

## Adaptable heat pump-assisted dividing-wall column design for intensified downstream processing of bio-propionic acid

Janković, Tamara; Straathof, Adrie J.J.; Kiss, Anton A.

**DOI**

[10.1016/j.seppur.2024.127832](https://doi.org/10.1016/j.seppur.2024.127832)

**Publication date**

2024

**Document Version**

Final published version

**Published in**

Separation and Purification Technology

**Citation (APA)**

Janković, T., Straathof, A. J. J., & Kiss, A. A. (2024). Adaptable heat pump-assisted dividing-wall column design for intensified downstream processing of bio-propionic acid. *Separation and Purification Technology*, 350, Article 127832. <https://doi.org/10.1016/j.seppur.2024.127832>

**Important note**

To cite this publication, please use the final published version (if applicable).  
Please check the document version above.

**Copyright**

Other than for strictly personal use, it is not permitted to download, forward or distribute the text or part of it, without the consent of the author(s) and/or copyright holder(s), unless the work is under an open content license such as Creative Commons.

**Takedown policy**

Please contact us and provide details if you believe this document breaches copyrights.  
We will remove access to the work immediately and investigate your claim.



# Adaptable heat pump-assisted dividing-wall column design for intensified downstream processing of bio-propionic acid

Tamara Janković, Adrie J. J. Straathof, Anton A. Kiss<sup>\*</sup>

Department of Biotechnology, Delft University of Technology, van der Maasweg 9, 2629 HZ Delft, the Netherlands

## ARTICLE INFO

Editor: Dr. B. Van der Bruggen

### Keywords:

Propionic acid  
Dividing-wall column  
Reactive distillation  
Fluid separation  
Process design

## ABSTRACT

Propionic acid is a valuable platform chemical that is usually produced via fossil routes. As these are energy-intensive and eco-unfriendly processes, fermentative production of propionic acid is becoming more attractive. However, the complex downstream processing (due to low achievable product concentrations, high water content, presence of by-products and thermodynamic constraints), presents a potential barrier to scale-up this technology. This original research proposes a novel intensified large-scale (production capacity of ~ 20 ktonne/y) process for the final recovery of propionic acid after the initial biomass and counterion removal. Vacuum distillation steps ensure the recovery of a high-purity succinic acid product (>99.9 wt%) and a water stream that may be recycled to the fermentation to reduce the fresh water requirements. The main unit that follows is a highly integrated heat pump-assisted dividing-wall column (DWC), which allows the effective separation of propionic (>99.9 wt%) and acetic acid (99.4 wt%) products. Alternatively, it can serve as a reactive DWC that performs the esterification and separation of methyl propionate (99.8 wt%), methyl acetate (91.0 wt%) as higher added value products, and water by-product. If needed, extractive distillation can be implemented additionally to recover methyl acetate at higher purity (99.9 wt%). Overall, both options are proven to be economically interesting (recovery costs of 0.399 – 0.469 \$/kg considering the product price of 1.349 – 1.802 \$/kg) and environmentally attractive (3.206 – 3.678 kWh/kg).

## 1. Introduction

The preference towards bio-based chemicals is rapidly expanding driven by concerns about environmental pollution, strict regulatory constraints on industrial production, concerns about energy security, swiftly escalating prices and diminishing availability of fossil fuels [1]. However, significant improvements are needed to make industrial biotechnology competitive with petrochemical processes [2]. While numerous studies focus on genetically engineering microorganisms to improve fermentation processes, development of advanced downstream processes for large-scale production of biochemicals has not been as prominently addressed. Since recovery costs for bio-based carboxylic acids production typically account for 30–40 % of the total production costs [3], efficient downstream processes are extremely important for ensuring the competitiveness of the overall production processes. In that respect, this original research advances the development of industrial biotechnology by proposing a highly integrated heat pump-assisted dividing-wall column (DWC), with a flexible design, for large-scale recovery of propionic acid from dilute aqueous solutions or with a

conversion option towards methyl esters as higher added value products.

Propionic acid is an important platform chemical with an estimated global market of 450,000 tonnes, accompanied by a yearly growth rate of about 2.7 % [4]. Due to its antimicrobial properties, the primary usage of propionic acid is as a preservative in feed, grain, bakery and dairy, as well as in herbicides. The usage of propionic acid as a preservative is generally based on the undissociated acid which can more easily penetrate through the cell wall. Alternatively, propionic acid salts can be used in the equivalent concentrations as the pure acid. However, their effect is slower and shorter because propionic acid must be released from the salt which is often a slow process. Moreover, propionic acid is typically preferred for preserving animal feed due to its more appealing taste for animals and possible increase in feed intake. Contrarily, propionic acid salts are usually more acceptable as preservatives for human consumption due to their neutral taste and smell [5]. Additionally, sodium propionate is widely used in the pharmaceutical industry for treating wound infections, conjunctivitis and dermatoses, while propionic esters can be used as artificial fruit flavors, plasticizers and volatile

<sup>\*</sup> Corresponding author.

E-mail addresses: [tonykiss@gmail.com](mailto:tonykiss@gmail.com), [a.a.kiss@tudelft.nl](mailto:a.a.kiss@tudelft.nl) (A.A. Kiss).

<https://doi.org/10.1016/j.seppur.2024.127832>

Received 18 March 2024; Received in revised form 25 April 2024; Accepted 5 May 2024

Available online 6 May 2024

1383-5866/© 2024 The Authors. Published by Elsevier B.V. This is an open access article under the CC BY license (<http://creativecommons.org/licenses/by/4.0/>).

solvents [6]. Due to its sweetness, methyl propionate is commonly used as a flavoring agent and fragrance. Furthermore, it is a frequent solvent and platform chemical in the production of paints, varnishes [7] and other valuable chemicals such as polymethyl methacrylate [8]. Similarly, methyl acetate is widely used as a solvent for resins and oils, in paints, adhesives, nail varnish removers, etc [9]. Presently, propionic acid is mainly produced petrochemically via Reppe or Larson processes, whereby BASF, The Dow Chemical Company, Eastman Chemical and Perstorp dominate the global propionate market [4]. Due to the drawbacks associated with highly energy-intensive processes, nonrenewable feedstocks and environmental concerns, the production of propionic acid via fermentation is becoming more attractive. The most suitable producer for an industrial-scale propionic acid fermentation is *Propionibacterium* [4]. So far, significant research effort has been invested into the fermentative production of propionic acid with the main focus on increasing fermentation titer [10–16], because the end-product toxicity remains the principal challenge for the feasibility of the large-scale process. The high permeability of cell membranes towards undissociated forms of carboxylic acids can cause acidification of cytoplasm, ATP exhaustion and dissipation of proton-motive force [3]. Even though fermentation at low pH has been attempted for some acids (acetic acid, citric acid, succinic acid, etc.), exposing *Propionibacterium* strains to low pH leads to inhibition and low product concentrations [17]. Contrarily, cell membranes are significantly less permeable to the anions of carboxylic acids. Thus, a commonly used approach is fermentation in a pH range higher than the  $pK_a$  of carboxylic acid (4.87 for propionic acid) which leads to the transformation of the carboxylic group to the carboxylate group. This is usually performed by the addition of a base (e. g. NaOH,  $Ca(OH)_2$ , or others), which results in salt formation [3].

In the fermentative production of propionic acid, common by-products are acetic and succinic acids [12–15]. However, due to the necessary pH regulation in the fermenter, these acids are also present as salts in the fermentation broth. Therefore, several steps are required for the downstream processing to obtain carboxylic acids as final products:

removal of cells and debris, counterion removal to obtain carboxylic acids from salts, water removal (preconcentration) and final purification to obtain high-purity end-products (Fig. 1). The initial removal of cells and debris after the fermentation is done by well-known methods such as filtration. Furthermore, several alternative methods have been studied for the removal of counterions. Precipitation using sulfuric acid, with co-production of gypsum ( $CaSO_4 \cdot 2H_2O$ ) can be very convenient if  $Ca(OH)_2$  is used for the pH control in the fermentation. This method has already been proven effective for citric acid and lactic acid fermentations [3]. However, using  $Ca(OH)_2$  for pH control in propionic acid fermentation would lead to the precipitation of calcium succinate (solubility of 8.9 – 19.3 g/L [3]) and its removal with cells in the initial filtration step. Furthermore, the amount of gypsum that would be produced is slightly higher than the amount of carboxylic acids that can actually be recovered. Additionally, part of the monosaccharides would be lost due to adsorption on the gypsum surface. Besides sugars, formed gypsum would contain additional impurities such as ash and lignin, which would increase the disposal cost in industrial processing [18]. Alternatively, bipolar membrane electrodialysis (BME) can be used to obtain an aqueous carboxylic acid solution, which is sent to further purification, and an aqueous base solution, which can be reused in the fermentation for pH control. This process can efficiently combine cation and anion exchange membranes to replace cations in carboxylic acid salts with hydrogen ions [19,20]. The main cost contribution of BME is due to the membranes and energy requirements [19]. To the best of our knowledge, no detailed research on propionic acid recovery from fermentation using BME has been reported and precise costs cannot be determined. Yet, the reported BME energy requirements for the recovery of different carboxylic acids (0.60 – 0.95 kWh/kg for lactic acid, 1.5 kWh/kg for pyruvic acid, or 2.30 – 4.88 kWh/kg for succinic acid which is dicarboxylic [19]) indicate substantial minimum costs. Recovery of a solution of the three acids (propionic, acetic and succinic) from ultra-filtered fermentation broth might be possible, as indicated by exploring studies with some carboxylic acids [21,22]. Furthermore, additional

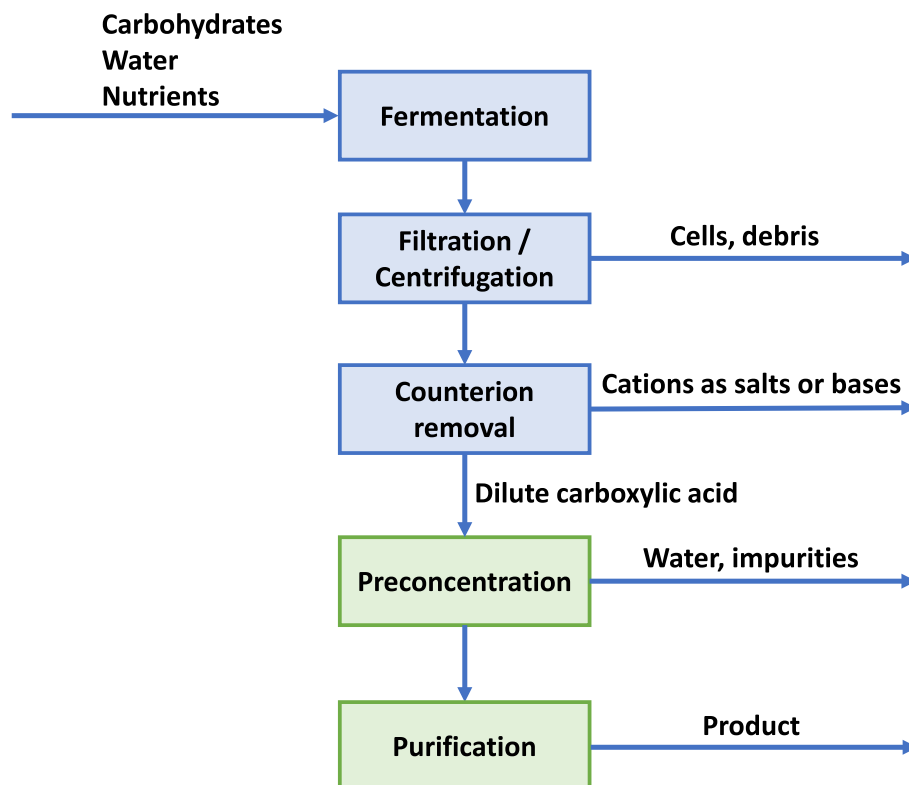


Fig. 1. General downstream processing steps for purification of carboxylic acids.

