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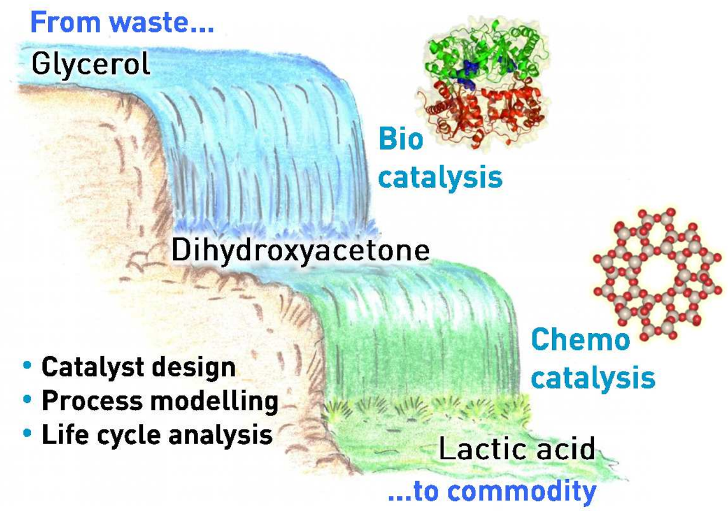
Merten Morales,‡ Pierre Y. Dapsens,‡ Isabella Giovinazzo, Julia Witte, Cecilia Mondelli*, Stavros Papadokonstantakis*, Konrad Hungerbühler and Javier Pérez-Ramírez*



The viability of a novel bio-/chemocatalytic process for lactic acid production from crude glycerol is demonstrated and its benefits with respect to the classical glucose fermentation quantified in terms of sustainability and economics.

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REPORT

Environmental and economic assessment of lactic acid production from glycerol using cascade bio- and chemocatalysis

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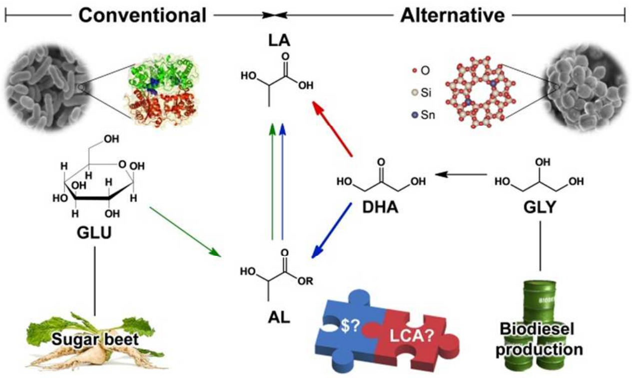
Recently, lactic acid has emerged as one of the most relevant platform molecules for the preparation of bio-chemicals. Due to the limited productivity of sugar fermentation, the dominant industrial technology practiced for its manufacture, new chemocatalytic processes are being developed in order to meet the expected demand for this intermediate. The Lewis-acid catalysed isomerisation of dihydroxyacetone has attracted particular interest. If the reaction is performed in water, lactic acid is attained directly, while if alcohols are used as the solvent the desired product can be obtained upon subsequent hydrolysis of the alkyl lactates formed. Herein, we (*i*) demonstrate tin-containing MFI zeolites prepared by scalable methods as highly active, selective and recyclable catalysts able to operate in concentrated dihydroxyacetone aqueous and methanolic solutions, and (*ii*) reveal by life cycle analysis that a process comprising the enzymatic production of dihydroxyacetone from crude glycerol and its chemocatalytic isomerisation in methanol is advantageous for the production of lactic acid compared to glucose fermentation in terms of both sustainability and operating costs. In particular, we demonstrate that the reduced energy requirements and CO_2 emissions of the cascade process originate from the valorisation of a waste feedstock and from the high performance and recyclability of the zeolite catalyst and that the economic advantage is strongly determined by the comparably low market price of glycerol. It is also shown that the bio-/chemocatalytic route remains ecologically and economically more attractive even if the purity of glycerol is as low as 38%.

