Original Research

Electrochemically Induced Crystallization as a Sustainable Method for Product Recovery of Building Block Chemicals: Techno-Economic Evaluation of Fumaric Acid Separation

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Abstract

Carboxylic acids are key platform chemicals for use as biobased alternatives for fossil-based applications. State-of-the-art fermentations of carboxylic acids at neutral pH with downstream product recovery by pH-shift crystallization are not sustainable due to the accompanied production of waste salts. To earn the label “sustainable,” salts should be converted into the original base and acid required for fermentation and product recovery. This paper shows that electrochemically induced crystallization (EIC) integrated with fermentation reduces both capital and operating expenditures (CAPEX and OPEX, respectively). EIC is a novel process that utilizes electrolysis of water to induce localized pH-shift crystallization of the carboxylic acid, while neutral bulk conditions are maintained.

In the current study, a techno-economic evaluation of the fermentative production of fumaric acid integrated with product removal by crystallization was performed. Three cases were evaluated, two based on classical pH-shift crystallization and one based on fermentation integrated with EIC. The first so-called base case produces a solid waste, while the second employs electrodialysis at the end of downstream processing to regenerate the waste salt and re-use the base and acid. With EIC, electrochemistry obviates the use of chemicals and hence solid waste cannot be formed. Cost estimates for EIC unit operation were based on the mass and energy balances of laboratory-scale experimental work and the resultant conceptual design. Assuming return on investment (ROI) of 15%, the first base case has the lowest required sales price of €2.2/kg fumaric acid (USD2.84/kg fumaric acid). EIC eliminates the necessity of water removal, hence reduces CAPEX and OPEX by 35% and 19%, respectively. Sensitivity analysis reveals that with an augmenting penalty on solid waste disposal (greater than €128/ton; USD164/ton), EIC is the most economical option. This paper demonstrates that the integration of fermentation with product recovery improves

the overall process economics and, by applying EIC, a more sustainable process is obtained.

Introduction

Biobased processes play a key role in the transition from a fossil resource-based industry toward a sustainable industry that converts renewable feedstocks into energy and materials. According to Sauer et al., the purpose of such processes is to provide building block chemicals, rather than attempt to produce every required compound.¹ These building blocks can be used as feedstock for a sustainable chemical industry that converts the substrate into high-value chemicals and materials.

Nine out of 12 building block chemicals identified by the US Department of Energy as top value-added chemical targets to produce from biomass are organic acids.² These are used in polymers and resins, as food and beverage additives, and as therapeutics, projecting the importance of market volume of carboxylic acids and amino acids. The production of bulk quantities of such building block chemicals requires efficient processes with high volumetric productivities. Fermentations in general, however, are often limited by inhibition and/or toxicity at high product concentrations.³ Hence, the product should be removed as soon as it is formed to prevent accumulation of the fermentation product. Low product concentrations can, in principle, be achieved by applying *in situ* product removal (ISPR) techniques.^{4–8}

State-of-the-art fermentative production of carboxylic acids is carried out at neutral pH.^{9,10} Consequently, the less toxic carboxylate is present in the fermentation broth, which provides a product sink alleviating the direct toxic effects of the carboxylic acid. However, high product concentrations may inhibit enzymatic conversions inside the whole-cell catalyst and cause osmotic pressure, both leading to decreased productivity. Efficient product removal of the charged carboxylate during fermentation is thus very important.

Conventional fermentations that are conducted at neutral pH require neutralization to counterbalance acidification of the broth by the production of carboxylic acids. Often, NaOH is used to provide the necessary base (OH[−]). The proton from the carboxylic acid reacts with OH[−] to form water, leaving the cation (e.g., Na⁺) and the carboxylate anion in solution. To obtain the carboxylic acid, pH-shift crystallization with H₂SO₄ is applied. The crystalline product is recovered using a solid-liquid separation (either filtration or centrifugation),

while the mother liquor contains Na_2SO_4 , a by-product of neutralization and pH-shift crystallization. Na_2SO_4 is often considered to be without value due to the large quantities in which it is produced as a by-product in various other processes. Consequently, conventional fermentation processes that produce carboxylic acids are not very sustainable, because 1 mole of waste is produced per mole of product. This corresponds to the production of 1 ton of waste per ton of product because of the similar molecular weight for the waste salt and the product. When a plant is close to the sea, the mother liquor can be discarded as wastewater, while in most other cases the salt needs to be removed from the mother liquor before disposal.

Nowadays, a penalty has to be paid to discharge the waste. From an economical point of view, this adds to the operational costs, although it can often be overlooked when compared to substrate prices or utility costs. However, it can reasonably be expected that in the near future processes that generate large quantities of waste will be restricted. Such processes need to be adapted to prevent the accompanied production of waste. In the case of carboxylic acids, the waste salt can, in principle, be regenerated by electrodialysis. The salt breaks up into an acid (e.g., H_2SO_4) and a base (e.g., NaOH), which can be recycled to the crystallizer and fermenter, respectively.

Electrodialysis can be applied not only to regenerate the waste salt, but also for product removal, thus replacing the use of insoluble salt complexes.^{11,12} However, the use of electrodialysis for product removal of slightly soluble carboxylic acids has been limited due to undesired crystallization inside the module. Rather complex solutions with elevated temperatures or mixed solvents have been used to overcome this constraint.^{13,14} Recently, we proposed the use of crystallization inside the electrodialysis module. Using electrochemically induced crystallization, product removal of slightly soluble carboxylic acids (cinnamic, fumaric, and p-hydroxybenzoic acid) has been demonstrated.¹⁵ In this process, pH-shift crystallization is achieved through electrolysis of water instead of the use of chemicals. Consequently, by applying EIC, carboxylic acids are removed from the neutral fermentation broth, while the production of an environmentally unfriendly waste salt is avoided.

Our previous paper demonstrated the potential of EIC for product removal of carboxylates. This paper presents a techno-economical evaluation of EIC and touches upon some sustainability aspects to underline the applicability of EIC for product removal in biochemical processes. Here the performance of EIC is compared with classical pH-shift crystallization with or without waste salt regeneration. For this purpose, three conceptual processes were designed for the three different cases following procedures described previously.^{16–19} Fumaric acid was chosen as a model compound because high production concentrations can already be obtained.^{20–22}

Process Design

The conceptual process design of EIC was based on crystallization experiments to determine the performance (e.g., crystallization efficiency and operating costs) of EIC. The designs of the processes in which classical pH-shift crystallization was applied were based on well-known conventional unit operations. The fermentative

production of model compound fumaric acid was considered identical for the three conceptual processes, and downstream processing steps of the three different designs are compared. The fermentation is taken into account for realistic cost price estimation.

BASIC ASSUMPTIONS

The fumaric acid production facility is located in the Netherlands, where discarding salt waste streams in the sea is not a possibility. Consequently, salt needs to be removed from the wastewater. The plant life for the design cases is assumed to be 10 years due to the use of unconventional equipment. In our calculations, an operational time of 7,920 hours per year is assumed.

In the process designs, fumaric acid is produced by the pellet form of *Rhizopus oryzae* in a stirred tank reactor in continuous mode. The metabolic pathway and kinetics of the reaction are described by Engel et al.²³ The design challenge is to produce 25 ktons per year of 99.7% pure fumaric acid. In this design, the feed material is glucose of satisfactory quality; hence, no additional pre-treatment is required. It should be noted that the fermentation medium consists of a number of different compounds in relatively low quantities. For simplicity, these components are neglected.

