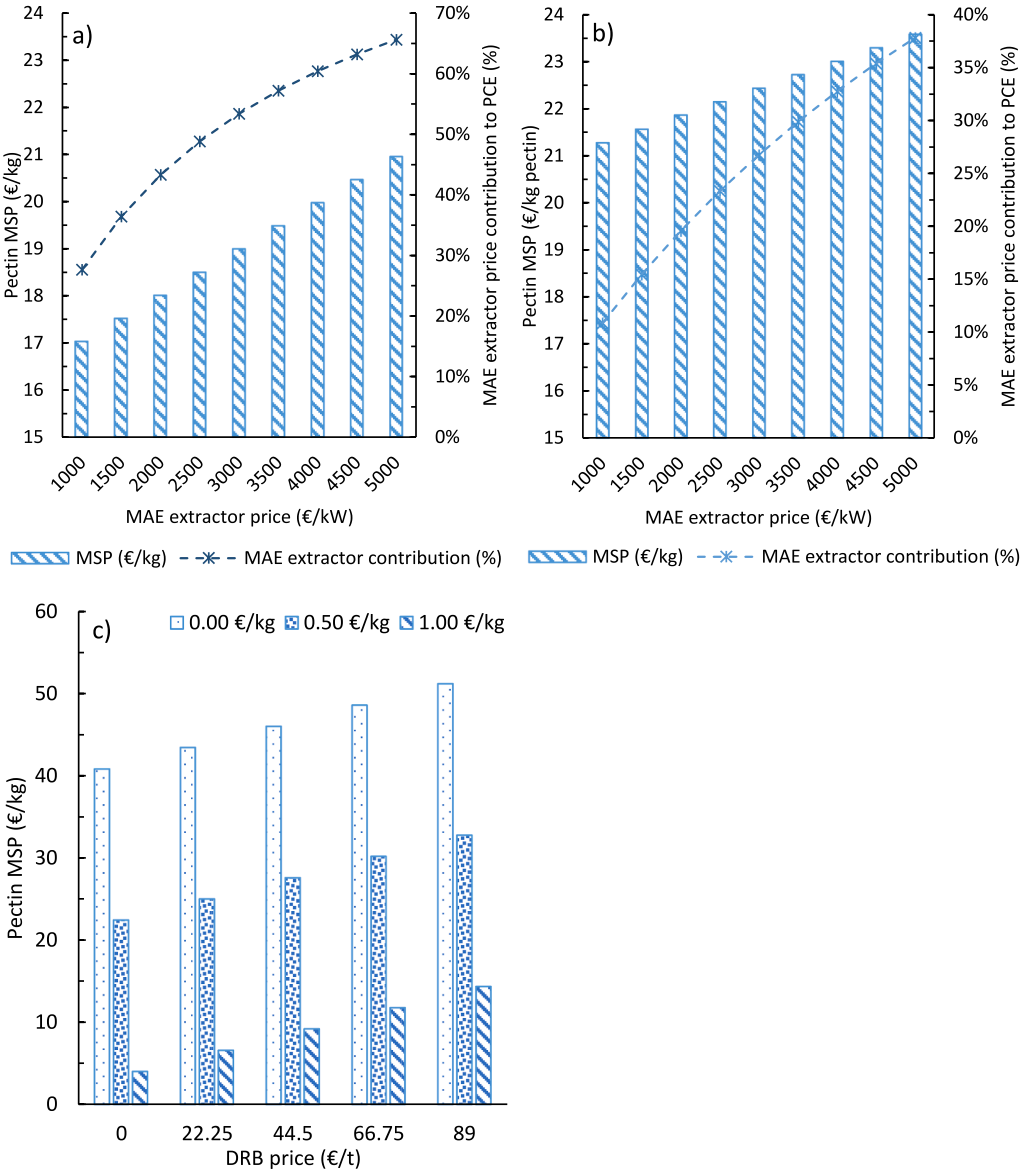

Fig. 7. Sensitivity analysis for tecno-economic assessment. Effect of MAE extractor price (€/kW) on MAE extractor contribution to total purchase cost of equipment (PCE) and pectin minimum selling price (MSP) in Alternative 1 SBP MAE (a) and Alternative 3 DRB MAE (b); effect of DRB price (0 – 89 €/t) and DRB juice selling price (0.00 – 1.00 €/kg) on pectin MSP in Alternative 3. DRB MAE (c).

3.4. Eco-efficiency assessment results

The value added and eco-efficiency results are presented in Table 3 and the overall eco-efficiency scores are provided in Fig. 8.