1 Introduction

In the last years, many research efforts have been directed to the conversion of bio-derived feedstocks into chemicals as a means to address the forthcoming oil shortage, mitigate global warming and meet the consumers' demand for greener products.[1] In this context, lactic acid (LA) has gained significant interest.[2] This compound not only can generate multiple commodity and intermediate chemicals (*e.g.*, acrylic acid, 1,2-propanediol, pyruvic acid, acetaldehyde, 2,3-pentanedione) that already belong to our traditional chemistry portfolio, but it can also be polymerised into a novel biodegradable plastic suitable for packagings, *i.e.*, polylactide (PLA). This polymer was already industrially produced in an amount of 180 kton in 2012 and its manufacture is projected to reach 1 Mton year^{-1} by 2020. This value might even be exceeded considering that PLA could replace PET (58 Mton in 2012) in 20% of its current applications.[2,3]

Nowadays, LA is prepared *via* the anaerobic batch fermentation of glucose (GLU) or sucrose, under strict temperature (< 313 K) and pH (5-7) conditions, followed by purification through subsequent esterification, distillation and hydrolysis (Scheme 1).[4] One prominent environmental issue related to this manufacturing method is the coproduction of large amounts of gypsum (1 ton per 1 ton of LA), which has to be disposed. Moreover, as typical of biocatalysis, the low productivity (1-13 g_{LA} L^{-1} h^{-1}) of the process may be a high hurdle for the expansion of this technology to fulfil the future demand for LA.[4] Accordingly, alternative pathways based on chemocatalysis have been explored,[5] such as the aqueous-phase isomerisation of 1,3-dihydroxyacetone (DHA) over Lewis acids.[2] Lanthanum and lead salts have proved highly active for this reaction, even enabling the one-pot production of LA from cellulose.[5] Still, owing to their complex and energy-demanding separation from the reaction mixture as well as their toxicity, solid catalysts have attracted more pronounced interest (Scheme 1).[6] So far, research studies have been mainly devoted to the identification of active and selective materials .[7] In this respect, Sn-containing BEA and MWW zeolites stand as the best performers, yielding > 96% LA.[6,8] The surprisingly lower LA yields over Sn-MFI zeolites is likely due to the suboptimal reaction conditions applied in their evaluation.[9] The LA yield



Scheme 1 Conventional lactic acid (LA) production *via* glucose (GLU) fermentation *versus* an alternative hybrid bio-/chemocatalytic route based on the enzymatic oxidation of glycerol (GLY) to 1,3-dihydroxyacetone (DHA) followed by isomerisation in water or in methanol *via* an alkyl lactate (AL) intermediate. Thick and thin arrows define the reactions for which experimental data have been derived from this work and from the literature, respectively.

over Al-based Lewis-acid solids has reached a maximum of 82% over alkaline-treated MFI zeolites.[10] In spite of these promising results, other critical aspects at the catalyst and process levels have been infrequently investigated in view of a potential large-scale application of this route.

At the *catalyst* level, crucial parameters are the scalability of the synthesis, the stability upon reaction and the ability to process concentrated feeds. In this context, dealumination of beta followed by solid- or liquid-phase tin incorporation has been introduced as a less demanding strategy than the complex and long hydrothermal synthesis to prepare Sn-BEA.[11] The method to produce Sn-MWW, comprising deboronation of hydrothermally-synthesised B-MWW and subsequent tin incorporation, features a comparable degree of complexity.[8] Hydrothermally-synthesised Sn-MFI and alkaline-treated MFI hold the brightest prospects as industrially-amenable catalysts. Indeed, the analogously prepared TS-1 zeolite has been manufactured by EniChem for nearly three decades for use in phenol hydroxylation [12] and desilication of commercial ZSM-5 has already been demonstrated at the pilot scale.[13] Concerning catalyst robustness, Sn-MWW has been shown to retain its performance in three consecutive catalytic runs, while hydrothermally-prepared Sn-BEA and alkaline-treated MFI to suffer from poor recyclability due to metal leaching.[8,10] The reusability of Sn-MFI has not been evaluated. The replacement of the aqueous medium by an alcoholic solvent has proved effective in preventing deactivation. Thus, the production of alkyl lactates (AL), relevant green solvents and chemical intermediates, by one-pot DHA isomerisation-esterification has been often preferred.[7,11a] In relation to LA productivity, no attempts have been made to verify that these zeolites could retain their outstanding activity and selectivity when applying higher DHA concentrations than 3 wt.% in the starting solution.