BATTERY LIMIT

All the involved unit operations are within the battery limit and concern the complete process from the preparation of the fermentation medium until the recovery of the product and the regeneration of auxiliary phases. The treatment of (biomass) waste and purge streams is outside the battery limit, but a penalty for the production of such waste is taken into account. Furthermore, it is assumed that utilities such as electricity, steam, and cooling water are present at the plant location. So the generation of utilities is outside the battery limit, while the associated costs are within the battery limit.

CONCEPTUAL DESIGN OF THE FERMENTATION PROCESS

The conceptual design of the fermentation process for the production of fumaric acid is derived from a Delft University of Technology (TUDelft) report.²⁴ It includes a growth fermenter and two production fermenters. The growth fermenter is an airlift-loop reactor operated in batch mode. The nutrient requirements to grow large but permeable and firm pellets consist of a carbon source (glucose), a nitrogen source (ammonia), inorganic salts [K_2SO_4 , MgSO_4 , $\text{Ca}(\text{H}_2\text{PO}_4)_2 \cdot \text{H}_2\text{O}$], vitamins, and some metal ions.²⁵ To maximize biomass production, an excess amount of feed and nutrients, as well as air, are fed to the reactor. The incubation temperature and the pH are set at 30°C and 3.36, respectively. After one batch cycle (158.5 hours), the biomass is fed into a holding tank. It is important to provide sufficient substrate to the holding tank for maintenance of the biomass.

The biomass contained within the biomass maintenance tank is sent to the production fermenters as a biocatalyst. Along with the biomass, the carbon source and neutralizing agent (NaOH) are added to the production fermenters. The production phase is nitrogen-limited to discourage growth of the biomass and stimulate

production of fumaric acid. The production fermenter is a continuous airlift-loop type, which operates at a temperature of 35°C and a pH of 5.0. Due to its continuous characteristics, 95% of the glucose is converted into fumaric acid, while the other 5% remains within the fermentation broth. Air and CO₂ are fed into the reactor for aeration and agitation and as the additional carbon source, according to the tricarboxylic acid (TCA) cycle. As fumaric acid is produced, the acidity of the fermentation broth increases and causes product inhibition. This is solved by neutralization of the broth by the continuous addition of NaOH. Ethanol is formed as a byproduct. Prior to downstream processing, 99.7% of the biomass is recovered from the fermentation broth by a rotary drum filter and recycled to the production fermenter.

PROCESS DESCRIPTION BASE CASE: CHEMICAL pH-SHIFT CRYSTALLIZATION

The process flow diagram (PFD) of the base case is presented in *Figure 1*. The required nutrients are premixed in a vessel (V1) before being distributed over the growth fermenter (V2), the biomass maintenance tank (V3), and the production fermenters (V4). Prior to downstream processing, the biomass is filtered (F1) and recycled to the reactor.

Biomass filtration and separation is strongly recommended and commercially practiced to isolate the microorganisms from exposure to the chemical processes and to maintain the optimal growth and production conditions in the fermentation step. Furthermore, cell and cell material leakage to the downstream process may cause numerous operational inconveniences, such as membrane bio-fouling and organic impurities (large molecules such as proteins) in the final product.^{26–28}

The permeate containing the dissolved product (NaHFum/Na₂Fum) is fed to two crystallizers in series (V5, V6), and H₂SO₄ is added

to both vessels to lower the pH to crystallize the fumaric acid. To obtain the fumaric acid crystals, the slurry is filtered and the crystals are collected from the rotary drum filter (F2). Furthermore, water is evaporated from the mother liquor of the pH-shift crystallizer to precipitate the Na₂SO₄ salt. For this purpose, a multi-stage tubular evaporator (E1) is used, of which it was assumed that crystallization of the salt could take place inside. A detailed design of such an evaporator was outside the scope of this paper. The generated slurry is filtrated (F3) to separate the salt from the mother liquor. The salt is disposed as waste, while the mother liquor can be recycled to the evaporator. Since the original mother liquor was saturated with fumaric acid, a fraction of the product produced by fermentation is discarded together with the waste disposal. The condensed water vapor (E2) can, in principle, be re-used in the fermentation process. For this purpose, it should be mixed with nutrients and/or neutralization agents. The details of the process flows and required information for cost estimation of this case, including equipment sizing, are given in Appendix A.

PROCESS DESCRIPTION BASE CASE WITH SALT REGENERATION

The PFD of the base case including salt regeneration is depicted in *Figure 2*. It differs from the base case at the end of the downstream processing section, where the wastewater, which is the mother liquor from the filtration (F2) of the slurry from the crystallizers, is treated with electrodialysis (ED1) instead of evaporation to obtain streams that can be re-used in the fermentation process. The Na₂SO₄ salt in the wastewater is converted into a NaOH solution and a H₂SO₄ solution. The maximum outlet concentration of the H₂SO₄ solution is determined by the stability of the electrodialysis membrane. The relatively low H₂SO₄ concentration (~1 M) requires a large membrane surface area for the conversion of the waste salt. The salt

Fig. 1. Process flow diagram of base case.

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Fig. 2. Process flow diagram of base case with salt regeneration.

separation by electrodialysis is limited to the driving force supplied by the concentration gradient of the salt at the permeate and retentate sides of the ED module. Therefore, complete conversion of Na_2SO_4 requires an infinitely large surface area, and thus, some Na_2SO_4 will still be present in the wastewater after electrodialysis.

The wastewater, stripped of Na_2SO_4 , is partially evaporated (E3) after electrodialysis to obtain pure water that can be reused in the fermentation process. The wastewater, which was the mother liquor of the pH-shift crystallization (V5, V6), is saturated with uncharged fumaric acid (H_2Fum), since uncharged molecules do not migrate through the electrodialysis membranes. Hence, a small amount of additional fumaric acid crystals will be obtained as a consequence of the evaporation of the wastewater. These crystals are separated from the mother liquor after filtration (F3). The mother liquor contains most of the impurities of the process, i.e., the remaining Na_2SO_4 after electrodialysis and some glucose. This mother liquor is partially discarded as waste to remove the accumulated impurities and partially recycled to the electrodialysis unit to minimize the Na_2SO_4 waste.

The fermentation process (V2, V4) requires the addition of base to neutralize the production of fumaric acid. A NaOH solution is produced by electrodialysis, of which the basicity is adjustable by selecting the proper flow rate and initial NaOH concentration of the solution that collects Na^+ . These parameters can be regulated in such a way that the NaOH concentration of the outlet solution matches with the fermentation requirements. Then, this NaOH solution can be recycled directly to the fermenter. On the other hand, the solution that collects Na^+ should be recycled as well. For this purpose, the produced NaOH solution is split into two parts. Condensed

water (E1), originating from the evaporation of wastewater, should be used to dilute the Na^+ collecting solution; otherwise, the NaOH concentration of the outlet solution would be too high due to the conversion of Na_2SO_4 into NaOH inside the electrodialysis module. The disadvantage of this approach is the low NaOH concentration inside the module, which leads to a relatively high resistance and accompanied energy consumption. Alternatively, the NaOH concentration of the Na^+ collecting solution can be set equal to the NaOH concentration of the solution that is recycled to the fermenter. In that case, the complete solution that leaves the electrodialysis module should be diluted with condensed water to such an extent that its NaOH concentration meets the required specification. The alternative approach should be more cost-effective due to the improved conductivity and accompanied process efficiency. This approach was therefore chosen in the conceptual process design of the base case with salt regeneration.