options for counterion removal are the usage of ion exchange adsorbents or extractants (which usually lead to the co-production of aqueous mineral salt solution) and thermal degradation of carboxylate salts (which may be very energy intensive) [3]. Nonetheless, whatever counterion removal method is used, further purification is required to obtain the pure carboxylic acid. Additionally, in-situ recovery of propionic acid by extractive fermentation, adsorption with anion-exchange resin and electrodialysis fermentation has been tested on a lab scale but has not been scaled up yet [6]. Nonetheless, additional purification to obtain high-purity product will still be needed following these initial recovery steps. Consequently, this research focuses on the preconcentration and purification steps in the recovery of carboxylic acids (Fig. 1) in case these acids are obtained in an aqueous solution after the removal of cells, dissolved biopolymers, and mineral ions by membrane filtrations and BME, for example. The main goal of this work is to enhance the fermentation of carboxylic acids by developing a highly advanced downstream process for propionic acid recovery from a dilute aqueous solution, with acetic acid and succinic acid as by-products. Even though different methods have been proposed for the recovery of carboxylic acids from dilute streams (e.g. liquid–liquid extraction, membrane separations, etc.), achieved recoveries and product purities are usually insufficiently high [23–26]. Extensive usage of additional chemicals may complicate further the processing, as the potential solubility of these components in water and toxicity to the microorganisms would make recycling of water to the fermentation difficult. Moreover, available data for the recovery of carboxylic acids were usually obtained from the lab scale without considering scale-up which might result in high operating costs and additional challenges [27]. Thus, distillation was chosen as the final fluid separation technique due to its common usage on an industrial level, the possibility of achieving high recoveries and high purities of all three considered acids and the absence of chemicals that would complicate the recovery process. Besides recovering high-purity fermentation products, distillation can also recover pure water that may replace fresh water in the process (e.g. in the upstream fermentation). Furthermore, even though conventional distillation of highly dilute aqueous streams might be very energy intensive, advanced process intensification methods might substantially reduce overall energy requirements.

In that context, this paper is the first one to propose an adaptable heat pump-assisted dividing-wall column that can be either used for the separation of propionic and acetic acid from an aqueous stream (case 1) or as a reactive DWC to produce and separate the methyl esters of these acids (cases 2a and 2b) as higher added value products. Given that concentrations of carboxylic acids after the counterion removal might differ from those in the fermenter, a sensitivity analysis was conducted to evaluate the performance of the developed processes if feed streams were more dilute.

## 2. Problem statement

The most common by-products reported in significant concentrations in the propionic acid fermentation are acetic and succinic acid [12,13,15,28,29]. Given the absence of available data on the concentrations of these acids in the aqueous mixture obtained after removing cells, dissolved biopolymers, and minerals, concentrations equal to those in the fermenter were assumed for the base case. Subsequently, a sensitivity analysis was performed to evaluate the effects of lower product concentrations in the aqueous solution. Thus, the feed stream to the recovery process is a dilute aqueous solution containing about 7.2 wt % propionic acid, 2.7 wt% succinic acid and 2.3 wt% acetic acid, while water makes the remaining > 87.8 wt% [13]. Acetic acid and succinic acid are substantial by-products in propionic acid fermentation, so they may be worth recovering and valorizing, although their separation is also very challenging. Therefore, the aim of this original research is to enhance the downstream processing of carboxylic acids after fermentation by developing an intensified process that can simultaneously

recover three considered acids (propionic, acetic and succinic acids) from an aqueous solution in a cost-effective and energy-efficient manner.

Depending on the used microorganism and its metabolic pathway for propionic acid synthesis, some lactic acid may be present besides already mentioned by-products. However, it has been shown that lactic acid concentrations in the fermentation broth can be minimized in several ways. Firstly, lactic acid can be catabolized by microorganisms to produce propionic acid. Accordingly, propionic acid titer and fermentation yield have been reported to increase even after substrate exhaustion until lactic acid was consumed [16]. Furthermore, a proper choice of operating conditions in the fermenter can lead to a complete absence of lactic acid in the final broth while concentrations of acetic and succinic acids are not nearly as much affected. For example, an increase in the fermenter's headspace pressure resulted in a significantly higher titer of propionic acids while lactic acid was completely depleted. Moreover, a lower agitation rate in the fermenter resulted in a lower concentration of lactic acid but this effect is less significant compared to the mentioned pressure increase [30]. As the presence of lactic acid in the final fermentation broth can be minimized or completely mitigated, it was not included in this study.

Besides the high content of water, which is lighter boiler compared to the carboxylic acids (Table 1), thermodynamic constraints (e.g. water – propionic acid azeotrope [31] and a tangent pinch between water and acetic acid – see Table 1) are additional complications for the recovery process. When the production of methyl acetate and methyl propionate is preferred over recovering carboxylic acids, four more azeotropes can potentially be formed: e.g. methanol – methyl acetate, water – methyl acetate, methanol – methyl propionate and water – methyl propionate (see Table 1) [31]. Therefore, a reliable property method is necessary to properly describe the interactions among various components and ensure the validity of the overall process design.

## 3. Results and discussion

This section gives an overview of the main results of this work. Rigorous simulations for every process operation were developed in Aspen Plus. Even though there is no theoretically guaranteed global optimum for non-convex mixed-integer nonlinear problems (MINLP), such as optimization of chemical processes, decreasing the external energy supply requirements will result in a reduction of the total recovery costs (which are dominated by OPEX). Notably, the thermal and electrical energy supply costs make a significant part of OPEX. Thus, the primary emphasis during the process design was on minimizing overall energy requirements. The considered decision variables were number of stages, location of the feed stage, reflux ratio, reboiler duty, bottoms rate, liquid split ratio, compression ratio, temperature difference for

**Table 1**  
Boiling points of pure components and azeotrope formation (at 1 bar).

Pure components		Azeotrope		
Component	Tb [°C]	Component	Mass fraction	T [°C]
Methyl acetate	57.05	Methanol	0.1810	53.60
Methanol	64.53	Methyl acetate	0.8190	
Methyl propionate	79.46	Water	0.0123	
Water	100.02	Methyl acetate	0.9877	61.91
Acetic acid	118.01	Methanol	0.4697	
Propionic acid	141.14	Methyl propionate	0.5303	
Dimethyl succinate	196.37	Water	0.0900	71.59
Monomethyl succinate	217.51	Methyl propionate	0.9100	
Succinic acid	317.60	Water	0.8120	99.52
		Dimethyl succinate	0.1880	
		Water	0.8769	
		Propionic acid	0.1231	100.01

heat exchange, operating pressure, etc. Moreover, numerous constraints were taken into account including temperature limitations, minimizing loss of valuable fermentation products, high purities of recovered products that satisfy market requirements, high purity of water streams to allow recycling to the fermentation, etc.

### 3.1. Thermodynamic modeling

UNIQUAC property method coupled with Hayden-O'Connell correction model for vapor phase association was chosen as the most suitable to describe the thermodynamic properties in a complex mixture of components [32]. Missing binary interaction parameters were estimated using the UNIFAC method based on group contributions. To ensure the validity of the developed property model, this was tested against experimental data from the literature [33–39]. Values of the obtained binary interaction parameters and comparison with experimental data, which shows a very good agreement, are presented in the [Supplementary Information](#) file.

### 3.2. Pre-concentration step

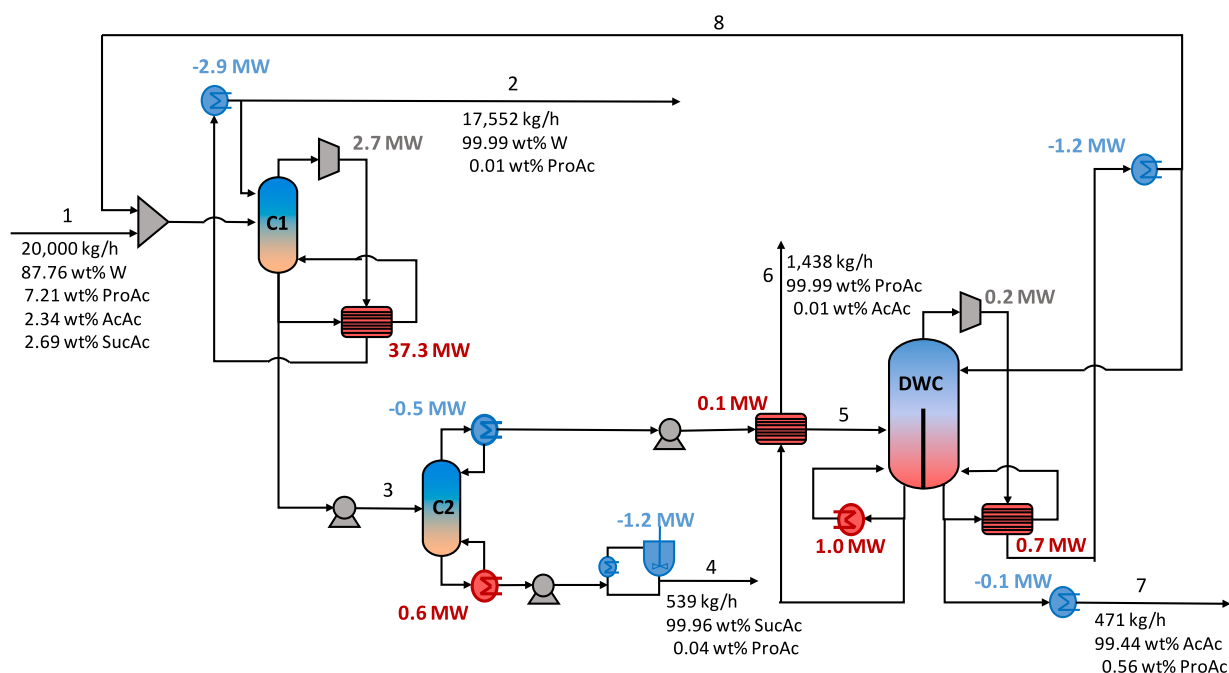
As the feed stream for the recovery process is very dilute (>87.8 wt% water), an initial step is to remove most of the water. This step is important to ensure efficient recovery of pure carboxylic acids and is critical in case of further esterification. Being the lightest component, water is obtained as the distillate, while a mixture of carboxylic acids with some water is obtained as the bottom product of column C1 ([Fig. 2](#), [Fig. 3](#) and [Fig. 4](#)). Due to the mentioned difficulties in water – acetic acid and water – propionic acid separations, this initial distillation is performed under vacuum. The operating pressure (0.08 bar) was chosen to minimize the energy requirements for this step while keeping the condensation temperature appropriate for the usage of inexpensive cooling water. Considering low-pressure operation, structured packing type Mellapak 250 (with a pressure drop of 0.225 mbar per theoretical stage [40]) was selected for column C1. The distillate-to-feed ratio for this column (0.87 on a mass basis) was determined to ensure the removal of about 99 % of the water from the feed while keeping the loss of acetic and propionic acid below 0.5 % each. The bottom product from

column C1 contains three carboxylic acids with some water and it requires further processing.