At the *process* level, critical aspects in determining economic and environmental performance are the number of unit operations, the reaction conditions (temperature, pressure, solvent), the types of separation methods and the cost, availability and cradle-to-gate life cycle impact of raw materials and chemical auxiliaries. Only some of these aspects have been addressed and on an individual basis. For instance, in view of eliminating a separation step, it has been attempted to combine the production and isomerisation of DHA. Thus, Sn-BEA and Sn-MWW ha~~s~~ve been used in water or alcohols to isomerise GLU, cleave the generated fructose by a retro-aldol reaction and further convert the obtained DHA to LA/AL, while noble metals supported on oxides or (tin-containing) zeolites have been employed to oxidise glycerol (GLY) and isomerise DHA in one pot.[8,14] Still, even substantially reduced product yields have been attained compared to the one-step transformation and catalyst stability issues have emerged, thus questioning the advantage of these intensified processes. With respect to the bio-derived feedstock used, GLY has been regarded as a preferable source of DHA compared to sugars as it is an abundant waste (125 Mton expected in 2016) of biofuels production.[15] Furthermore, it can be enzymatically oxidised to the desired triose with very high yields (97% after 30 h of reaction) under acidic pH conditions, avoiding the generation of gypsum waste streams.[16] Based on these observations, it is clear that a quantitative assessment of the economic profitability and environmental sustainability of the new value chains for LA production would be extremely relevant in order to adequately define their industrial potential.[17] In this respect, evaluating the relative impact of variables such as catalyst stability and number of production steps is particularly important. Indeed, LA could be also recovered from alkyl lactates, whose production undermines the integrity of the Lewis-acid solids to a lower extent, through an additional hydrolysis step in water, in analogy to the biocatalytic route (Scheme 1).

Herein, we first evaluate the performance and the reusability of Sn-containing MFI zeolites in the isomerisation of concentrated (15 wt.%) DHA aqueous and methanolic solutions. We then perform rigorous process modelling and flowsheeting for a hybrid bio-/chemocatalytic route comprising enzymatic GLY oxidation and water- or alcohol-based DHA isomerisation and for the biochemical pathway to contrast the alternative approach to the current industrially-practiced technology. For this purpose, we combine the experimental results herein gathered on DHA isomerisation with available data for the fermentation of GLY to the ketonic triose and of sugar beet-derived GLU to LA. The comparison is made on the basis of economic and ecological metrics following a systematic life cycle assessment (LCA) methodology.[17]

2 Materials and methods

2.1 Experimental

Catalyst synthesis and characterisation

The Sn-containing MFI zeolites investigated in this study comprise Sn-MFI and MFI-ATSn. The former was prepared by hydrothermal synthesis in hydroxide medium and the latter by post-synthetic alkaline-assisted stannation of silicalite-1, according to reported procedures.[18,19] The catalysts were characterised in terms of porous and acidic properties, crystallinity, and tin and carbon content prior to and after use in the catalytic tests. A detailed description of the synthetic protocols and of the characterisation methods is available in the Electronic Supplementary Information (ESI).

Catalyst testing

Batch catalytic tests were performed under autogenous pressure in 100-cm^3 thick-walled glass vials (Ace, pressure tubes, front seal) dipped in an oil bath heated at 413 K. The vials were loaded with 7.6 g of DHA (Sigma-Aldrich, 98%, dimer), 4.12 g of catalyst and 40 g of solvent. The latter comprised deionised water or methanol (Scharlau, 99.9%). The mixture was allowed to react under vigorous stirring during 0.5 to 3 h. Thereafter, the reaction was quenched using an ice bath and the catalyst removed by filtration under reduced pressure using cellulose acetate filters (Satorius, 0.45 μm). Reusability tests in water and methanol were performed for 1 h under the same conditions. Prior to every reuse, the catalyst was calcined in static air at 823 K (5 K min^{-1}) for 5 h. Continuous-flow experiments were conducted using a homemade continuous-flow reactor setup composed of (*i*) an HPLC pump, (*ii*) a stainless-steel tubular reactor (Swagelok SS-T4-S-035, o.d. = ¼ inch, i.d. = 4.6 mm) heated by an oven and (*iii*) a backpressure regulator (Swagelok, LH2981001). The reactor was loaded with Sn-MFI (0.825 g) diluted with quartz (1.0 g, sieve fraction = 0.25-0.36 mm) and heated at 403-413 K. Thereafter, the reaction was started by admitting a solution comprising 5 wt.% DHA in methanol at a rate of 0.2 cm^3 min^{-1}. To prevent solvent loss by evaporation, the system was pressurised to 25 bar prior to heating. Samples were periodically collected from the outlet stream for analysis.