As can be observed in Table 3, even though Alternative 3 has the lowest value added out of the three alternatives, it achieves higher eco-efficiency in 6 out of 8 environmental indicators (CC, FD, FC, HT-C, HT-NC, LU), as it has the lowest environmental impacts for those categories. Alternative 1, on the other hand, has the highest eco-efficiency for fine particulate matter formation for the same reason. In the case of terrestrial acidification, Alternatives 1 and 2 have similar impacts (see Fig. 4), however differences in their value added explain why Alternative 2 performs best for this environmental indicator. The Reference case is the least eco-efficient option as it has the highest environmental impacts and the lowest value added.

The overall eco-efficiency results presented in Fig. 8 confirmed that the Reference case, with a score of 0.05, is the least eco-efficient option for pectin production. Among the three Alternative processes, the MAE-based options, Alternative 3 (0.91) and Alternative 1 (0.73), achieved the highest eco-efficiency scores, followed by Alternative 2 (0.62). These findings demonstrate that developing MAE extraction processes for the valorization of agricultural residues (SBP) and discarded produce (DRB) is a promising pathway that should be further investigated, with DRB showing a particular potential as an attractive feedstock for pectin production.

Fig. 8. Overall eco-efficiency score for the different pectin production processes assuming equal importance of all environmental impacts [higher scores signify higher eco-efficiency].

4. Conclusions

This study demonstrates the potential of alternative agricultural residues, SBP and DRB, as sustainable feedstocks for the production of high-value RG-I pectin. Through the integration of LCA, TEA and eco-efficiency, well-established sustainability analysis tools, and using scaled-up data from experimental work using process simulation and other advance calculations, this work has provided quantitative evidence to guide the design of more environmentally and economically viable processes beyond conventional acidic extraction and ethanol precipitation process.

The results show that replacing alcohol precipitation with membrane purification and adopting MAE for extraction significantly reduces the environmental impacts, with MAE consistently outperforming HE. DRB, considered for pectin production for the first time, emerged as the most promising feedstock when coupled with MAE.

From an economic perspective, the alternative processes achieved lower MSPs than commercial prebiotics such as inulin and FOS, highlighting their strong market potential. In addition, the sensitivity analysis concluded that the high purchase cost of microwave extractor did not compromise the feasibility of the process and valorizing DRB through a multiproduct biorefinery approach can further enhance economic feasibility.

Integrating LCA and TEA through eco-efficiency confirmed the clear advantage of MAE-based routes. The overall ranking was: A3. DRB MAE > A1. SBP MAE > A2. SBP HE > > RC. SBP CE. These findings underline the dual benefits of MAE, higher yields and lower environmental burdens, and position DRB as a novel, high-value raw material for sustainable pectin production.

Altogether, this work highlights how emerging technologies such as MAE, and membrane filtration combined with waste valorization strategies, can drive the transition toward greener, safer, and economically competitive bioprocesses, directly supporting SDG 9 “Build resilient infrastructure, promote inclusive and sustainable industrialization and foster innovation” and 12 “Ensure sustainable consumption and production patterns”.

CRediT authorship contribution statement

Susana Lucas: Writing – review & editing, Supervision, Project administration, Formal analysis, Conceptualization. **Esther del Amo-Mateos:** Writing – review & editing, Writing – original draft, Methodology, Investigation. **Rosa M. Cuéllar-Franca:** Writing – review & editing, Project administration, Formal analysis, Conceptualization.

Table 3
Value added and eco-efficiency results and heat map for the different pectin production processes and environmental impact categories [for eco-efficiency: light green denotes higher eco-efficiency, amber denotes medium, with dark green signifying the highest and red the lowest eco-efficiency].

Parameter		RC SBP CE	A1 SBP MAE	A2 SBP HE	A3 DRB MAE	Unit
Value added	VA	22.2	59.2	62.0	55.8	€/kg pectin
Climate change	VA/CC	0.3	11.2	8.2	14.0	€/kg CO_2 eq.
Fine particulate matter formation	VA/FPMF	55.5	592.0	310.0	279.0	€/kg PM2.5 eq.
Fossil depletion	VA/FD	0.8	37.0	27.0	42.9	€/kg oil eq.
Freshwater consumption	VA/FC	12.3	296.0	310.0	558.0	€/m^3 freshwater
Human toxicity – cancer	VA/HT-C	11.1	197.3	206.7	279.0	€/kg 1,4-DB eq.
Human toxicity – non-cancer	VA/HT-NC	0.6	15.6	14.8	16.9	€/kg 1,4-DB eq.
Land use	VA/LC	1.3	29.6	31.0	1116.0	€/crop eq.·year
Terrestrial acidification	VA/TA	8.2	49.3	51.7	39.9	€/kg SO_2 eq.

Piya Gosalvitr: Supervision, Methodology, Formal analysis. **Solange I. Mussatto:** Supervision, Formal analysis. **Giuliano Dragone:** Supervision, Formal analysis. **Mónica Coca:** Writing – original draft, Supervision, Formal analysis, Conceptualization. **M. Teresa García-Cubero:** Supervision, Formal analysis.

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.2025.120786.

Data availability

Data will be made available on request.

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