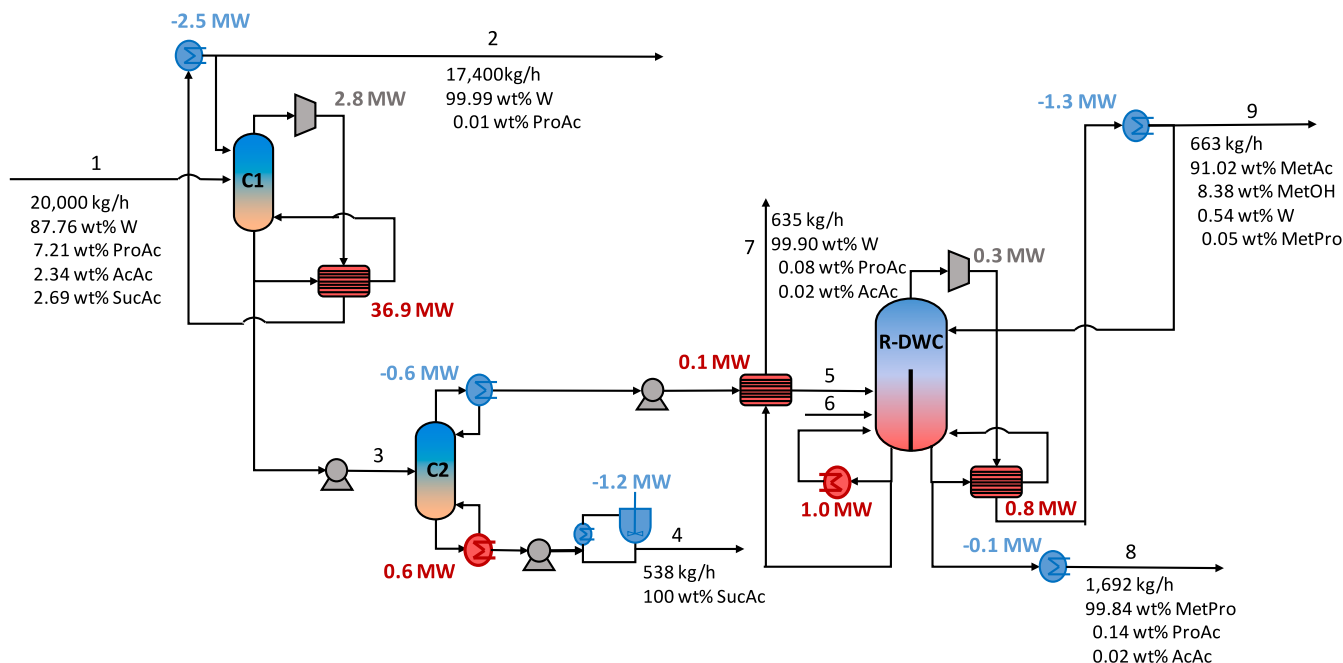
Despite being very energy intensive, this distillation separates relatively close boiling components, presenting an opportunity for the implementation of heat pumps. Mechanical vapor recompression (MVR), which implies using compressed top vapor from the distillation column to evaporate the bottom liquid, is an appropriate heat pump design for this column [41]. The effectiveness of the implemented heat pump system can be evaluated by calculating the coefficient of performance (COP), which is the ratio of upgraded heat (exchanged between the compressed vapor and bottom liquid) and required compressor duty. COP values greater than the electrical to thermal conversion factor (conservative value of 2.5 [42]) confirm the possible energy savings (n. b. the ideal COP for heat pumps is above 2.0 [43]). The COP of the MVR system applied to column C1 is 13.2 – 13.6, proving that very high energy savings are feasible. Besides the significant reduction in energy requirements, the implementation of MVR system is an important step toward electrification as it replaces large amounts of thermal energy with much less electrical energy.

### 3.3. Succinic acid separation

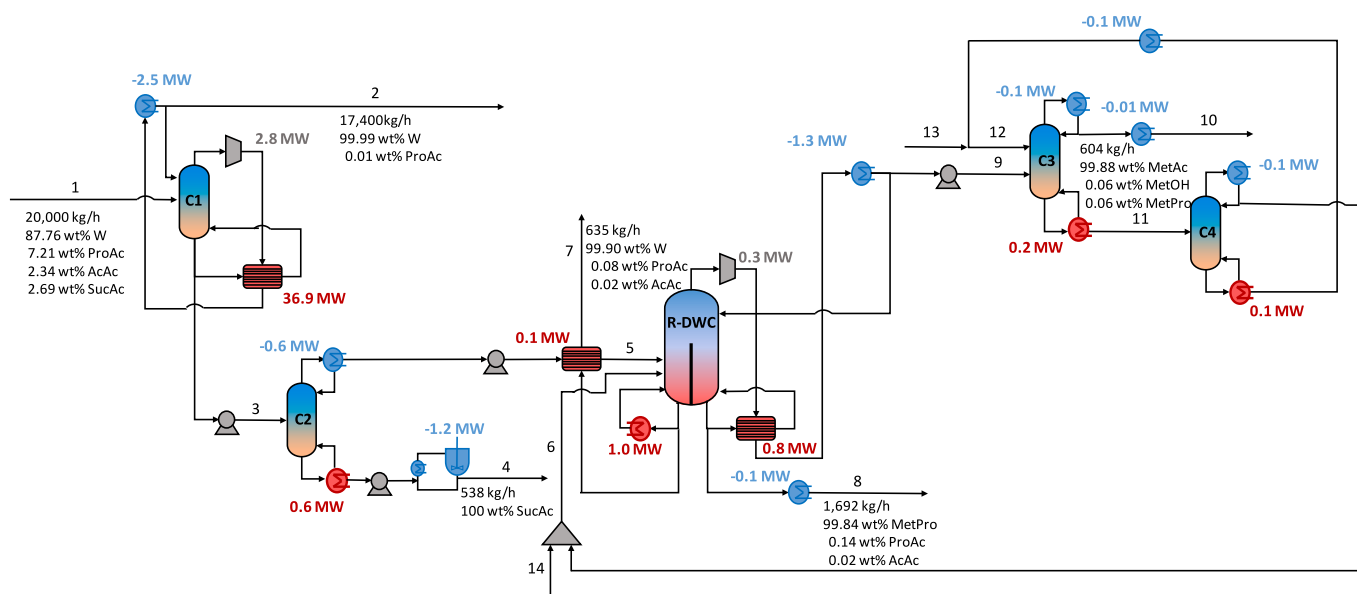
As succinic acid has a much higher boiling point than propionic acid, it can easily be separated from the rest of the mixture as a bottom product in distillation column C2 ([Fig. 2](#), [Fig. 3](#) and [Fig. 4](#)). Due to the dicarboxylic nature of succinic acid, its esterification would lead to the formation of both mono- and di-methyl succinate [44]. Additionally, the formation of a water – dimethyl succinate azeotrope (see [Table 1](#)) might further complicate the recovery process. Therefore, the best option on a large-scale, is to remove succinic acid prior to further separation of acetic and propionic acid or their methyl esters. Thus, separation of succinic acid is performed in distillation column C2. Due to the large boiling point difference, this separation is relatively easy. However, reduced pressure is required (0.10 bar) to ensure that the reboiler temperature (~243 °C) is still suitable for the usage of high-pressure steam. As a result of the reduced operating pressure, Mellapak structured packing type 250 was also assumed for this column [40]. Despite the high reboiler temperature, there is no risk of succinic acid



**Fig. 2.** Downstream process for the recovery of propionic, acetic and succinic acid – Case 1, conditions and compositions of the numbered process streams are given in [Table 2](#) (abbreviations: W – water, ProAc – propionic acid, AcAc – acetic acid, SucAc – succinic acid).



**Fig. 3.** Downstream process for the recovery of propionic, acetic and succinic acid – Case 2a, conditions and compositions of the numbered process streams are presented in Table 3 (abbreviations: W – water, ProAc – propionic acid, AcAc – acetic acid, SucAc – succinic acid, MetPro – methyl propionate, MetAc – methyl acetate, MetOH – methanol).



**Fig. 4.** Downstream process for the recovery of propionic, acetic and succinic acid – Case 2b, conditions and compositions of the numbered process streams are presented in Table 3 (abbreviations: W – water, ProAc – propionic acid, AcAc – acetic acid, SucAc – succinic acid, MetPro – methyl propionate, MetAc – methyl acetate, MetOH – methanol).

decomposition at the defined conditions [45]. The bottom product from column C2 is pure liquid succinic acid (100 wt%), while the top product is an acetic acid – propionic acid – water mixture that needs to be further treated. Liquid succinic acid product can be pumped to the atmospheric pressure and cooled which would result in crystallization. In case some non-volatile impurities were not removed in the previous processing steps (filtration, BME, etc.), they will end up in this stream and additional purification might be required.

### 3.4. Case 1: Recovering pure propionic and acetic acid

Due to the thermodynamic constraints, several steps are required for recovering pure propionic acid and also pure acetic acid. The flowsheet of this process is presented in Fig. 2, while the composition and condition of the main process streams are given in Table 2.

Our initial reasoning was that propionic acid might be obtained as the bottom product of the first distillation column as it has the highest boiling point in the acetic acid – propionic acid – water mixture that is obtained as the top product from column C2. The top product from this column would be a water – acetic acid mixture with a small amount of



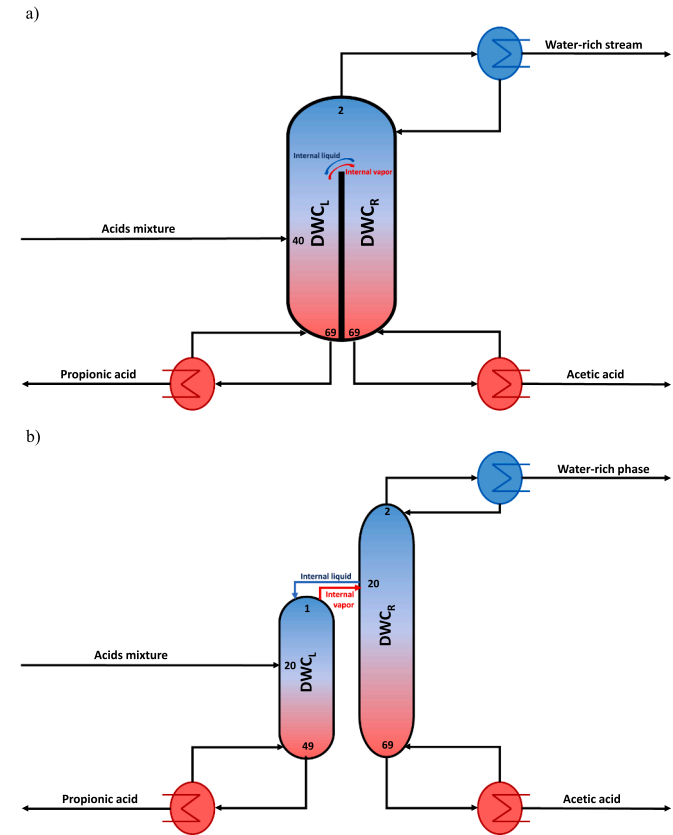
**Table 2**  
Conditions and compositions of the main streams from Fig. 2 (Case 1).