DHA and LA were isolated by high-performance liquid chromatography (HPLC) in a Merck LaChrom system equipped with a Biorad Aminex HPX-87H column heated at 308 K and a refractive index detector (Hitachi Chromaster model 5450) set at 303 K, using an aqueous eluent of 0.005 M H_2SO_4 (pH = 2.30) flowing at 0.6 cm^3 min^{-1}. Quantification was obtained by integration of their respective peaks. Methyl lactate (ML) was analysed using a gas chromatograph (HP 6890) equipped with an HP-5 capillary column and a flame ionisation detector. He was used as the carrier gas (flow rate = 4.3 cm^3 min^{-1}, pressure = 1.4 bar) and an injection volume of 0.2 μL was applied. The initial temperature of 328 K was held for 2 min before heating to 473 K (30 K min^{-1}). ML (Acros, 99%) was used as a reference and isooctane (Fluka, 99.5%) as a standard, which was added to the reaction mixture after catalyst removal. The DHA conversion was calculated as the mole of DHA reacted divided by the mole of DHA fed, whereas the LA/ML yields as the mole of LA/ML formed divided by the initial moles of DHA.

2.2 Process modelling

Rigorous process models were developed for racemic LA production based on the routes depicted in Scheme 1. The simulation software Aspen Plus® V8.2 was used for this purpose An overview of the models is presented in Table 1. In all cases, the main reactor units operate in batch mode, while the downstream purification units in steady-state continuous mode. For sake of clarity, the intermediate storage tanks, facilitating the connection between the batch and the continuous part, were not represented. The process waste streams are treated in conventional industrial waste water treatment plants or incineration units depending on their composition.[20] The basic features and respective flowsheets of each process model are described in the following sections whereas detailed information about the process conditions, amount of raw materials and of chemical auxiliaries, reaction conversions, production yields and separation procedures is available in the ESI.

Table 1 Overview of the process models developed for LA production

Model description	Acronym
Biocatalytic DHA production from pure GLY (99.99 wt.%)	*DHA-1*
Biocatalytic DHA production from crude GLY destined for the incineration plant[a]	*DHA-2*
Biocatalytic DHA production from crude GLY destined for the waste water treatment plant[a]	*DHA-3*
Biocatalytic DHA production from crude glycerol according to *DHA-3* followed by chemocatalytic LA production from DHA using the MFI-AT catalyst (3 consecutive runs) in water	*LA-1*
Biocatalytic DHA production from crude glycerol according to *DHA-3* followed by chemocatalytic LA production from DHA using the Sn-MFI catalyst (3 consecutive runs) in water	*LA-2*
Biocatalytic DHA production from crude glycerol according to *DHA-3* followed by chemocatalytic LA production from DHA using the Sn-MFI catalyst (7 consecutive runs) in water	*LA-3*
Biocatalytic DHA production from crude glycerol according to *DHA-3* followed by chemocatalytic LA production from DHA using the Sn-MFI catalyst (7 consecutive runs) in methanol	*LA-4*
Industrially applied biocatalytic LA production from GLU	*LA-5*

[a] The DHA-2 and DHA-3 models are identical, except for the distinct fate that crude GLY would follow if it was not used as a chemical feedstock.