The relatively dilute H_2SO_4 solution that leaves the electrodialysis unit requires evaporation (E4) of the water before re-use in the crystallizer, since otherwise the fumaric acid inside the crystallizers would be diluted too much, reducing yields. The mother liquor from the crystallizer also contains some charged HFum^- as a result of speciation of fumaric acid at the pH at which crystallization occurs. During electrodialysis, the negatively charged compounds will migrate towards the H_2SO_4 solution together with SO_4^{2-} anion. It is assumed that the neutralized fumaric acid does not remain soluble in the concentrated H_2SO_4 solution during evaporation of the water. Hence, a slurry will be obtained, from which the H_2Fum crystals will be separated by filtration (F4). A small fraction of the concentrated acid—the mother liquor of F4—is combined with the condensed

Fig. 3. Process flow diagram of EIC case.

water (E1) and recycled to the electrodialysis unit to collect SO_4^{2-} anions, while the majority of the concentrated acid is re-used for the pH-shift crystallization. The amount of crystalline fumaric acid obtained in this method increases the yield of the process. The details of the process flows and required information for cost estimation of this case, including equipment sizing, are given in Appendix B.

PROCESS DESCRIPTION EIC CASE

The PFD of the biobased process using EIC is presented in *Figure 3*. In this design, the crystallizers are replaced by a stack of modules in which EIC is conducted (EIC1). The filtered fermentation broth flows through these modules and can directly be recycled to the fermenter (V4). Inside these modules, electrolysis of water generates protons that induce crystallization of fumaric acid at the membrane.¹⁵ Simultaneously, OH^- is produced, which neutralizes the pH of the fermentation broth such that less or no NaOH has to be added to the fermenter. The crystalline layer that develops in time is regenerated with warm ethanol periodically (EIC2). The ethanol is cooled to obtain the fumaric acid crystals (V5), which are retrieved from the mother liquor after a solid-liquid separation (F2). During EIC, fermentation broth is incorporated into the crystalline layer. Upon dissolution of this layer with ethanol, the interstitial liquid (consisting mainly of water and glucose) will mix with the regeneration solvent. Therefore, a small part of the mother liquor of F2 is distilled in the distillation column (D1) to discard the excess of water, otherwise, water and glucose will accumulate in the system. The evaporated ethanol is condensed and mixed with the other part of the mother liquor and is as such re-used to dissolve fumaric acid crystals grown inside the EIC stack. The details of the process flows and required information for cost estimation of this case, including equipment sizing, are given in Appendix C. The description of the experimental study of the design and optimization of the EIC unit is presented in Appendix D.

ECONOMIC ANALYSIS

The purchased equipment costs (PEC) corresponding to the process designs for the different cases are given in *Table 1*, along with the operational costs (OPEX) and investments costs (CAPEX). Break-downs of the PEC are shown in *Figure 4*, where unit operations are categorized.

The costs for the fermenters involve the mixing, holding, growth, and production vessels. The filters category concerns the biomass and product filters. The two categories are comparable across the three different cases due to equal sizing of the fermentation section and comparable production rates.

The crystallizers category concerns the operation in which fumaric acid is crystallized; this is equal for both base cases, while crystallization of fumaric acid takes place twice in the EIC process. The first crystallization event takes place inside the EIC modules, while the second event takes place in a crystallizer downstream of the modules. Therefore, the crystallizers category is relatively large for the EIC case.

Table 1. Cost Price of Raw Materials

TYPE OF RAW MATERIAL	COST PRICE (€/TON) ^a
Glucose	448 ^b
Sodium hydroxide	334
Sulfuric acid	73
Process water	0.35
CO ₂	100
Ethanol	425

^a1€ ~ 1.29 USD.

^bSee Ref. 29.

Fig. 4. Purchased equipment costs of base case (A); base case with salt regeneration (B); and the EIC case (C).

The waste treatment category involves all unit operations that deal with the mother liquor after filtration of the product. For the base case, it involves evaporation of the mother liquor and subsequent filtration of the solid waste. For the base case with salt regeneration, it includes the electrodialysis unit, two evaporators, and two filtrations. The evaporator and rotary drum filter that handle the concentrated H_2SO_4 stream are assumed to be twice as expensive compared to normal solutions/suspensions. For the EIC case, waste treatment only involves the distillation unit. In total, the base case with salt regeneration has the highest PEC, which originates from the wastewater treatment, while the PEC of EIC is higher than that of the base case due to the use of the more expensive EIC modules and the additional cooling crystallizer that replaces the relatively cheap pH-shift crystallizers.

The CAPEX are calculated from the PEC using fixed percentages for direct costs, indirect costs, working capital, and start-up costs (Table 2). Therefore, the values of the CAPEX scale with the values of the PEC. The OPEX are calculated with raw materials (costs given in Table 1), waste disposal penalty, utilities, operating supplies, and CAPEX as input. The former four inputs define the variable costs, while fixed costs are defined by fixed percentages of the CAPEX (capital charge, maintenance, local taxes, and insurance) and OPEX (labor, plant overhead, and general expenses).

Three types of waste were identified: wastewater, biomass waste, and solid waste, with disposal penalties of €1/ton (USD 1.3/ton), €10/ton (USD 12.9/ton), and €50/ton (USD 64.2/ton), respectively. For the two cases involving some form of electrodialysis, the cost item for maintenance includes the replacement of membranes every 2.5 years (the lifetime of such ion selective membranes). Conventionally, economic criteria are calculated based on a given sales price of the product. In our case the projected sales price of fumaric acid produced by fermentation is unknown, since fumaric acid is currently produced petrochemically. Therefore our approach was to achieve a fixed return on investment (ROI) of 15%, and the required sales price per kg to obtain this target was calculated. If, in reality, the product can only be sold for a lower price, then the achieved ROI will be lower than 15%. The required sales price is a measure of the competitiveness of the process; i.e., a lower required sales price translates into a better position in the market. To be able to calculate the required sales price, the discount rate (DCR), the rate of return that could be earned on an investment in the financial market with similar risk, was

assumed to be 8% annually. From the ROI and the DCR, the pay-back time can be calculated. Since these instruments are both fixed, the pay-back time is fixed at 9 years. Without the DCR, the pay-back time would be 6.7 years. With the described approach, the required net cash flow (CF) is determined by the ROI and is thus fixed at 15% of the initial investment.

Figure 5 shows the cumulative net present value (CNPV) in millions of Euros plotted against time (year), illustrates the fixed pay-back time with and without the DCR applied, and shows the relation

Table 2. Final Economic Figures of the Three Cases

COST ITEMS	BASE CASE*	BASE CASE WITH SALT REGENERATION*	EIC CASE*
Purchased equipment costs	3.2	9.3	6.0
CAPEX	20.1	58.2	37.9
Direct costs	11.0	31.7	20.3
Indirect costs	3.7	10.8	7.4
Working capital	4.0	11.6	7.6
Start-up costs	1.4	4.1	2.7
OPEX	51.3	65.8	53.3
Fixed costs	21.7	36.9	27.3
Related to CAPEX	5.6	16.1	10.5
Labor	6.4	8.2	6.7
Plant overhead	2.6	3.3	2.7
General expenses	7.2	9.2	7.5
Variable costs	29.5	28.9	26.0
Raw materials	23.8	16.9	17.9
Utilities	3.7	10.4	7.5
Waste disposal	1.7	0.02	0.03
Operating supplies	0.3	1.8	0.6

*In millions of € (1€ ~ 1.29 USD).