Stream	1	2	3	4	5	6	7	8
Temperature [C]	30.0	41.5	53.9	30.0	113.6	155.2	30	99.7
Pressure [bar]	1.000	0.080	0.150	1.000	1.300	1.552	1.552	1.000
Flowrate [kg/h]	20,000	17,552	2,623	539	2,084	1,438	471	175
Mass fractions								
Water	0.8776	0.9999	0.0668	0.0000	0.0841	0.0000	0.0000	0.9979
Propionic acid	0.0721	0.0001	0.5494	0.0004	0.6912	0.9999	0.0056	0.0008
Acetic acid	0.0234	0.0000	0.1785	0.0000	0.2247	0.0001	0.9944	0.0013
Succinic acid	0.0269	0.0000	0.2053	0.9996	0.0000	0.0000	0.0000	0.0000

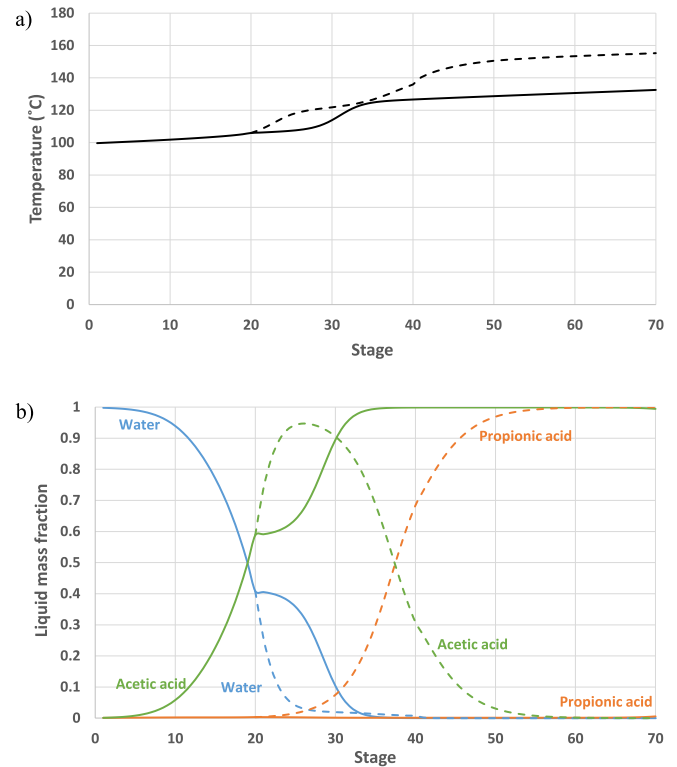
propionic acid, that needs to be further treated in the next distillation column. Acetic acid would be obtained as the bottom product from this column, while the top product would be mainly water, with small amounts of propionic and acetic acid. To avoid any loss of valuable fermentation products, this stream can be recycled to column C1. Therefore, a sequence of two distillation columns would lead to high-purity propionic acid and acetic acid products. In an improved design (Fig. 2), these two columns were merged into one dividing-wall column (DWC) with a common overhead section and divided bottom section (Fig. 5). This highly integrated system consists of only one column shell, one condenser and two reboilers. Since the DWC unit is not readily available in Aspen Plus, it can be simulated as the thermodynamic equivalent combination of distillation columns (Fig. 5). The left and right parts of DWC are presented as  $DWC_L$  and  $DWC_R$ , respectively. The number of stages, feed stage location, reflux ratio, bottom rate and liquid split were varied to minimize the energy requirements for DWC while ensuring sufficient purities of the recovered acids. Thus, propionic acid (>99.9 wt%) is recovered as the bottom product from  $DWC_L$ , while acetic acid (>99.9 wt%) is recovered as the bottom product from  $DWC_R$ .

The DWC has 70 stages in total, whereby the convention in Aspen Plus implies that the first stage is the condenser and the last stage is the reboiler. The wall that separates  $DWC_L$  and  $DWC_R$  is placed along the bottom 49 stages. The feed stream containing a mixture of water, propionic and acetic acid, enters the DWC at the 40th stage (20th stage of  $DWC_L$ ). Since  $DWC_L$  does not have a condenser, all its top vapor goes to  $DWC_R$ . About 52 % of the top liquid flowing down in  $DWC_R$  is redirected to  $DWC_L$ , to ensure sufficient liquid flow at this side of the DWC. As the temperature difference (Fig. 6) on both sides of the wall in the DWC is not large, special thermal insulation is not required. Fig. 6 illustrates the mass fractions of components in the liquid phase. As expected, mass fractions of propionic and acetic acid increase up to almost 1 at the bottom of the left and right sides of the DWC ( $DWC_L$  and  $DWC_R$ ), while the mass fraction of water increases at the top of the DWC. Sieve trays with a pressure drop of 8 mbar per tray were assumed for DWC internals [46]. The required diameter of the DWC is 1.2 m, whereby the required tray areas for  $DWC_L$  and  $DWC_R$  are approximately equal.

There is no need to perform the described separations under reduced pressure since low-pressure steam can provide sufficient heat for both reboilers. Therefore, the water – acetic acid – propionic acid mixture obtained from the top of column C2 needs to be pumped to the



**Fig. 5.** DWC designed for case 2a (a) and the equivalent sequence of distillation columns (b), numbers in the column indicate column trays.



**Fig. 6.** Temperature (a) and composition (b) profiles of DWC designed for case 1 (right side – full line, left side – dotted line).

atmospheric pressure. Afterward, it becomes a subcooled liquid that can be heated using the propionic acid bottom product to reduce total reboiler duty for the DWC. Furthermore, MVR can be applied using the top compressed vapor from the DWC to heat the bottom liquid from  $DWC_R$ . This reboiler is a better option for MVR than a  $DWC_L$  reboiler due to the lower temperature and lower compressor duty required. The COP of the applied heat pump system is around 3.1, hence higher than the 2.0 that is required to be effective [43], which proves the energy efficiency of this heat pump system.

### 3.5. Case 2: Recovering methyl esters of propionic and acetic acids

Instead of recovering pure propionic acid and acetic acid, an alternative is to obtain their methyl esters, which have higher market values. The processes proposed for recovering propionic and acetic acid in the forms of methyl esters will be referred to as case 2a and case 2b, for a final purity of methyl acetate of 91.0 and 99.9 wt%, respectively. These two cases are presented in Fig. 3 and Fig. 4, respectively, while the composition and condition of the main process streams are given in Table 3. Thus, the flexibility of the recovery process was increased by transforming the previously designed DWC into a reactive dividing-wall column (R-DWC). Methanol was selected as the reactive agent because it forms lighter esters than higher alcohols do. Moreover, unlike higher alcohols, methanol does not form an azeotrope with water. Thus, the purification after esterification is more straightforward if methanol is used as a reactive agent. The simulations in Aspen Plus were performed using a combination of equilibrium models for vapor–liquid equilibrium, and rate-based models for kinetics. The kinetic model includes two reversible reactions that are equilibrium-limited:



As previously recommended for the reactive esterification of lactic acid [44], Amberlyst 36 catalyst (strong acid catalyst type, density of 800 g/L, minimum capacity of 5.40 eq/kg, with maximum operating

temperature of 150 °C) was chosen [47]. Kinetic data for the esterification and hydrolysis reactions were gathered from the literature [48,49] and adjusted for the power law kinetics expression that is commonly used in Aspen Plus for defining the kinetics of rate-based reactions. The obtained pre-exponential factors were expressed per mass of catalyst, while catalyst density (800 g/L) allows conversion into catalyst volume-based units. Mole fractions of components were used as the basis for the rate expressions, whereby liquid was defined as the reactive phase. Table 4 provides the kinetic parameters.

A sequence of at least two distillation columns is required to perform esterification and separate the formed products (water, methyl acetate and methyl propionate). The esterification reactions are performed in the first reactive distillation column. Being the highest boiling product of the esterification reactions, water can be separated at the bottom of this column. Furthermore, methyl propionate and methyl acetate are separated as the bottom and top products from the second distillation column. As in case 1, these two columns can be integrated into one reactive dividing-wall column (R-DWC) with a common overhead section and divided bottom section (Fig. 7). Generally, R-DWC has been previously proven to be highly competitive compared to the sequence of conventional distillation columns [50,51]. Chemical reactions and separation of more than two products can effectively be performed in only one equipment unit, as proven in real-life industrial examples [50,52,53]. This column was simulated as an equivalent sequence of two

**Table 4**

Kinetic parameters for the esterification and hydrolysis reactions.

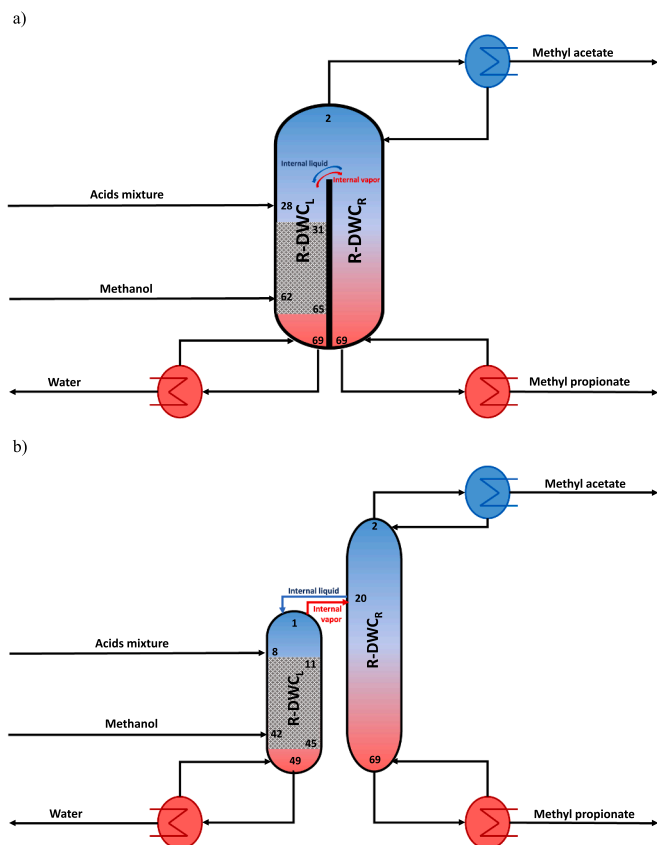
Equilibrium reaction	Reaction type	Pre-exponential factor (kmol/kg <sub>cat</sub> s)	Activation energy (kJ/kmol)
$ProAc + MetOH \rightleftharpoons MetPro + W$	Esterification	$4.27 \cdot 10^3$	45,282
	Hydrolysis	$8.47 \cdot 10^{-1}$	26,650
$AcAc + MetOH \rightleftharpoons MetAc + W$	Esterification	$8.20 \cdot 10^3$	45,600
	Hydrolysis	$4.87 \cdot 10^2$	41,924

**Table 3**

Conditions and compositions of the main streams from figures Fig. 3 and Fig. 4 (Cases 2a and 2b).

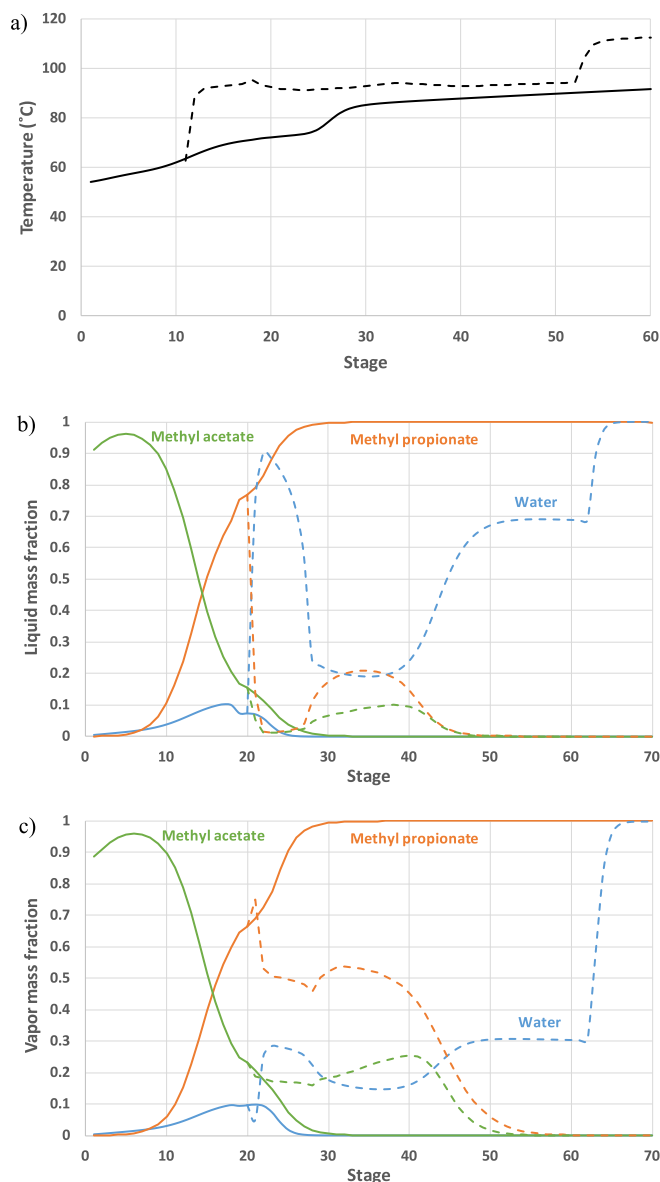
Stream	1	2	3	4	5	6	7
Temperature [°C]	30.0	41.05	54.9	30.0	86.8	56.0	53.8
Pressure [bar]	1.000	0.080	0.150	1.000	1.400	1.600	1.552
Flowrate [kg/h]	20,000	17,400	2,600	538	2,062	927	635
Mass fractions							
Water	0.8776	0.9999	0.0589	0.0000	0.0742	0.0000	0.9990
Propionic acid	0.0721	0.0001	0.5541	0.0000	0.6987	0.0000	0.0008
Acetic acid	0.0234	0.0000	0.1800	0.0000	0.2270	0.0000	0.0000
Succinic acid	0.0269	0.0000	0.2071	1.0000	0.0000	0.0000	0.0002
Methanol	0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	0.0000
Methyl propionate	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methyl acetate	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
DMSO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Stream	8	9	10	11	12	13	14
Temperature [°C]	30.0	54.1	30.0	157.4	60.0	60	56.0
Pressure [bar]	1.552	1.000	1.000	1.232	1.200	1.200	1.600
Flowrate [kg/h]	1,692	663	604	1,162	1,103	< 1	872
Mass fractions							
Water	0.0000	0.0054	0.0000	0.0031	0.0000	0.0000	0.0000
Propionic acid	0.0014	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Acetic acid	0.0002	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Succinic acid	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Methanol	0.0000	0.0838	0.0006	0.0475	0.0000	0.0000	1.0000
Methyl acetate	0.0000	0.9102	0.9988	0.0000	0.0000	0.0000	0.0000
Methyl propionate	0.9984	0.0005	0.0006	0.0000	0.0000	0.0000	0.0000
DMSO	0.0000	0.0000	0.0000	0.9494	1.0000	1.0000	0.0000