Conventional LA production

This method is used by 90% of the LA producing companies, such as Cargill Inc. and Purac Biochem BV.[21] The process comprises five main steps: (*i*) the enzymatic generation of racemic LA from GLU with LA bacteria and its progressive neutralisation with $Ca(OH)_2$, (*ii*) the H_2SO_4-catalysed hydrolysis of the calcium lactate formed, (*iii*) LA esterification with methanol, (*iv*) the purification of the obtained ML by distillation and finally, (*v*) ML hydrolysis to LA with water.[22] The flowsheet of the *LA-5* process model is shown in Fig. S3 in the ESI.

Alternative LA production from GLY

With respect to the alternative production of LA by chemocatalytic isomerisation of DHA, it was necessary to create a process model for the production of DHA, since its price (*ca.* 4500 USD ton^{-1}) is significantly higher than that of LA (1700-2100 USD ton^{-1}) and the LA market has a greater volume than that of DHA.[23,24] As mentioned in the introduction, GLY fermentation was selected as the production method of the triose substrate.[25] The process comprises the following main steps: (*i*) fermentation of GLY (*ca.* 10 wt.% in water), (*ii*) concentration of the reaction mixture and crystallisation of DHA out of butanol and (*iii*) recovery of butanol from the water-containing mother liquor through heterogeneous azeotropic distillation. Two process models were developed using pure GLY (99.99 wt.%, *DHA-1*) and crude GLY (75.8 wt.%, *DHA-2,3*). The composition of the latter was derived from the study by Hansen *et al.*[26] Impurities comprise ash (5.3 wt.%), water (6.0 wt.%), matter organic non glycerol (MONG, 9.6 wt.%) and methanol (3.3 wt.%). The *DHA-2,3* models differ from the *DHA-1* model with respect to the additional treatment of crude GLY in order to remove ash, methanol and MONG prior to the fermentation. Their flowsheet is shown in Fig. 1 and the relative life cycle inventory (LCI) data are compiled in Table 2.

The product stream of the DHA manufacture was directed into the LA production process, which was conducted using either a Lewis-acid zeolite prepared by alkaline-treatment of a commercial Al-containing MFI sample (denoted as MFI-AT) or Sn-MFI. The performance (*i.e.*, productivity, recyclability, *etc.*) of the former catalyst is discussed in detail in [10] and that of the latter material in Section 3.1. Two process models were developed for racemic LA production over the two catalysts, which are based on: (*i*) the catalytic conversion of DHA to LA in water with acetol and pyruvic acid as the main by-products, (*ii*) the recovery of the catalyst by filtration and regeneration *via* calcination and *(iii-a)* the removal of impurities through distillations and crystallisation with butanol and the recovery of unreacted DHA for the process using MFI-AT or (*iii-b*) the concentration of the reaction mixture and the recovery of LA through vacuum distillation for the process using Sn-MFI. The different downstream work-up procedures originate from the distinct degree of conversion of DHA, which is partial over MFI-AT and full over Sn-MFI. The flowsheets for these processes are shown in Fig. S4 in the ESI (MFI-AT) and Fig. 2a (Sn-MFI) while the LCI data are compiled in Table 3 for both scenarios. The consumption of DHA is similar in either process as DHA is recovered if unconverted.

Replacing water with methanol in the Sn-MFI catalysed isomerisation resulted in some process flowsheet modifications. As DHA is fully transformed into ML generating negligible amounts of soluble by-products (see Section 3.1), ML and the excess of methanol can be recovered by distillation. ML is then hydrolysed *via* a low-pressure reactive distillation producing pure LA and a mixture of methanol and water, which is further separated by conventional distillation to recover the organic fraction. The choice of methanol instead of ethanol or butanol as alcoholic medium was mainly based on this last distillation step, which would be azeotropic instead of conventional when using the C_2 and C_4 alcohols. The respective process flowsheet is shown in Fig. 2b and the associated LCI data are presented in

Table 2 LCI data for the biocatalytic production of DHA from GLY

Materials and energy	*DHA-2,3*	Units
Crude glycerol	1.5	kg kg_{DHA}^{-1}
Process water	10.2	kg kg_{DHA}^{-1}
Catalyst	–	kg kg_{DHA}^{-1}
Nutrients	0.05	kg kg_{DHA}^{-1}
Air	3.1	kg kg_{DHA}^{-1}
Butanol	0.06	kg kg_{DHA}^{-1}
Steam (6 bar)	44.7	MJ kg_{DHA}^{-1}
Cooling water	818	kg kg_{DHA}^{-1}
Waste	11.0	kg kg_{DHA}^{-1}