Fig. 5. Cumulative NPV plot illustrating fixed pay-back time as a function of ROI (15% and 30%) and DCR (0% and 8%) for the EIC case. Initial investments of the three cases are plotted at the y-axis. The slope of the curve is the discounted CF from which the net CF can be calculated.

between the height of the investment and the required net CF (related to the slopes of the lines). It also indicates that with a higher ROI, the pay-back time is shorter, but the required net CF is higher, resulting in a higher required sales price. The required sales price per kg product is obtained by dividing the required annual sales revenues (millions of €/y) by the annual production of fumaric acid (kg/y). The former is the summation of the gross cash flow (with the tax rate set to 35%), the depreciation of equipment (10% of PEC annually),

Table 3. Economic Criteria of the Three Cases

CRITERIA	BASE CASE	BASE CASE WITH SALT REGENERATION	EIC CASE
ROI (%)	15	15	15
DCR (%)	8	8	8
Net CF*	3.0	8.7	5.7
Tax rate (%)	35	35	35
Depreciation*	0.3	0.9	0.6
Gross CF*	4.7	13.4	8.8
Required sales*	56.2	80.1	62.6
Amount of product (ton/year)	25488	27796	26412
Required sales price (€/kg)	2.2	2.9	2.4

*In millions of € per year.

and the OPEX. The economic criteria of the three cases are given in Table 3.

It can be observed that the base case with salt regeneration has the highest required sales price, which is due to the large investments in the waste treatment section to avoid the disposal of solid waste and the accompanied high operational costs (utilities). The salt-free production of fumaric acid by fermentation integrated with EIC is considerably more competitive, in terms of both CAPEX and OPEX. The use of EIC is slightly more expensive than the base case, in which solid waste is disposed against a penalty of €50/ton (USD64/ton). This is only due to the investments, since the variable part of the OPEX of EIC is lower (Table 1). This is in turn due to the use of electricity instead of the addition of chemicals.

Sensitivity Analysis

A sensitivity analysis was performed to evaluate the effect of glucose price, electricity cost, steel price, and penalty for solid waste (Fig. 6) on the required sales price of fumaric acid for the three different cases. It can be observed that the glucose price is hardly differentiating between the three cases; this is because glucose is dominating the raw materials and, to a large extent, the OPEX.²⁹ At glucose prices above €1,400/ton (USD1,799/ton), fermentation integrated with EIC becomes the most competitive process. This is due to the fact that EIC is more efficient in terms of glucose consumption. In the EIC case, the glucose that is not converted during fermentation is directly recycled to the fermenter after passing through the stack of EIC modules. The non-converted glucose, assumed to be 5% in all cases, ends up in the waste treatment section in both base cases and is, therefore, not recycled to the fermenter.

The effect of electricity on the required sales price of the base case with salt regeneration and the EIC case is larger than on the required sales price of the base case. Therefore, at lower electricity prices (below €0.025/kWh; USD0.032/kWh), EIC becomes economically more attractive than the base case. The difference in dependency is explained by the electrodialysis unit, which is dominating the utility costs in two of the three cases.

The steel price dependency is calculated by varying the total PEC to 50% or 150% of the original calculated value. Since the PECs were significantly different (€3, €6, and €9 million; USD3.9, 7.7, and 11.6, respectively), the effect of the steel price is quite significant. At lower steel prices (below 60% of the original PEC), the EIC case is economically most attractive. However, at higher steel prices, the base case clearly remains the cheapest process.

Since the two cases in which electrodialysis is applied do not produce solid waste, the base case is the only case affected by variations in the solid waste penalty. The effect of this penalty is quite drastic, as shown in Figure 6D. Already at €128/ton (USD164/ton), EIC is economically more attractive than the base case, while at €365/ton (USD469/ton) the disposal of solid waste becomes more expensive than the regeneration of the salt into the original acid and base.

In reality, it is expected that glucose will be more expensive due to competition between biofuels, biochemicals, and the food market.

Fig. 6. Effect of glucose price (A), electricity costs (B), steel price (C), and penalty for solid waste (D) on cost price of fumaric acid.

On the other hand, the trend is to replace glucose with cheaper, second-generation feedstocks that are not competing with food. Furthermore, it is hard to forecast the electricity price, but with the current trends of increasing consumption, it is not realistic to expect decreasing electricity costs. The same is true for the steel price, which has rapidly increased during the last decade due to economic growth of the BRIC countries [Brazil, Russia, India, and China]. These considerations indicate that EIC will most likely not be much cheaper than classical pH-shift crystallization. On the other hand, sustainability is a driver toward clean processes that produce zero waste. In that respect, waste penalties might increase, such that EIC could soon become economically attractive. As discussed in the introduction, the disposal of solids might well be prohibited, and hence, the base case is

not valid. Figure 6 shows that in any situation, EIC is the most economically attractive option of the two remaining possibilities.

Discussion

Waste streams in the process designs discussed in the appendices have not been considered for recycling, which might be of particular interest for the two cases that do not generate solid waste. These processes both produce wastewater, which after further treatment can be reused in the processes. The base case with salt regeneration has a purge stream that contains a considerable amount of glucose (17 wt%), making recycling to the fermenter an interesting option. A drawback might be the amount of salt (3 wt%) that is present in this purge stream as well, which can lead to undesired osmotic pressure

during fermentation. The presence of salt downstream of the fermentation process is not a problem since electrodialysis is applied to regenerate the salt. The EIC case has two purges; the first minimizes the ethanol content during fermentation, while the second purge prevents accumulation of water and glucose in the ethanol used for regeneration of the EIC modules. The first purge can easily be recycled to the fermenter after evaporation of the ethanol. This can, for instance, be accomplished by pervaporation, which is less CAPEX intensive than distillation. Complete removal is not required—so the added cost is not large—and valuable components such as glucose, NaOH, and some dissociated fumaric acid are recovered. The second purge consists of ethanol, water, fumaric acid, and glucose. This stream can also be recycled to the fermenter after evaporation of the ethanol. In essence, a process with 100% glucose efficiency will be obtained in both cases, which will positively affect the OPEX of both processes, while less waste is generated.