**Fig. 7.** R-DWC designed for case 2b (a) and the equivalent sequence of distillation columns (b), numbers in the column indicate column trays, grey patterned part indicates reactive area.

distillation columns (Fig. 7) and designed according to the procedure recommended in the literature [54]. To extend the flexibility of the proposed recovery process design in case 1, the developed DWC design was adjusted for reactive distillation. The feed stages, catalyst holdup, reactive stages, liquid split ratio, reflux ratio and reboiler duties were varied until the desired product specifications were obtained. The top product from column C2, an aqueous mixture of propionic and acetic acid, and methanol are fed to the 28th and 62nd stages of DWC (8th and 42nd stage of  $R-DWC_L$ ), respectively. The esterification reactions are performed in  $R-DWC_L$ , whereby the catalyst is placed from the 11th to 45th stage (from the 31st to 65th stage of  $R-DWC$ ) with 11.7 kg of catalyst per stage.  $R-DWC_R$  performs separation between the formed methyl propionate and methyl acetate. The top vapor from  $R-DWC_L$  goes to  $R-DWC_R$ , while about 44 % of the top liquid from column  $R-DWC_R$  is redirected to  $R-DWC_L$ . The bottom product from  $R-DWC_L$  is a pure water stream that is formed in the esterification reactions. This stream can be used to preheat the  $R-DWC_L$ 's feed stream with a mixture of acids, and then recycled to the fermentation to reduce water requirements. The bottom product from  $R-DWC_R$  is high-purity methyl propionate (99.8 wt %), while the top product is 91.0 wt% methyl acetate. Due to the operation at atmospheric pressure, common trays hosting the catalyst in bags design [55] were assumed for the  $R-DWC$  internals. The total diameter of  $R-DWC$  is 1.3 m, whereby the required tray area for  $R-DWC_R$  is about 2.5 larger compared to  $R-DWC_L$ . Fig. 8 shows the profiles of temperature, liquid and vapor composition along the DWC. The temperature in  $R-DWC_L$  rapidly decreases above the reactive zone and increases below this area, while the temperature in the reactive zone varies less than 5 °C. This is due to the fact that the rectification section above the reactive zone performs the separation of light methyl esters, while the stripping section below the reactive zone performs the



**Fig. 8.** Temperature (a) and composition (b and c) profiles of R-DWC designed for cases 2a and 2b (right side – full line, left side – dotted line).

separation of higher boiling water. Due to the formation in the esterification reactions, the water concentration increases at the bottom of  $R-DWC_L$ , while concentrations of methyl acetate and methyl propionate increase at the top and bottom of  $R-DWC_R$  in both liquid and vapor phases. The concentrations of methyl esters increase in the reactive zone in both phases, while the water concentration decreases. Above the reactive zone, the concentrations of methyl esters increase in the vapor phase, while the water content increases in the liquid phase. The top vapor stream that is being sent to  $R-DWC_R$  contains 75.1 wt% methyl propionate, 18.8 wt% methyl acetate and only 5.0 wt% water.

The obtained methyl acetate with a purity over 90 wt% (which is of commercial quality) can be used in the production of artificial leather, as flavoring, paint remover and solvent for varnishes, resins, oils and nitrocellulose [56]. However, additional steps need to be taken if higher purity methyl acetate is required, due to the formation of a methanol – methyl acetate azeotrope. Extractive distillation with dimethyl sulfoxide (DMSO) is a common way for this separation [57,58]. The addition of DMSO to methanol – methyl acetate mixture changes the relative volatility of the two components and facilitates separation. As shown in Fig. 4, the extractive distillation is performed in column C3, while the

DMSO is recovered for reuse in column C4. The exact flowrate of DMSO (1.4 times higher, on a mole basis, than the flowrate of the methanol – methyl acetate feed stream) was determined by minimizing the total energy requirements for the methyl acetate purification (columns C3 and C4). The top product from column C3 is a high-purity methyl acetate product (99.9 wt%), while the bottom product is a DMSO – methanol mixture that is sent to the solvent recovery column C4. The top product from this column is methanol that can be recycled to R-DWC, while the bottom product is recovered DMSO that can be reused in column C3. Columns C3 and C4 can potentially be merged into one dividing-wall column with a common bottom and divided overhead section. This column would have only one reboiler and two condensers. However, this would lead to somewhat higher operating costs as high-pressure steam would have to be used for heating. Alternatively, using a sequence of two distillation columns allows the usage of low-pressure steam for the first column (reboiler temperature of about 157 °C) and high-pressure steam only for the second column (reboiler temperature of about 193 °C). In addition, an MVR heat pump system can be implemented to R-DWC by using the top compressed vapor to evaporate the bottom liquid from R-DWC<sub>R</sub>. The COP of this heat pump system is about 2.8.

### 3.6. Economic evaluation

A complete economic analysis following the NREL methodology [59] was performed to evaluate and compare the developed processes. The calculated capital expenses (CAPEX) include costs related to equipment purchase and installation, warehouse, site development, additional piping, proratable expenses (fringe benefits, burdens and insurance of the construction contractor), field expenses, home office and construction, working capital, etc. The equipment costs were estimated using the published cost correlation, with the Marshall and Swift cost index of 1,638.2 (end of 2018) [60]. Reliable market data was used to determine the vacuum systems' costs [61]. The operating expenses (OPEX) were determined from costs for utilities, operating labor, maintenance, property insurance, consumables, etc. Even though the costs for utilities strongly depend on the exact plant site, the following values were considered: 60.48 \$/MWh for electricity, 28.01 \$/ MWh for low-pressure steam, 29.59 \$/MWh for medium-pressure steam, 35.59 \$/MWh for high-pressure steam, 1.27 \$/MWh for cooling water and 15.95 \$/MWh for chilled water [51]. The total annual costs (TAC) take into account both CAPEX and OPEX with a payback period of 10 years [59] and were calculated using the following formula:  $TAC = CAPEX/PBP + OPEX$ . The effect of the payback period on TAC is further analyzed to expand the flexibility of the performed economic evaluation. The specific cost data was calculated based on the total products' flowrate, while the average market price (AMP) was determined as a weighted average price of the products.

The calculated economic indicators for both developed processes are presented in Table 5 and Fig. 9. The total costs for installing all process equipment units are 13,887 k\$ for case 1, 14,407 k\$ for case 2a and 16,592 k\$ for case 2b. The largest part of the equipment costs is due to distillation columns (46 – 52 %), followed by the cost for compressors (20 – 32 %) and heat exchangers (22 – 28 %). The CAPEX for the designed processes are 25,334, 26,284 and 30,221 k\$ for cases 1, 2a and 2b, respectively. The costs for equipment make the largest part of CAPEX (~55 %) for all processes. While the CAPEX for case 1 and case 2a are very similar, the CAPEX for case 2b is somewhat higher mainly due to the additional equipment (distillation columns, heat exchangers, pumps).

The OPEX for cases 1, 2a and 2b are 5,274, 7,635 and 7,619 k\$/y, respectively. The contribution of different costs to the total OPEX for case 1 is the following: 57 % for operating labor, 28 % for electricity, 4 % for low-pressure steam, 3 % for high-pressure steam, 1 % for cooling water and 7 % for other costs. The reason for much higher OPEX for cases 2a and 2b is the cost for methanol (2,292 and 2,156 k\$/y, with methanol price of 0.310 \$/kg as reported in the market data from 2023

**Table 5**

Key performance indicators in terms of economics and sustainability.

	Case 1	Case 2a	Case 2b
<b>Economic indicators</b>			
CAPEX [k\$]	25,334	26,284	30,221
OPEX [k\$/y]	5,274	7,635	7,619
OPEX [\$/kg <sub>product</sub> ]	0.269	0.330	0.336
TAC [k\$/y]	7,807	10,264	10,641
TAC [\$/kg <sub>product</sub> ]	0.399	0.444	0.469
<b>Sustainability metrics</b>			
Thermal energy requirements [kW <sub>th</sub> /kg <sub>product</sub> ]	0.644	0.535	0.635
Electrical energy requirements [kW <sub>e</sub> /kg <sub>product</sub> ]	1.214	1.068	1.091
Primary energy requirements [kW <sub>th</sub> /kg <sub>product</sub> ]	3.678	3.206	3.361
Water consumption [m <sup>3</sup> /kg <sub>product</sub> ]	0.417	0.364	0.387
Water loss [m <sup>3</sup> /kg <sub>product</sub> ]	0.029	0.025	0.027
CO <sub>2</sub> emissions [kg <sub>CO2</sub> /kg <sub>product</sub> ]*	0.658 /	0.573 /	0.598 /
	0.105	0.085	0.101
Material intensity [kg <sub>waste</sub> /kg <sub>product</sub> ]	0.000	0.000	0.000
Pollutant emissions [kg <sub>pollutant</sub> /kg <sub>product</sub> ]	0.000	0.000	0.000
Toxic emissions [kg <sub>toxic material</sub> /kg <sub>product</sub> ]	0.000	0.000	0.000

\* Grey / green electricity.

[62]) which makes 30 and 28 % of the total OPEX. The lower cost for methanol in case 2b is due to the fact that recovered methanol from column C4 can be reused in R-DWC. Apart from the methanol cost, costs for operating labor (39 % of OPEX) and electricity (20 % of OPEX) contribute significantly to OPEX for case 2a and case 2b, while costs for low-pressure steam (3 % of OPEX), high-pressure steam (2 – 3 % of OPEX), cooling water (<1 % of OPEX) and other costs (5 – 6 % of OPEX) are much lower. Per mass of recovered product, the OPEX for the designed recovery processes is 0.269, 0.330 and 0.336 \$/kg for cases 1, 2a and 2b, respectively.