Table 3 LCI data for the chemocatalytic production of LA from DHA in water

Materials and energy	MFI-AT, 3^a	Sn-MFI, 7^a	Units
DHA	1.1	1.2	kg kg_{LA}^{-1}
Process water	6.5	5.8	kg kg_{LA}^{-1}
Butanol	0.02	-	kg kg_{LA}^{-1}
Catalyst	0.3	0.1	kg kg_{LA}^{-1}
Steam (6 bar)	18.2	13.7	MJ kg_{LA}^{-1}
Cooling water	434	271	kg kg_{LA}^{-1}
Waste	6.9	6.1	kg kg_{LA}^{-1}

^{a} 3 and 7 indicate the number of cycles in which the catalyst was reused without loss in performance.

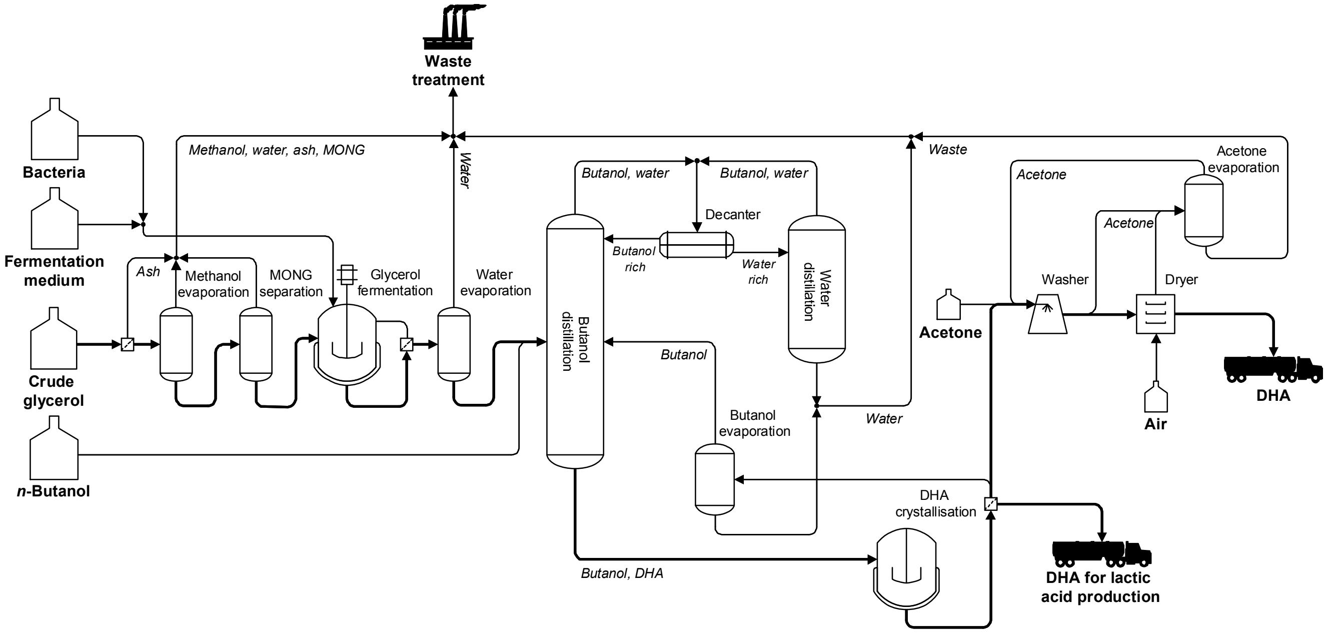


Fig. 1 Flowsheet for the DHA production process from crude glycerol according to the *DHA-2,3* models.