In the process design of the EIC case, the unit operation EIC consists of a number of stacks that each contains a large number of modules. This is determined by geometrical considerations derived from experimental work. The actual design (module complexity, manifold, etc.) and operation (distribution of flows, monitoring, etc.) of these modules has similarities with microreactor technologies. Lessons learned in this evaluation can probably be applied to solve some of the issues that might be encountered in the development of large-scale EIC technology.³⁰

In this paper, the base case without salt regeneration is considered the least sustainable process due to the formation of 1.3 tons of salt per ton of product. Another commonly used parameter that determines the sustainability of a process is the CO₂ footprint. A detailed analysis was outside the scope of this paper, but a few interesting comments about the overall process can be posted. First, the metabolic pathway to produce fumaric acid utilizes approximately 1 mole of CO₂ per mole of produced fumaric acid.²³ As such, the unit operation fermentation has a negative CO₂ footprint, which obviously is positive for the sustainability of the overall process. Second, although the other unit operations do not consume or emit greenhouse gases directly, the utilities (e.g., electricity, steam, etc.) required for these processing steps are generated from fossil resources. Hence, these steps indirectly emit CO₂, which needs to be taken into account in the sustainability assessment. At €3.7 million (USD4.8 million), the base case without salt regeneration has the lowest utility costs (versus €10.4 million (USD13.3 million) and €7.5 million (USD9.6 million) for the base case with salt regeneration and the EIC case, respectively), and therefore appears to have the lowest CO₂ footprint. The other two cases require more utilities because they either recycle or prevent the formation of waste salt. It seems that preventing the formation of waste requires greenhouse gas emissions. However, considering the production processes of NaOH and H₂SO₄, utilities are required as well. NaOH is produced by electrodialysis of water, which is exactly the same process that is applied by EIC, while H₂SO₄ is also an energy-intensive process. These utilities are generated from fossil resources, such that the production of the required chemicals is accompanied with CO₂

emissions as well. Hence, the base case without salt regeneration has a CO₂ footprint comparable to the EIC case. It might be even higher due to loss of conversion efficiency, transport of chemicals, etc. Therefore, the EIC case is the more sustainable process, since no waste salt is produced.

Current biotechnological developments in the field of carboxylic acid production are focused on low pH fermentations.^{31–34} Here, the acid is formed directly, obviating the need for a pH shift. This process will thus intrinsically exhibit a lower CO₂ footprint compared to state-of-the-art neutral fermentations. The carboxylate product sink will become obsolete, such that integrated product recovery becomes even more pivotal to alleviate the toxicity and inhibitory constraints of the produced carboxylic acid. Potential product recovery techniques applicable for low pH fermentations are in-stream cooling and evaporative crystallization, co-crystallization, adsorption, and extraction.^{35–40} In the long term, the novel low-pH fermentation process will be the logical choice. Until then, the sustainability of current fermentations that produce carboxylic acids at neutral pH can be improved by applying EIC.

Conclusions

The base case has the lowest required product sales price, followed by the EIC case and base case with salt regeneration. The application of EIC will only be cheaper than classical pH-shift crystallization when the penalty of solid waste increases to above €128/ton (USD164/ton). When the disposal of solid waste is prohibited, EIC is the most economically attractive option for continuous product recovery from fermentation.

The base case produces 1.3 tons of solid waste per ton of product, while the other two processes do not generate solid waste at all. Hence, the base case is the least sustainable process in terms of waste. The metabolic utilization of CO₂ for the production of fumaric acid is beneficial, while on the other hand, CO₂ is indirectly emitted by the utilities consumed. Although a considerable amount of electricity is required, the EIC case is the most sustainable process in terms of CO₂ footprint since the production of the chemicals required in the base case also leads to indirect CO₂ emissions.

This techno-economic evaluation shows that changing the purpose of electrodialysis from the regeneration of waste salt to crystallization of the fermentation product is cost effective by reducing the required capital investments and the amount of water that has to be evaporated.

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Author Disclosure Statement

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(Appendix follows →)

Appendix A—Process Analysis of the Base Case

A simplified process flow-sheet for the base case as depicted in Figure 1 is shown in Figure A.1, where the mass streams (ton per ton product), important process conditions (T and P), and important equipment sizes are indicated. The thicker lines show the flow of feedstock glucose and product fumaric acid. The overall process yield of fumaric acid on glucose of the base case was 81% during the production of fumaric acid in 7,920 hours.

A total of 33.7 ktons solid waste is produced after the evaporation of wastewater. A small part of the evaporated wastewater cannot be reused due to the water required for biomass transport into the fermenter and the formation of water as a result of broth neutralization. As such, the evaporated water is discarded as wastewater (20.8 ktons). The biomass purge is also discarded as waste (1.20 ktons with 61 wt% biomass). In

total, 55.7 ktons waste is produced. Besides the waste, some ethanol can be retrieved from the evaporation of wastewater as well. It was assumed that a 1:1 mixture of water and ethanol can easily be obtained by fractional condensation. The ethanol can, in principle, be sold as a by-product, which is not considered in this process design. More details of the most important unit operations are given in Table A.1.

The sizes of the streams, given in Figure A.1, are determined by assumptions concerning the production rate of fumaric acid (0.111 kg/kg microorganism), and the concentration of the microorganisms in the continuous production fermenter (10 wt%). The size of this fermenter is based on the total amount (15.9 ton) of biomass required for the production of 25.5 ktons fumaric acid, accounting for fumaric acid in the purge and waste (in total 2.5 ktons), the density of the fermentation broth (1,025 kg/m³), and for aeration (33%) and working volume (85%) of the fermenter. Sizes of the batch

Fig. A.1. Simplified block scheme diagram of base case.

Table A.1. Detailed Information of Important Unit Operations of Base Case

SPECIFICATION	FERMENTER	CRYSTALLIZER	EVAPORATOR	FILTERS		
	V4	V5 & V6	E1	F1	F2	F3
Type	Airlift-loop	MPV ^a	Multistage (4 stages)	RDF ^b	RDF	RDF
Volume (m ³)	243	45	-	-	-	-
Area (m ²)	17.3	5.6	21.0	6.5	6.7	7.6
Flux (m ³ /m ² h)	-	-	-	4.0	3.6	0.88
Number	2	2	1	1	1	1
Series	1	2	1	1	1	1
Parallel	2	-	-	-	-	-

^aMulti purpose vessel with agitation accessories.

^bRotary drum filter.

growth fermenter (254 m^3) and corresponding vessels (mixing: 198 m^3 ; maintenance: 210 m^3) are based on the amount of biomass that is purged after filtration (0.3% of the total amount of required biomass). These additional values for the batch section of the process design are used as such in the economic evaluation.

Filtrations are carried out using rotary drum filters, with an assumed submerged angle of 120° at 1 rpm. The thickness of the cake layer is assumed to be 0.008 m, while the cake densities for biomass and fumaric acid crystals are assumed to be $1,250 \text{ kg/m}^3$ and $1,500 \text{ kg/m}^3$, respectively. From the fluxes given in Figure A.1, the required surface area for filtration was calculated.

Two crystallizers in series were used, with 75% of the acid dosed in the first crystallizer and the remaining 25% in the second, to establish a narrow size distribution suitable for filtration. For the calculation of the size of these crystallizers, an overall growth rate of 10^{-8} m/s is assumed and a particle size of $100 \mu\text{m}$. Furthermore, it is assumed that the size of each of these crystallizers in series is 50% of the size of a crystallizer that would be required for the residence time (2.78 h) as calculated from the parameters mentioned previously. Additionally, a fluid density of $1,025 \text{ kg/m}^3$ and a working volume of 75% are assumed.

The calculation of the size of the tubular evaporator accounts for the mole fraction of water. It is assumed that the complete amount of water and ethanol are evaporated. The required heat transfer is enhanced by forced circulation, and an averaged value for the heat transfer coefficient ($1.95 \text{ kW/m}^2\text{K}$) is obtained (www.cheresources.com/uexchangers.shtml). The temperature of saturated steam (179.9°C) at 10 bars is obtained (<http://www.econosto.com/A4/Res/A4/data/sectie27/653.pdf>). The boiling point elevation

($\Delta T = 1.9^\circ\text{C}$) was calculated from the amount of salt in the wastewater (15 wt%).