Finally, the calculated TAC are 7,807 k\$/y (or 0.399 \$/kg) for case 1, 10,264 k\$/y (or 0.444 \$/kg) for case 2a and 10,641 k\$/y (or 0.469 \$/kg) for case 2b. However, as propionic, acetic and succinic acids are the products in case 1, while methyl esters (methyl propionate, methyl acetate) and succinic acid are recovered in cases 2a and 2b, comparison of the TAC with mass-averaged market prices (AMP) of the product is needed. According to reliable market data [56,63,64], AMP for the products are 1.349, 1.761 and 1.802 \$/kg for cases 1, 2a and 2b, respectively. Thus, the difference between AMP and TAC is much higher in cases 2a and 2b (1.318 and 1.332 \$/kg) as compared to case 1 (0.950 \$/kg). Despite more expensive recovery processes, the higher prices of methyl acetate and methyl propionate (as compared to acetic and propionic acid) lead to a higher margin between AMP and TAC. Additionally, the purification of methyl acetate from 91.0 to 99.9 w% results in a slightly higher margin. To assess profitability, costs of fermentation and prior downstream processing (biomass removal, counterion removal) also need to be considered.

Depending on the business strategy, different payback periods might be required by the investors. Thus, an analysis of the payback period (PBP) effect on the total recovery costs was performed and is presented in Fig. 10. For case 1, TAC are lower than AMP for PBP of 2 years, while shorter PBP leads to a more expensive recovery process compared to the market prices of products. Contrarily, TAC for cases 2a and 2b are lower than AMP even with a PBP of 1 year. Thus, the margin between AMP and TAC strongly depends on the specified PBP.

### 3.7. Sustainability assessment

The environmental impacts of the designed processes were evaluated using key sustainability metrics: energy intensity, water consumption,

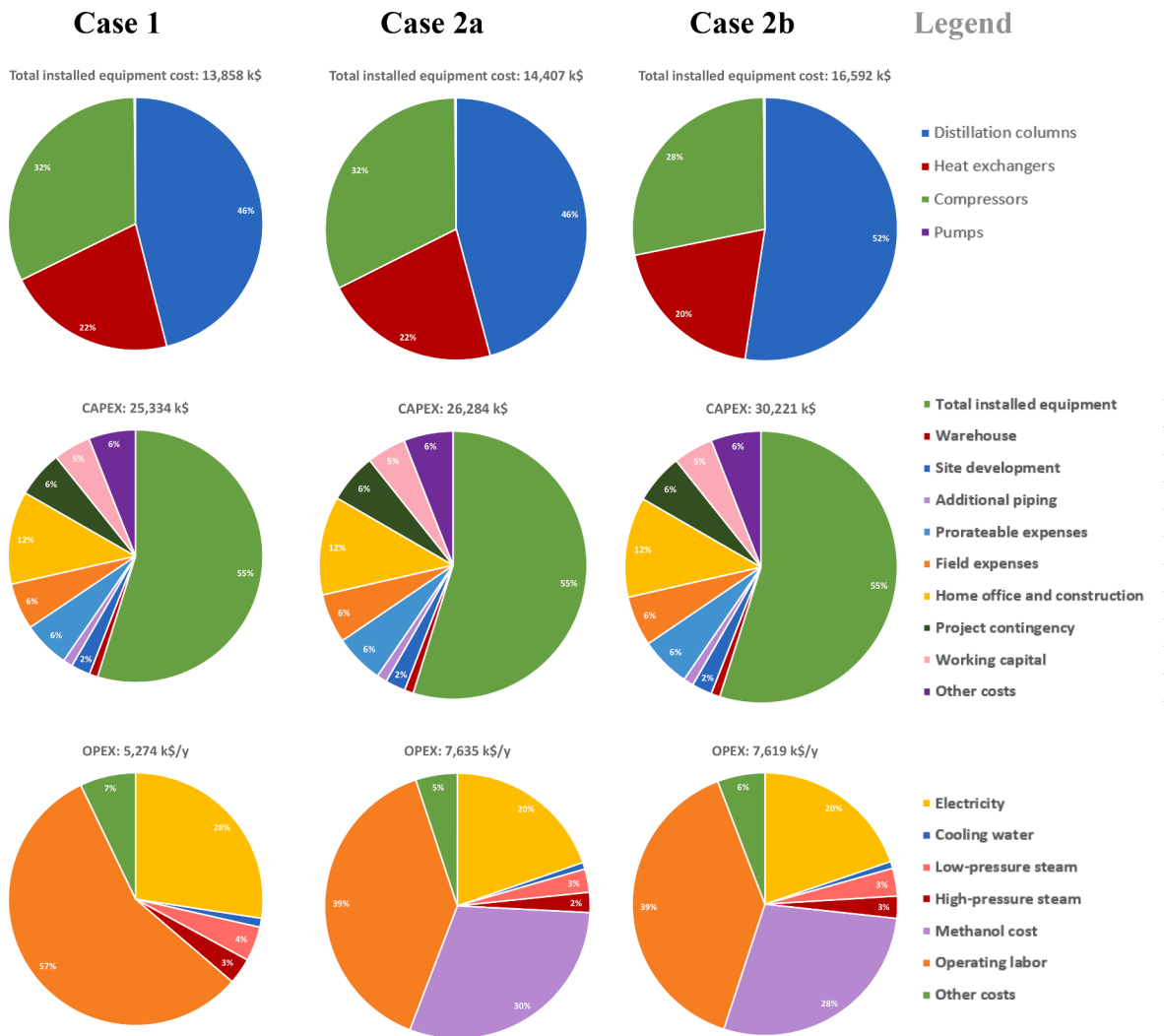


Fig. 9. Comparison of economic indicators for the designed recovery processes.

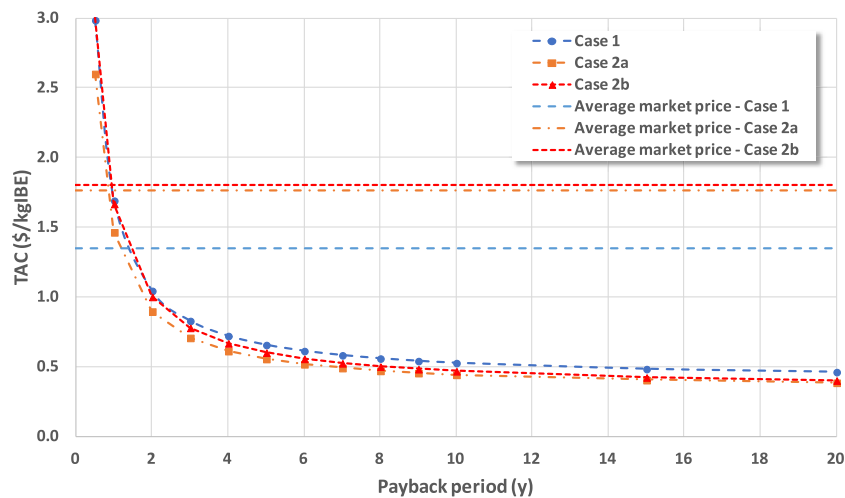


Fig. 10. Influence of the payback period on TAC.

greenhouse gas emissions, material intensity, pollutant and toxic emissions [65]. The lower values of these metrics indicate better process performance in terms of sustainability. The calculated metrics are summarized in Table 5. All the specific values were obtained based on

the total products' flowrate.

- *Energy intensity* is a measure of total energy, thermal and electrical, required to recover one kilogram of product. Thermal energy

requirements present the amount of thermal energy expressed per kilogram of product, while electrical energy requirements are the measure of electrical energy per kilogram of product. Primary energy requirements take into account both thermal and electrical energy, whereby all inefficiencies in power generation and distribution are covered through the electrical to thermal conversion factor – considering a conservative value of 2.5 [42]. Thus, 0.644, 0.535 and 0.635 kW<sub>th</sub>/kg of thermal energy is needed for case 1, case 2a and case 2b, respectively. Additionally, electrical energy requirements for these processes are 1.214, 1.068 and 1.091 kW<sub>e</sub>/kg, respectively. The primary energy requirements are 3.678 kW<sub>th</sub>/kg for case 1, 3.206 kW<sub>th</sub>/kg for case 2a and 3.361 kW<sub>th</sub>/kg for case 2b. Notably, recovering the methyl esters instead of pure acids is less energy-demanding, due to the lower boiling points and larger relative volatility of esters. Given that the total energy requirements for the propionic acid production from glycerol were assumed to be 120.4 – 138.9 MJ/kg, whereby the energy for the fermentation makes only a minor contribution [66], the process proposed in this study is a major step towards more competitive production of propionic acid by industrial fermentation. Additionally, since the designed preconcentration and final purification steps can be expected to be more energy-demanding compared to the prior operations (e.g. counterion removal by BME [19]), minimizing energy requirements for these steps is of crucial importance for the viability of the overall production process.

- **Water consumption** indicates the amount of water needed per kilogram of product, accounting for cooling water with an estimated loss of 7 % [65] and inefficiencies in the steam cycle with an estimated condensate recovery of 70 % [67]. Due to thermal energy requirements, the water consumption and loss are the highest for case 1 (0.417 and 0.029 m<sup>3</sup><sub>w</sub>/kg, respectively). Water consumption and loss are somewhat lower for case 2a (0.364 and 0.025 m<sup>3</sup><sub>w</sub>/kg) and case 2b (0.387 and 0.027 m<sup>3</sup><sub>w</sub>/kg).
- **Greenhouse gas emissions** indicate the amount of carbon-dioxide (CO<sub>2</sub>) emitted per kilogram of product. The recommendations from the published literature, based on thermal and electrical energy usage, were used to evaluate these emissions [60,68]. Furthermore, differentiation was made between the usage of grey (from fossil fuels) and green electricity (from renewable sources). CO<sub>2</sub> emissions are the highest in case 1 due to the highest energy requirements (0.658 and 0.105 kg<sub>CO2</sub>/kg for grey and green electricity usage, respectively). Recovering methyl esters of propionic and acetic acids leads to lower CO<sub>2</sub> emissions both in case of grey (0.573 and 0.598 kg<sub>CO2</sub>/kg for cases 2a and 2b) and green electricity (0.085 and 0.101 kg<sub>CO2</sub>/kg for cases 2a and 2b) usage. Available data on the carbon footprint of propionic acid (4.40 kg<sub>CO2</sub>/kg [69]), acetic acid (1.67 kg<sub>CO2</sub>/kg [70]) and ethyl acetate (1.67 kg<sub>CO2</sub>/kg [71]), may indicate the total greenhouse emissions for these products.
- **Material intensity** indicates the amount of total waste per kilogram of recovered product. In case 1, besides propionic, acetic and succinic acids products, the outlet stream is pure water from the top of column C1 which can be used elsewhere. Besides methyl propionate, methyl acetate and succinic acid products, the outlet streams in cases 2a and 2b are the top aqueous product of column C1 and the bottom water product from DWC<sub>L</sub>. These streams are also pure enough for use elsewhere. Thus, as there are no waste outlet streams in any of the recovery processes, the material intensity metrics are equal to zero.
- **Pollutant and toxic materials** are measure of formed pollutants and toxic materials per kilogram of product. There is no formation of pollutants or toxic materials in the developed processes. Also, the DMSO used for the extractive distillation in case 2b can be completely recovered in the solvent recovery column C4. Thus, the values of these metrics are equal to zero for all recovery processes.

The comparison of these metrics between cases should only serve as

an indication due to the different natures of the recovered products. Nonetheless, since all of the sustainability metrics are reflected in the total recovery costs (e.g. costs of electrical energy, heating and cooling utilities, etc.), the most reliable comparison is the one between TAC and AMP, as discussed in section 3.6.