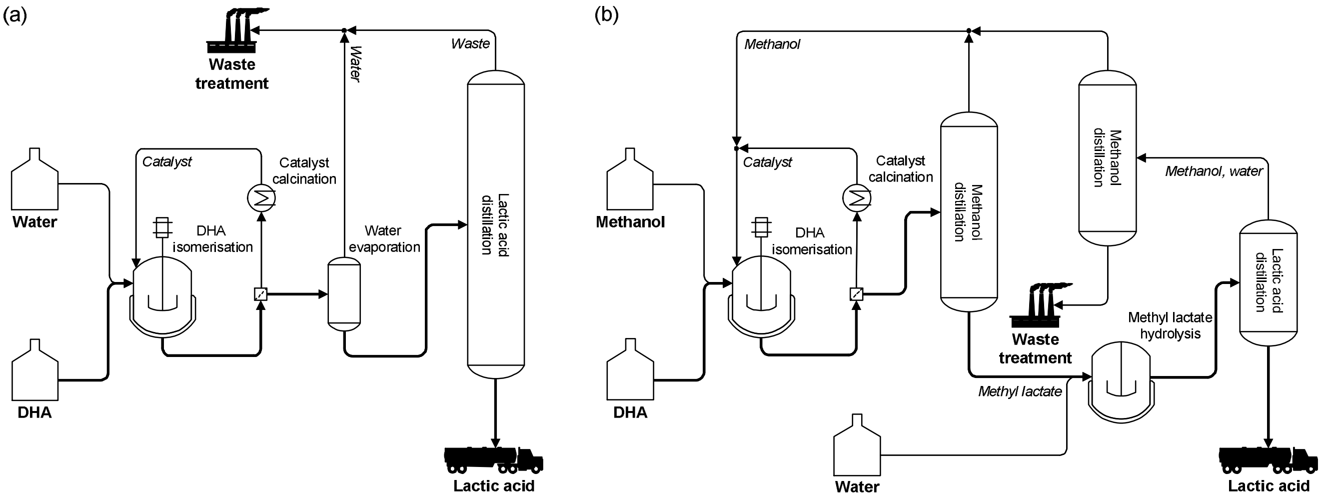


Fig. 2 Flowsheet for the chemocatalytic LA production process from DHA using the Sn-MFI catalyst in (a) water and (b) methanol.

Table 4 LCI data for the chemocatalytic production of LA from DHA in methanol

LCI data	Sn-MFI, 7[a]	Units
DHA	1.09	kg kg$_{LA}^{-1}$
Process water	0.41	kg kg$_{LA}^{-1}$
Methanol	0.03	kg kg$_{LA}^{-1}$
Catalyst	0.10	kg kg$_{LA}^{-1}$
Steam (6 bar)	12.7	MJ kg$_{LA}^{-1}$
Cooling water	649	kg kg$_{LA}^{-1}$
Waste	0.59	kg kg$_{LA}^{-1}$

[a] 7 corresponds to the number of cycles in which the catalyst was reused without loss in performance.

Table 4. Because of the recycling of methanol, less waste is generated but a higher cooling utility consumption is required compared to the process using MFI-AT in water. DHA, catalyst and steam consumption are equivalent in the LA production scenarios using Sn-MFI in water or methanol.

2.3 Environmental and economic assessment

The goal and scope of the environmental assessment following the LCA methodology is the comparison of the alternative process options presented in Scheme 1 for the production of 1 kg of LA. The process models were used for estimating the relevant LCI data (*i.e.*, consumption of resources and process emissions) in a cradle-to-gate approach. The respective environmental impacts in the use phase and the environmental fate of LA after its use were not considered here (*i.e.*, no cradle-to-grave impacts were calculated). The materials for the plant construction were also not included. Three well-known LCA metrics were followed, *i.e.*, the cumulative energy demand (CED), the global warming potential (GWP-100a) and the eco-indicator 99 (EI99). These metrics have been reported in previous chemical process assessment studies to cover different aspects of the environmental impacts.[27] In the case of CED, we calculated here the non-renewable CED to focus on the depletion aspect for resource protection, which is the main motivation for a bio-based production. A more detailed discussion on this topic can be found elsewhere.[28]

For the allocation of environmental impacts to crude GLY as a co-product of the biodiesel production, the approach and assumptions proposed by Weidema *et al* were followed.[29] Thus, crude GLY was considered as a partially utilised co-product from a consequential LCA perspective. Therefore, the respective LCA metrics calculated in this study are valid as long as the amount of GLY employed in LA production is not restricted by the biodiesel production. In the case of pure GLY, the values allocated in the Ecoinvent database were used.[30] Due to the lack of literature data, the impact of the separation of MONG *via* an acidulation process [31] on the LCA assessment was not considered.