Appendix B—Process Analysis of the Base Case with Salt Regeneration

A simplified process flow sheet for the base case with salt regeneration, depicted in Figure 2, is given in Figure B.1. Here the mass streams, important process conditions, and equipment sizes are indicated, while more details are provided in Table B.1. As explained above, a small amount of extra fumaric acid will be produced inside the electrodialysis unit. The values given for the process streams are not based on the additional obtained amount, but only on the amount of fumaric acid retrieved from the crystallizer. The overall process yield of the base case with salt regeneration is 89% when the extra amount of fumaric acid from electrodialysis is taken into account. Otherwise, the same yield as the base case without salt regeneration is obtained. A total of 27.8 ktons of fumaric acid is produced in 7,920 hours.

No solid waste is produced due to electrodialysis, while the amount of wastewater is also significantly reduced (from 20.8 to 8.8 ktons). The wastewater consists predominantly of water (79 wt%) and glucose (17 wt%), while it contains only a fraction of Na_2SO_4 (3 wt%) originating from the incomplete conversion during electrodialysis (since otherwise an infinitely large surface area would be required). The water in this waste stream is in excess, due to the addition of water to the fermentation process as a result of biomass transport into the fermenter. Since no solid waste is produced, this excess of water is not evaporated, but purged. The waste stream should be purged; otherwise glucose will accumulate in the electrodialysis unit. The

Fig. B.1. Simplified block scheme diagram of base case with salt regeneration.

Table B.1. Detailed Information of Important Unit Operations of Base Case with Salt Regeneration

SPECIFICATION	FERMENTER	CRYSTALLIZER	ED MODULE	EVAPORATORS		FILTERS			
	V4	V5 & V6	ED1	E1	E2	F1	F2	F3	F4
Type	Airlift-loop	MPV ^a	Classical	Multi-stage	Multi-stage	RDF ^b			
Volume (m ³)	243	45	-	-	-	-	-	-	-
Area (m ²)	17.3	5.6	13,684	19.8	54.5	6.5	6.7	0.22	0.38
Flux (m ³ /m ² h)	-	-	-	-	-	4.0	3.6	5.3	4.0 ^c
Number	2	2	1	1	1	1	1	1	1
Series	1	2	1	1	1	1	1	1	1
Parallel	2	-	-	-	-	-	-	-	-

^aMulti purpose vessel with agitation accessories.^bRotary drum filter.^cThe flux through the filter is calculated with a fluid density of 1,800 kg/m³ (predominantly H₂SO₄).

same amount of biomass waste is produced as with the base case, such that the total amount of waste amounts to 10.0 ktons annually.

Similar to the base case, some ethanol can be retrieved from the evaporation of wastewater as well. It was assumed that a 1:1 mixture of water and ethanol can easily be obtained by fractional condensation. Additionally, electrodialysis results in the formation of gas (both O₂ and H₂), which can be sold as byproducts (not considered in this process design).

The same considerations for the fermenters, filters, crystallizers, and evaporators as in the base case were taken into account. It should be noted that the simplified electrodialysis unit also consists of the two evaporators (E1, E2) and their corresponding filters (F3, F4). The size of the stream (7.9 tons per ton of product) that has to be processed by evaporator E1 is comparable to the size of the wastewater stream (8.8 tons per ton of product) treated by the evaporator of the base case. It is slightly smaller, since most (99%) of the Na₂SO₄ that was in the wastewater is converted into NaOH and H₂SO₄ by electrodialysis. The size of the stream (21.3 tons per ton of product) processed by evaporator E2 is considerably larger. This is due to the maximum allowable H₂SO₄ concentration in the electrodialysis unit, which results in diluted, and thus large, streams. The boiling point elevation in evaporator E1 amounts to 0.4°C, which is caused by fumaric acid, glucose, and the remaining amount of Na₂SO₄ (in total 3.3 wt%). The boiling point elevation in evaporator E2 is 0.6°C, due to contributions of fumaric acid and H₂SO₄ (in total 4.9 wt%). These additional numbers were used to calculate the size of the evaporators and filters inside the simplified electrodialysis unit. The values given in *Figure B.1* and *Table B.1* were used for equipment sizing and the calculation of operating costs.

Appendix C—Simplified Process Analysis of the EIC Case

A simplified process flow sheet of the EIC case, presented in *Figure 3*, is given in *Figure C.1*. Here the mass streams, important process conditions, and equipment sizes are indicated, while more details are provided in *Table C.1*. The dashed arrows to and from the EIC unit

operation represent the semi-continuous character of those flows. The unit operation EIC consists of multiple stacks of electrolysis modules. Crystallization takes place in the majority of these stacks, while in the remaining stacks the generated crystalline layer is dissolved in ethanol. So the flows are continuous, but are distributed batch-wise over the different stacks. The cell-free fermentation broth is circulated over the stacks and directly recycled to the fermenter. Compared to both the base cases, no evaporation of water prior to its reuse is required. The overall process yield of fermentations integrated with the EIC concept is 88% with a production of 26.4 ktons per year. Although the production is lower, the yield of the EIC case is equal to the yield of the base case with salt regeneration due to the recycling of fermentation broth, which ensures that the non-consumed glucose can be re-used.

No solid waste is produced during EIC, but in addition to the biomass purge stream (1.2 ktons), two other purge streams are required: the purge after the EIC modules (7.7 ktons with 80 wt% water) and the purge after distillation (8.7 ktons with 67 wt% water and 18 wt% ethanol). These purges are required to prevent the accumulation of glucose and water in the EIC process and the regeneration process with ethanol, respectively. In total, the production of waste amounts to 17.6 ktons.

Similar to the electrodialysis unit in the base case with salt regeneration, EIC results in the formation of gas. Here it was assumed that H₂ was recycled to the fermenter alongside with the fermentation broth, where it was vented together with the exhaust gas mixture.

The calculation of the sizes of the batch and continuous sections of the fermentation process are calculated with the same approach used for both base cases. The mass flows are slightly higher, which is mainly due to accumulation of ethanol by the applied recycle from the EIC unit operation. The sizes and fluxes of the filters are calculated following the procedure described above. The size of the EIC equipment is calculated based on experimental work, which is described in the following section.

ELECTROCHEMICALLY INDUCED CRYSTALLIZATION

Fig C.1. Simplified block scheme diagram of EIC case.

The size of the crystallizers is calculated according to the approach used for both base cases. Instead of acidification, cooling is applied to generate a driving force. Similar to the calculation of the base cases, 75% of the driving force is applied to the first crystallizer, while the remaining 25% is applied to the second one. All parameters, except for the fluid density, are assumed to be equal to the base cases. The fluid density of the mixture of fumaric acid, water, and predominantly ethanol is assumed to be 850 kg/m³.