### 3.8. Sensitivity analysis

Since concentrations of carboxylic acids in the feed to the designed process can vary depending on the prior processes, these concentrations were varied in a sensitivity analysis while the production capacity remained constant. Fig. 11 presents the results of the sensitivity analysis. Lower concentrations of carboxylic acid in the feed stream lead to higher TAC, energy requirements, CO<sub>2</sub> emissions in case of grey electricity usage, and water consumption. Especially concentrations lower than 10 wt% (>90 wt% water) result in poorer performance. To the best of our knowledge, the costs of the propionic acid recovery from a highly dilute aqueous stream have not been reported. Nonetheless, the reported energy requirements for the recovery of high-purity acetic and lactic acid from dilute streams by liquid–liquid extraction might give some indication (~1.7 – 6.9 kWh/kg and ~11.1 – 55.6 kWh/kg for 10 and 1 wt% of acetic acid in the feed, or ~15.3 kWh/kg and ~138.9 kWh/kg for 10 and 1 wt% of lactic acid in the feed) [72]. Thus, the proposed processes for recovery of carboxylic acids from aqueous solution are competitive even with more dilute streams.

## 4. Conclusion

The proposed intensified downstream process was demonstrated to successfully recover high-purity propionic, acetic and succinic acid or methyl propionate, methyl acetate and succinic acid, while simultaneously allowing recycle of water to the fermentation. The initial distillation step to remove most of the water from the fermentation broth should be performed under reduced pressure to facilitate the separation, decrease product loss and increase efficiency. Additionally, a mechanical vapor recompression heat pump system can be applied to reduce energy requirements by more than 13 times. Moreover, separating succinic acid as the next step in the process allows flexibility in further recovery of propionic and acetic acid or of their methyl esters (as higher added value products). At the heart of the process is a highly integrated heat pump-assisted dividing-wall column which allows the separation of the high-purity propionic and acetic acid from aqueous solution, or it can serve as a reactive dividing-wall column (same design but adapted to host a solid catalyst on the reactive stages) that can efficiently perform the esterification with methanol and the separation of the methyl esters and water by-product. Due to the formation of an azeotrope (methanol – methyl acetate), an additional extractive distillation step might be needed if methyl acetate purities higher than 91 wt% are required. Even though recovering both carboxylic acids and their esters are promising options for recovery from dilute aqueous solution on a large-scale (capacity of ~20 ktonne/y), the esterification of propionic and acetic acid is a superior option due to the higher margin that could be obtained (1.318 and 1.332 \$/kg as compared to 0.950 \$/kg for a payback period of 10 years). However, it should be noted that the costs of fermentation, filtration / centrifugation and counterion removal will further decrease the above-mentioned margins. As these costs are the same for all cases discussed in this work, recovering the esters (methyl propionate and methyl acetate) will still be more cost-effective than recovering the acid (propionic and acetic acids).

Overall, this original research contributes to advancements in the fermentative production of carboxylic acids by proposing an efficient process design that can be adapted to the recovery of propionic and acetic acids or esters (methyl propionate, methyl acetate) and succinic acid, depending on the market demand. Future research exploring the presence of other by-products or impurities would be beneficial to expand the versatility of the proposed downstream process design.



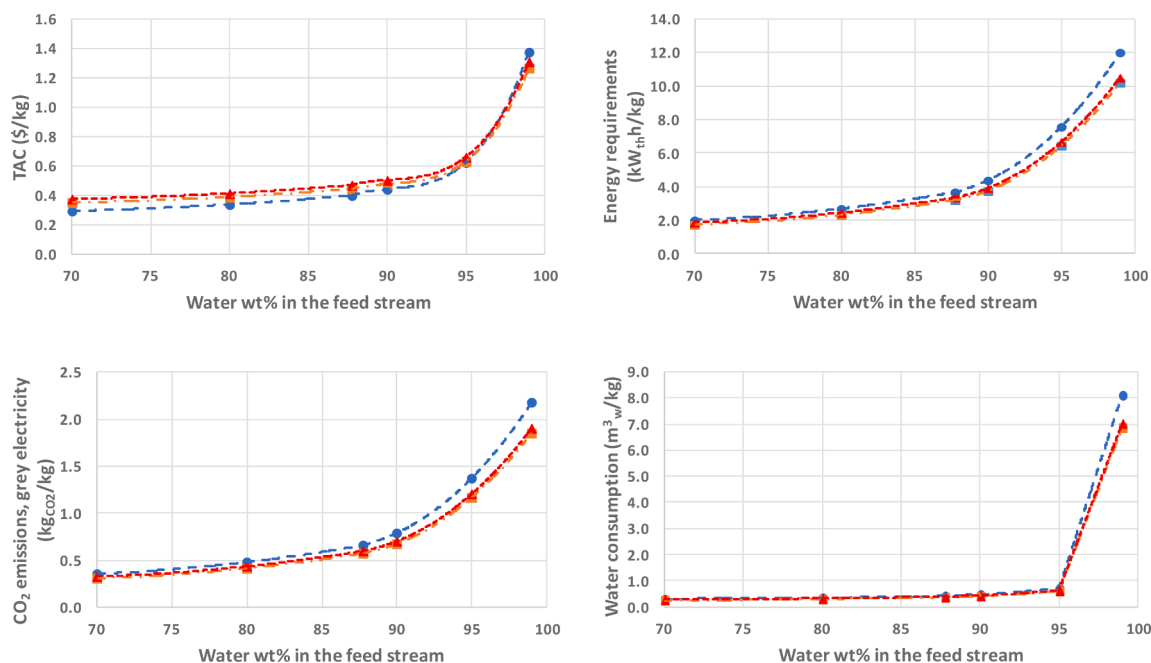


Fig. 11. Sensitivity analysis results: influence of the water content in the feed stream on the main economic and sustainability indicators: Case 1—●—, Case 2a—■— and Case 2b—▲—.

#### CRedit authorship contribution statement

**Tamara Janković:** Writing – review & editing, Writing – original draft, Visualization, Validation, Software, Methodology, Investigation, Data curation, Conceptualization. **Adrie J. J. Straathof:** Writing – review & editing, Validation, Supervision, Methodology, Formal analysis, Conceptualization. **Anton A. Kiss:** .

#### Declaration of competing interest

The authors declare the following financial interests/personal relationships which may be considered as potential competing interests: [Prof. Anton A. Kiss was a Guest Editor for this Special issue, but he was not involved in the handling or the peer review of the manuscript.].

#### Appendix A. Supplementary material

Supplementary material to this article can be found online at <https://doi.org/10.1016/j.seppur.2024.127832>.

#### References

- [1] F.M. Liew, M.E. Martin, R.C. Tappel, B.D. Heijstra, C. Mihalcea, M. Köpke, Gas fermentation - a flexible platform for commercial scale production of low-carbon-fuels and chemicals from waste and renewable feedstocks, *Front. Microbiol.* 7 (2016) 694, <https://doi.org/10.3389/fmicb.2016.00694>.
- [2] Z. Usmani, M. Sharma, A.K. Awasthi, T. Lukk, M.G. Tuohy, L. Gong, P. Nguyen-Tri, A.D. Goddard, R.M. Bill, S.C. Nayak, V.K. Gupta, Lignocellulosic biorefineries: the current state of challenges and strategies for efficient commercialization, *Renew. Sustain. Energy Rev.* 148 (2021) 111258, <https://doi.org/10.1016/j.rser.2021.111258>.
- [3] C.S. López-Garzón, A.J.J. Straathof, Recovery of carboxylic acids produced by fermentation, *Biotechnol. Adv.* 32 (2014) 873–904, <https://doi.org/10.1016/j.biotechadv.2014.04.002>.
- [4] R.A. Gonzalez-Garcia, T. McCubbin, L. Navone, C. Stowers, L.K. Nielsen, E. Marcellin, Microbial propionic acid production, *Fermentation* 3 (2017) 1–20, <https://doi.org/10.3390/fermentation3020021>.
- [5] U.-R. Samel, W. Kohler, A. Garner, Keuser, Propionic acid and derivatives, in: *Ullmann's Encyclopedia of Industrial Chemistry*, 2011: pp. 295–309.
- [6] Z. Wang, J. Sun, A. Zhang, S. Yang Propionic Acid Fermentation in: *Bioprocessing Technologies in Biorefinery for Sustainable Production of Fuels, Chemicals, and Polymers*, 2013: pp. 331–350.
- [7] Methyl propionate, PubChem. <https://pubchem.ncbi.nlm.nih.gov/compound/Methyl-propionate>.
- [8] J.P.C. Pereira, L.A.M. van der Wielen, A.J.J. Straathof, Perspectives for the microbial production of methyl propionate integrated with product recovery, *Bioresour. Technol.* 256 (2018) 187–194, <https://doi.org/10.1016/j.biortech.2018.01.118>.
- [9] Global Methyl Acetate Market Outlook, Expert Market Research. (2024). <https://www.expertmarketresearch.com/reports/methyl-acetate-market>.
- [10] C.C. Stowers, B.A. Rodriguez, B.M. Cox, Fermentation based hydrolyzed corn and/or sugar can mash to produce propionic acid, US 2016/017384 A1, 2016.
- [11] S.-T. Yang, E. Ammar, C.C. Stowers, B.A. Rodriguez, Process for producing n-propanol and propionic acid using metabolically engineered *Propionibacteria*, US 10,150,951 B2, 2018.
- [12] X. Feng, F. Chen, H. Xu, B. Wu, H. Li, S. Li, P. Ouyang, Green and economical production of propionic acid by *Propionibacterium freudenreichii* CCTCC M207015 in plant fibrous-bed bioreactor, *Bioresour. Technol.* 102 (2011) 6141–6146, <https://doi.org/10.1016/j.biortech.2011.02.087>.
- [13] Z. Liu, C. Ma, C. Gao, P. Xu, Efficient utilization of hemicellulose hydrolysate for propionic acid production using *Propionibacterium acidipropionici*, *Bioresour. Technol.* 114 (2012) 711–714, <https://doi.org/10.1016/j.biortech.2012.02.118>.
- [14] S. Suwannakham, S.T. Yang, Enhanced propionic acid fermentation by *Propionibacterium acidipropionici* mutant obtained by adaptation in a fibrous-bed bioreactor, *Biotechnol. Bioeng.* 91 (2005) 325–337, <https://doi.org/10.1002/bit.20473>.
- [15] A. Zhang, S.T. Yang, Engineering *Propionibacterium acidipropionici* for enhanced propionic acid tolerance and fermentation, *Biotechnol. Bioeng.* 104 (2009) 766–773, <https://doi.org/10.1002/bit.22437>.
- [16] B.A. Rodriguez, C.C. Stowers, V. Pham, B.M. Cox, The production of propionic acid, propanol and propylene via sugar fermentation: an industrial perspective on the progress, technical challenges and future outlook, *Green Chem.* 16 (2014) 1066–1076, <https://doi.org/10.1039/c3gc42000k>.
- [17] J.L. Rehbergert, B.A. Glatz, Response of cultures of *propionibacterium* to acid and low pH : tolerance and inhibition, *J. Food Prot.* 61 (1998) 211–216, <https://doi.org/10.4315/0362-028X-61.2.211>.
- [18] H.A. Seong, J.S. Lee, S.Y. Yoon, W.Y. Song, S.J. Shin, Fermentation characteristics of acid hydrolysates by different neutralizing agents, *Int. J. Hydrogen Energy* 41 (2016) 16365–16372, <https://doi.org/10.1016/j.ijhydene.2016.05.003>.
- [19] É. Hülber-Beyer, K. Bélafi-Bakó, N. Nemestóthy, Low-waste fermentation-derived organic acid production by bipolar membrane electrodialysis—an overview, *Chem. Pap.* 75 (2021) 5223–5234, <https://doi.org/10.1007/s11696-021-01720-w>.
- [20] P. Boyaval, C. Corre, Production of propionic acid, *Lait* 75 (1995) 453–461, [https://doi.org/10.1016/0023-7302\(96\)80128-X](https://doi.org/10.1016/0023-7302(96)80128-X).
- [21] D. Arslan, Y. Zhang, K.J.J. Steinbusch, L. Diels, H.V.M. Hamelers, C.J.N. Buisman, H. De Wever, In-situ carboxylate recovery and simultaneous pH control with tailor-configured bipolar membrane electrodialysis during continuous mixed culture fermentation, *Sep. Purif. Technol.* 175 (2017) 27–35, <https://doi.org/10.1016/j.seppur.2016.11.032>.
- [22] M. Szczygielda, J. Antczak, K. Prochaska, Separation and concentration of succinic acid from post-fermentation broth by bipolar membrane electrodialysis (EDBM), *Sep. Purif. Technol.* 181 (2017) 53–59, <https://doi.org/10.1016/j.seppur.2017.03.018>.