The relatively large ethanol streams and the distillation of the purge are required to minimize the effect of water and glucose on the dissolution of fumaric acid from the membrane and its crystallization, provoked by cooling of the ethanol. The size of the ethanol stream follows from the assumption of 20 wt% interstitial liquid, the

separation factor, and the sizes of the streams that are distilled and purged. It was assumed that a separation factor of 0.93 can be achieved (initially 80 wt% ethanol). A detailed design of the distillation unit was outside the scope of this paper. Therefore, it was assumed that the same tubular evaporator with forced circulation used for both base cases can also be used to approximate distillation. Hence, the same heat transfer coefficient and saturated steam pressure were used to calculate the required surface area. To calculate the heat duty of the tubular evaporator-based distillation unit, the boiling point of the ethanol-water mixture is required. It is estimated from the phase diagram of ethanol and water and amounts to 79.5°C with a mole fraction of ethanol of 0.61. This relatively low mole fraction is due to the considerably large amount of interstitial liquid

Table C.1. Detailed Information of Important Unit Operations of EIC Case

SPECIFICATION	FERMENTER	EIC	CRYSTALLIZER	DISTILLATION	FILTERS	
	V4	EIC1&2	V5	D1	F1	F2
Type	Airlift-loop	Custom	MPV ^a	Multistage	RDF ^b	RDF
Volume (m ³)	243	-	236	-	-	-
Area (m ²)	17.3	2,766	17.0	2.2	6.5	6.9
Flux (m ³ /m ² h)	-	-	-	-	4.2	18.5
Number	2	1	2	1	1	1
Series	1	1	2	1	1	1
Parallel	2	-	-	-	-	-

^aMulti purpose vessel with agitation accessories.

^bRotary drum filter.

(20 wt%). Furthermore, the fraction of the stream (used for regeneration of the EIC module) that is distilled is relatively small (1.5 tons per ton product), such that only a small amount of water and impurities is discarded. With an increased fraction of distilled regeneration solvent, the mole fraction of ethanol would increase. However, this would require a larger distillation unit.

Appendix D—Explanation of EIC

The stacks of modules in which EIC is performed are the central component of the PFD of fermentations integrated with EIC. A brief explanation of the governing principle is schematically depicted in *Figure D.1*. The cell-free fermentation broth containing the dissolved sodium fumarate salt flows through the cathode compartment (*right*). An acidic buffer (1 M H₂SO₄) is directed through the anode compartment to optimize the conductivity of the solution inside the module. The acid is auxiliary; it is not consumed in the process.

When an electrical potential is applied, water oxidizes at the anode to form H⁺ and O₂ and reduces at the cathode to form OH⁻ and H₂:

As a result of the applied electric field, H⁺ and other cations migrate towards the cathode, while OH⁻ and other anions (e.g., carboxylate ions) migrate towards the anode. The cation exchange

membrane (CEM) is selective for cations. Therefore, H⁺ ions present in the anode compartment migrate through the membrane. Consequently, a localized acidic environment is created in the vicinity of the CEM. The continuous flow of the fermentation broth guarantees a supply of carboxylates (RCOO⁻) throughout the whole cathode compartment. Carboxylates that are in the vicinity of the CEM associate with protons that are migrating from the CEM towards the cathode. The concentration of uncharged carboxylic acid (RCOOH) at the selective membrane is determined by the product specific dissociation constant *K*, the proton concentration (H⁺) and the carboxylate concentration (RCOO⁻) near the CEM:

$$(3) \quad K = (\text{RCOO}^-)(\text{H}^+) \div (\text{RCOOH})$$

Equation (3) shows that the concentration of RCOOH will exceed its solubility ($C > C_0^*$), when both the bulk concentration of RCOO⁻ and the proton concentration are high enough. Due to the concentration gradient of H⁺ across the cathode compartment, the driving force for crystallization ($C > C_0^*$) will be maximal at the CEM. There is a large chance that the formed crystals attach to the membrane, as association of RCOO⁻ with H⁺ takes place directly when protons permeate the membrane.

The efficiency of product removal by EIC is influenced by the broth velocity through the module parallel to the membrane, the migration velocity of ions, and the carboxylate concentration in the bulk of the fermentation broth, as concluded in previous work.¹⁵ The ratio of broth velocity and migration velocity, illustrated by the arrows in the insert of *Figure D.1*, effectively determines the amount of carboxylates that travels from the bulk towards the CEM where crystallization takes place. With a decreased ratio, more carboxylates can be converted into crystalline carboxylic acid in a single pass of broth through the module. These three parameters were the subject of experimental investigation to determine the crystallization efficiency of EIC. The CAPEX and OPEX of this unit operation were deduced from the experimental data.

MATERIALS AND METHODS

Fumaric acid (99%) and sodium hydroxide (98%) were obtained from Sigma-Aldrich and used as received. Demineralized water was used to prepare solutions of sodium fumarate. The experimental setup, described in previous work, was used to perform the experiments to determine the efficiency of EIC.¹⁵ Flow rates of 150, 300, and 450 mL/min; current densities of 30, 40, and 50 mA/cm²; and fumarate concentrations of 20, 40, and 70 g/L were used.

EXPERIMENTAL RESULTS OF EIC WITH FUMARIC ACID

Figure D.2 presents the results of EIC experiments in which flow rate (proportional to the broth velocity), current density (proportional to the migration velocity), and initial fumarate concentration were varied. The amount (g) of crystalline fumaric acid accumulated at the membrane is plotted against time. Symbols represent measurements (recalculated from bulk concentration), and lines represent linear trendlines to guide the eye. *A* and *B* show results from experiments with an initial fumarate concentration of 40 g/L. In *A*, the flow rate

Fig. D.1 Principle of EIC. The fermentation broth is circulated through the cathode compartment, the anolyte (contents of anode compartment) consists of H₂SO₄. Electrolysis of water occurs at the electrodes. Cations migrate towards the cathode and anions towards the anode. Protons migrate through the cation exchange membrane (CEM), create a local acidic environment and associate with carboxylate ions (RCOO⁻, opposite charge). When the carboxylic acid concentration exceeds the solubility, solid material is deposited alongside the CEM (indicated by shaded area). The insert schematically depicts the important ratio between flow rate and migration velocity.

Fig. D.2. Experimental results of EIC. (A) and (B) experiments with 40 g/L; (A) with varied flow rate (mL/min), (B) with varied current density (mA/cm²). (C) and (D): all data from experiments with 20 (red), 40 (green), and 70 g/L (blue), (C) with varied flow rate, (D) with varied current density. Color images available online at www.liebertpub.com/ind

was varied with constant current density (50 mA/cm²), while in B the current density was varied with constant flow rate (150 mL/min).

These results show that the use of a lower flow rate leads to larger amounts of solid material deposited at the membrane. This indicates that decreased flow rates result in increased crystallization efficiencies. Furthermore, with a higher current density more fumaric acid accumulated (B). C and D show the effect of initial fumarate concentration on the performance of EIC. The green data in these figures are identical to the data used in A and B, as indicated by the arrows. In general, it can be observed that higher concentrations (blue versus red) lead to increased crystallization performance. For higher concentrations, the effect of the flow rate becomes less pronounced, while the current density remains important. This is probably

explained by the fact that at high fumarate concentrations the proton concentration (proton supply by electrolysis), rather than the fumarate concentration (fumarate supply by broth circulation), becomes limiting (Equation 3). According to the crystallization performance, high product concentrations are desirable as well as high current densities and low flow rates.