- [23] V. Inyang, D. Lokhat, Propionic acid recovery from dilute aqueous solution by emulsion liquid membrane (ELM) technique: optimization using response surface methodology (RSM) and artificial neural network (ANN) experimental design, *Sep. Sci. Technol.* 57 (2022) 284–300, <https://doi.org/10.1080/01496395.2021.1890774>.
- [24] A. Kaur, D.K. Vohra, Study of bulk liquid membrane as a separation technique to recover acetic and propionic acids from dilute solutions, *Indian J. Chem. Technol.* 17 (2010) 133–138.
- [25] Y.S. Aşçı, I. İnci, Extraction equilibria of propionic acid from aqueous solutions by Amberlite LA-2 in diluent solvents, *Chem. Eng. J.* 155 (2009) 784–788, <https://doi.org/10.1016/j.cej.2009.09.024>.
- [26] A. Keshav, S. Chand, K.L. Wasewar, Recovery of propionic acid from aqueous phase by reactive extraction using quaternary amine (Aliquat 336) in various diluents, *Chem. Eng. J.* 152 (2009) 95–102, <https://doi.org/10.1016/j.cej.2009.03.037>.
- [27] Q.-Z. Li, X.-J. Feng, H.-B. Zhang, H.-Z. Liu, M. Xian, C. Sun, J.-M. Wang, X.-L. Jiang, Recovery processes of organic acids from fermentation broths in the biomass-based industry, *J. Microbiol. Biotechnol.* 26 (2016) 1–8, <https://doi.org/10.4014/jmb.1505.05049>.
- [28] S.-T. Hsu, S.-T. Yang, Propionic acid fermentation of lactose, *Biotechnol. Bioeng.* 38 (1991) 571–578.
- [29] Z. Wang, Y. Jin, S. Yang, High Cell Density Propionic Acid Fermentation with an Acid Tolerant Strain of *Propionibacterium Acidipropionici* 112 (2015) 502–511, <https://doi.org/10.1002/bit.25466>.
- [30] C.C. Stowers, B.M. Cox, B.A. Rodriguez, Development of an industrializable fermentation process for propionic acid production, *J. Ind. Microbiol. Biotechnol.* 41 (2014) 837–852, <https://doi.org/10.1007/s10295-014-1423-6>.
- [31] J. Gmehling, J. Menke, J. Krafczyk, K. Fischer, *Azeotropic Data*, Wiley-VCH, 2004.
- [32] Aspen Physical Property System, Aspen Technology, Bedford, 2020.
- [33] B.P. Dakshinamurthy, G.J. Rao, C.V. Rao, Vapour-liquid equilibria in the system water-propionic acid, *J. Appl. Chem.* 11 (1961) 226–228.
- [34] C. Yang, X. Yin, S. Ma, Organic salt effect of tetramethylammonium bicarbonate on the vapor-liquid equilibrium of the dimethyl carbonate + methanol system, *J. Chem. Eng. Data* 57 (2012) 66–71, <https://doi.org/10.1021/jc200697m>.
- [35] V.H. Alvarez, S. Mattedi, M. Iglesias, R. Gonzalez-Olmos, J.M. Resa, Phase equilibria of binary mixtures containing methyl acetate, water, methanol or ethanol at 101.3 kPa, *Phys. Chem. Liq.* 49 (2011) 52–71, <https://doi.org/10.1080/00319100903012403>.
- [36] D. Gao, D. Zhu, H. Sun, L. Zhang, H. Chen, J. Si, Isobaric vapor - liquid equilibria for binary and ternary mixtures of methanol ethanoic acid, and propanoic acid, *J. Chem. Eng. Data* 55 (2010) 4002–4009, <https://doi.org/10.1021/jc1000473>.
- [37] S. Bernatov, K. Aim, I. Wichterle, Isothermal vapour - liquid equilibrium with chemical reaction in the quaternary water + methanol + acetic acid + methyl acetate system, and in five binary subsystems, *Fluid Phase Equilib.* 247 (2006) 96–101, <https://doi.org/10.1016/j.fluid.2006.06.005>.
- [38] C.T. Hsieh, W.Y. Ji, H. mu Lin, M.J. Lee, Multiphase equilibria for mixtures containing water, acetic acid, propionic acid, methyl acetate and methyl propionate, *Fluid Phase Equilib.* 271 (2008) 69–75, <https://doi.org/10.1016/j.fluid.2008.07.004>.
- [39] R.H. Perry D.W. Green Distillation, in: *Perry's Chemical Engineering Handbook*, McGraw-Hill 1997 13.
- [40] Sulzer, Structured packings, (2023). <https://www.sulzer.com/en/products/separation-technology/structured-packings>.
- [41] A.A. Kiss, C.A. Infante Ferreira, Mechanically Driven Heat Pumps, in: *Heat Pumps in Chemical Process Industry*, CRC Press, Boca Raton, 2016: pp. 189–251.
- [42] BP, Approximate Conversion Factors, Statistical Review of World Energy. (2021). <https://www.bp.com/en/global/corporate/energy-economics/statistical-review-of-world-energy/using-the-review/definitions-and-explanatory-notes.html>.
- [43] SPRSUN, What Is a Good COP for an Air to Water Heat Pump?, (2021). <https://sprsunheatpump.com/what-is-a-good-cop-for-an-air-to-water-heat-pump.html>.
- [44] I. Pazmiño-Mayorga, M. Jobson, A.A. Kiss, Conceptual design of a dual reactive dividing wall column for downstream processing of lactic acid, *Chem. Eng. Process. - Process Intensif.* 164 (2021), <https://doi.org/10.1016/j.cep.2021.108402>.
- [45] F.J. Caires, L.S. Lima, C.T. Carvalho, M. Ionashiro, Thermal behaviour of succinic acid, sodium succinate and its compounds with some bivalent transition metal ions, *Thermochim Acta* 500 (2010) 6–12, <https://doi.org/10.1016/j.tca.2009.11.015>.
- [46] E. Volker, How to design and optimise sieve trays, *WelChem*. (2020).
- [47] T.M. DuPont, T.M. Amberlyst, *Polymeric Catalysts* (2021).
- [48] Y.T. Tsai, H. mu Lin, M.J. Lee, Kinetics behavior of esterification of acetic acid with methanol over Amberlyst 36, *Chem. Eng. J.* 171 (2011) 1367–1372, <https://doi.org/10.1016/j.cej.2011.05.049>.
- [49] Y.T. Tsai, H. mu Lin, M.J. Lee, Kinetics of catalytic esterification of propionic acid with methanol over Amberlyst 36, *Ind. Eng. Chem. Res.* 50 (2011) 1171–1176, <https://doi.org/10.1016/j.cej.2011.05.049>.
- [50] A.A. Kiss, J.J. Prag, C.J.G. van Strien, Reactive dividing-wall columns-how to get more with less resources? *Chem. Eng. Commun.* 196 (2009) 1366–1374, <https://doi.org/10.1080/00986440902935507>.
- [51] A.A. Kiss, Design, control and economics of distillation, in: *Advanced Distillation Technologies: Design, Control and Application*, Wiley, 2013: pp. 37–66.
- [52] A.A. Kiss, C.S. Bildea, A control perspective on process intensification in dividing-wall columns, *Chem. Eng. Process.* 50 (2011) 281–292, <https://doi.org/10.1016/j.cep.2011.01.011>.
- [53] Ö. Yildirim, A.A. Kiss, E.Y. Kenig, Dividing wall columns in chemical process industry: a review on current activities, *Sep. Purif. Technol.* 80 (2011) 403–417, <https://doi.org/10.1016/j.seppur.2011.05.009>.
- [54] W.L. Luyben, C.C. Yu, *Reactive Distillation Design and Control*, John Wiley & Sons, Hoboken, 2008.
- [55] Reactive Distillation Column, Maleta Cyclic Distillation. <https://www.maletacd.com/index.php/distillation-column/reactive-distillation-column>.
- [56] Methyl acetate 90%, VICHEMIC INDUSTRIES. <https://vichemic.pl/en/produkt/methyl-acetate-90-79-20-9/>.
- [57] E. Graciová, B. Sulgan, S. Barabas, P. Steltenpohl, Methyl acetate-methanol mixture separation by extractive distillation: economic aspects, *Front. Chem. Sci. Eng.* 12 (2018) 670–682, <https://doi.org/10.1007/s11705-018-1769-9>.
- [58] L. Berg, A.I. Yeh, The separation of methyl acetate from methanol by extractive distillation, *Chem. Eng. Commun.* 30 (1984) 113–117, <https://doi.org/10.1080/00986448408911119>.
- [59] D. Humbird R. Davis L. Tao C. Kinchin D. Hsu A. Aden P. Schoen J. Lukas B. Olthof M. Worley D. Sexton D. Dudgeon Process Design and Economics for Biochemical Conversion of Lignocellulosic Biomass to Ethanol 2011.
- [60] J. Mantingh, A.A. Kiss, Enhanced process for energy efficient extraction of 1,3-butadiene from a crude C4 cut, *Sep. Purif. Technol.* 267 (2021) 118656, <https://doi.org/10.1016/j.seppur.2021.118656>.
- [61] Roots vacuum pumps, Marpa Vacuum, [marpavacuum.com/en/roots-vacuum-pumps](http://marpavacuum.com/en/roots-vacuum-pumps).
- [62] Methanol Price Trend and Forecast, ChemAnalyst. (2023). <https://www.chemanalyst.com/Pricing-data/methanol-1>.
- [63] Pricing data, ChemAnalyst. <https://www.chemanalyst.com/Pricing/Pricingoverview>.
- [64] Methyl acetate 99.5%, VICHEMIC INDUSTRIES. <https://vichemic.pl/en/produkt/methyl-acetate-99-5-ar-79-20-9/>.
- [65] J. Schwarz, B. Beloff, E. Beaver, Use sustainability metrics to guide decision-making, *Chem. Eng. Prog.* 98 (2002) 58–63.
- [66] J.A. Posada, C.A. Cardona, Propionic acid production from raw glycerol using commercial and engineered strains, *Ind. Eng. Chem. Res.* 51 (2012) 2354–2361, <https://doi.org/10.1021/ie201300d>.
- [67] N.P. Lieberman, E.T. Lieberman, Steam Generation, in: *A Working Guide to Process Equipment*, McGrawHill, 2022: pp. 261–276.
- [68] A.A. Kiss, D.-J.-P.-C. Suszwalak, Innovative dimethyl ether synthesis in a reactive dividing-wall column, *Comput. Chem. Eng.* 38 (2012) 74–81, <https://doi.org/10.1016/j.compchemeng.2011.11.012>.
- [69] Propionic acid, fossil based, CarbonCloud. <https://apps.carboncloud.com/climatehub/product-reports/id/143249778560>.
- [70] Acetic acid, CarbonCloud. <https://apps.carboncloud.com/climatehub/product-reports/id/67148333700>.
- [71] Ethyl acetate, CarbonCloud. <https://apps.carboncloud.com/climatehub/product-reports/id/3223056066806>.
- [72] E. Reyhanitash, T. Brouwer, S.R.A. Kersten, A.G.J. van der Ham, B. Schuur, Liquid-liquid extraction-based process concepts for recovery of carboxylic acids from aqueous streams evaluated for dilute streams, *Chem. Eng. Res. Des.* 137 (2018) 510–533, <https://doi.org/10.1016/j.cherd.2018.07.038>.