In time, the thickness of the crystalline layer, deposited at the CEM, progressively increases. In previous work it was shown that the crystallization performance decreased in time. Regeneration of the module, such that the crystalline layer dissolved in an auxiliary solvent, was required to restore initial product removal rates.¹⁵ The data given in Figure D.2 demonstrate the same trend. The amount of fumaric acid accumulated at the CEM during the experiment with

40 g/L fumaric acid at 50 mA/cm² and 150 mL/min seemed to stagnate after approximately 40 minutes. This can also be observed at higher concentrations, while a plateau is reached after approximately 100 minutes at lower concentrations. Since these data seem to confirm the need for regeneration, the mass balance concerning the EIC case was designed such that regeneration of the module was taken into account. Here, the broth containing the fermentation product was circulated through the module during 40 minutes, after which the crystalline layer was dissolved in ethanol. It was assumed that dissolution is complete in 5 minutes, while another 5 minutes are required to wash the modules with water to avoid accumulation of ethanol in the fermenter. Multiple stacks of modules were used to guarantee a semi-continuous removal of product from the broth.

Module dimensions can be calculated from the optimal settings for product removal by EIC. Here not only the crystallization performance is of importance, but also the costs associated with the selected process condition. *Figure D.3* depicts the measured electricity consumption rate (*left*) and the specific electricity consumption per kg of crystallized fumaric acid (*right*) as a function of the applied current density and flow rate (50 mA/cm²). At the left, the results are as expected. Higher current densities consume more electricity, while higher concentrations result in improved conductivity, such that they consume less electricity compared to lower concentrations. The figure at the right is obtained when the electricity consumption rate at the left of *Figure D.3* is divided by the slope of the crystallization performance. This slope is determined using the data from *Figure D.3* up to 40 minutes. *Figure D.3 (right)* shows that the specific electricity consumption is high at low product concentrations, which is due to

the unfavorable combination of low conductivity and limited performance. At higher concentrations this ratio becomes more favorable. The specific electricity consumption is minimal at high product concentrations (70 g/L) and low flow rate (i.e., low broth velocity). For the high product concentrations, it seems the improved crystallization performance at elevated current densities cancels the increased electricity consumption. *Figure D.3* thus demonstrates that the specific electricity consumption for EIC at high concentration and low flow rate is more or less constant. The performance, however, increases with increasing current density, such that the amount of recovered crystals increases. Therefore, less membrane surface area is required compared to lower current densities. Consequently, the CAPEX decreases at increasing current densities. The optimal conditions with minimal costs that have been used in the design of the EIC process are therefore high product concentrations (> 70 g/L), high current density 50 mA/cm², and low flow rate (150 mL/min).

The module dimensions that correspond to these experimentally determined optimal process settings can be estimated from the experimental data. In the experiments, the distance between the CEM and the electrode (the width of one compartment) was 0.01 m, while the depth was 0.1 m. With the optimal flow rate of 150 mL/min, the broth velocity parallel to the membrane surface becomes 0.0025 m/s.

Industrially, a compartment width of 0.004 m is used for electro-dialysis to decrease the electrical resistance inside the module. Therefore, this value is used in the process design of EIC. To ensure equal crystallization conditions within the module, the broth velocity parallel to the membrane should be identical to the broth velocity in the module with a compartment width of 0.01 m, rather than

Fig. D.3. Experimental electricity consumption rate (*left*); experimental specific electricity consumption (*right*).

matching the flows. With the compartment width fixed at 0.004 m, the depth of the module scales with the applied flow rate when the original broth velocity is kept constant. In the design, the flow rate of 150 mL/min is used to stay close to the experimental observations. Considering this flow rate, the depth of a module should be increased to 0.25 m to maintain the broth velocity of 0.0025 m/s.

The height of the module is related to the maximum migration distance and to the ratio of the broth velocity and the migration velocity. When OH^- would be able to migrate towards the CEM, it will dissolve the crystalline layer at the membrane, diminishing the crystallization efficiency. The maximum migration distance is determined by the width of the module compartment and the maximum thickness of the crystalline layer on the CEM. In the calculations, this maximum thickness is assumed to be 0.001 m, such that the migration distance becomes 0.003 m. The calculation of the migration velocity is explained in our previous article on EIC.¹³ Its value becomes 9.5×10^{-5} m/s with a temperature of 35°C and a current density of 50 mA/cm² (corresponding to a $\Delta\Phi$ of approximately 10V). Hence, the velocity ratio becomes 26.4, while the multiplication of the maximum migration distance with this ratio yields the maximum height 0.079 m. When the height of the module would be higher than this value, OH^- would reach the crystalline layer and dissolve the carboxylic acid crystals. *Figure D.4* illustrates the module dimensions calculated above.

The semi-continuous character of crystallization, regeneration, and washing can be established by using a system similar to a simulated moving bed. A graphical representation is depicted in *Figure D.4* (right). The number of modules required for crystallization, regeneration, and washing, as well as the actual OPEX and CAPEX of the EIC unit operation can be determined from the experimental data based on the module dimensions as presented above. The required number of modules can be deduced from the compartment dimen-

sions and the required total membrane surface area for EIC. The surface area is a function of the annual production and the experimentally determined "crystallization flux" in kg/m²h. This flux is the slope of the most optimal crystallization performance corrected for the amount of surface area used in the EIC experiments. For the process setting 70 g/L, 50 mA/cm, and 150 mL/min, this flux is 1.51 kg/m²h. With the annual production rate of 26.41 ktons in 7,920 hours the required membrane surface area for crystallization is 2,213 m². The surface area of the membrane within one module is $0.079 \times 0.25 = 0.01975$ m², such that the number of modules becomes 112,046. Additional modules are required for regeneration and washing. Each step is finished in 5 minutes, while crystallization proceeds for 40 minutes. Hence, 20% $[(5+5)/(40+5+5)]$ of the total number of modules is required for regeneration and washing. Thus, the total number of modules is 140,057, while 2,766 m² of membrane surface area is required in total. The number of modules per stack, where each stack represents a time frame of 5 minutes, is 14,006. The total volume of installed equipment can be estimated from the total cathode compartment volume corrected for the presence of the anode compartment ($\times 2$) and wiring and tubing ($\times 2$). The total cathode compartment volume is $140,057 \times 0.01975 \times 0.004 = 11.1$ m³, such that the volume of the installed equipment becomes 44.4 m³, with a specific surface area for crystallization of 62.5 m²/m³.

The OPEX is related to the electricity consumption and calculated by multiplying the annual production of fumaric acid with the specific electricity requirement of the module and the cost price of electricity (€0.1/kWh; USD0.13/kWh). The experimentally determined specific electricity consumption belonging to the optimal process conditions is 4.0 kWh/kg product. It is assumed that this value can be translated to other module designs using the ratio of cathode compartment widths. Therefore, the specific electricity consumption of modules that are 0.004m wide corresponds to

Fig. D.4. Module dimensions (*left*); hypothetical representation of stacks (*right*), with EIC modules, including stacks for regeneration (REG) and washing (WASH).

$0.004/0.01 \times 4.0 = 1.61$ kWh/kg. With an annual production capacity of 26.41 ktons fumaric acid, the OPEX related to electricity consumption in the EIC unit operation becomes €4.2 million per year (USD54 million/yr).

The CAPEX are related to the required total membrane surface area ($2,766 \text{ m}^2$) of the stacks of modules. The costs of one pair of electrodes

and one CEM are estimated to be €650/ m^2 (USD831/ m^2), while an additional correction of 25% is used to include costs for wiring, cables, and connections such that the purchased equipment costs related to the required surface area becomes €2.25 million (USD2.88 million). The life-span of electrodes is approximately 10 years, while the life span of the cation exchange membranes is 2.5 years.