The economic assessment was only based on operating costs. Investment costs are outside of the scope of the present study, since the production plants were not optimised for a specific capacity. Although no capital cost estimations are provided, all flowsheets have similar LA productivity. All background data for the environmental and economic assessment are provided in the ESI.

3 Results and discussion

3.1 Tin zeolites for LA production

The Sn-containing zeolites prepared for LA production have a rather similar tin content (1.2 and 1.6 wt.% for Sn-MFI and MFI-ATSn, respectively). Their mesoporous surface is also comparable (Table 4), but, owing to the different preparation procedures, the hydrothermally-synthesised zeolite is exclusively microporous while the solid obtained by alkaline-assisted stannation also features intracrystalline mesopores.[3~~0~~2] Additionally, the structural characteristics and placement of the catalytic tin centres in the samples are distinct, *i.e.*, tin is fully integrated in the framework and uniformly distributed in the whole volume in the former, while it is totally or partially incorporated in the framework and predominantly at the external surface of the crystals in the latter.[32]

These materials were evaluated in DHA (15 wt.%) isomerisation at 413 K in water and methanol. Based on the kinetic profiles collected, the LA yield rapidly augmented with time attaining 85% at 97% DHA conversion for Sn-MFI (Fig. 3a) and 75% at 95% DHA conversion for MFI-ATSn (Fig. 3b) after only 30 min. Increasing the reaction time to 1 h enabled to convert the remaining DHA as well as the glyceraldehyde and pyruvaldehyde intermediates to the desired product, thus reaching LA yields of nearly 100% for both catalysts. LA was stable in the reaction mixture since its concentration remained unaltered for longer reaction times. Equivalent profiles were observed for the methanol-based isomerisation, the 1-h ML yield being only slightly lower for MFI-ATSn (*ca*. 93%) compared to Sn-MFI (*ca*. 98%) (Fig. 3c,d). The outstanding performance of Sn-MFI compared to previous reports [7b,9] indicates that optimal conditions were applied to the DHA isomerisation reaction. The similar catalytic behaviour of the stannated zeolite catalyst suggests that, as observed for the diffusion-free glyoxal substrate,[1~~5~~8] the advantage of a greater accessibility of tin sites compared to the hydrothermally-prepared sample is counterbalanced by the negative impact of their moderate structural heterogeneity. Investigation of the used catalysts by N_2 sorption and CHN analysis (Table 4) revealed substantial depletion of the porous volume due to the deposition of carbon-containing species (6 and 3 wt.% C on either catalysts in water and methanol, in the order). Accordingly, upon evaluating the stability of the solids in repeated 1-h runs, a calcination step was introduced prior to each reuse. For sake of clarity, Fig. 4 only displays the results of the first, second and last test (the complete set of data can be found in Fig. S1 in the ESI). The catalytic performance of Sn-MFI was unaltered in methanol, while a slight deactivation was observed in water. A similar behaviour was observed for MFI-ATSn. Average LA yields of 93 and 99% and ML yields of 98 and 97% at full DHA conversion were determined for the hydrothermally-synthesised and the stannated samples, respectively. As indicated by the reduced tin content and amount of Lewis-acid sites (measured by elemental analysis and infrared spectroscopy of adsorbed pyridine, respectively) and by the retained crystallinity (determined by X-ray diffraction) of the recycled materials, the activity loss was related to metal leaching (Fig. 4 and Fig. S2 in the ESI). As mentioned above, this phenomenon has been reported as the main cause of deactivation of catalysts in DHA isomerisation, usually being more pronounced in aqueous than in alcoholic media. Even if very moderate and undetrimental for the catalyst recyclability, our results follow the same trend (*i.e.*, −10 and 17 % in water and −3 and 10% in methanol for Sn-MFI and MFI-ATSn, respectively).

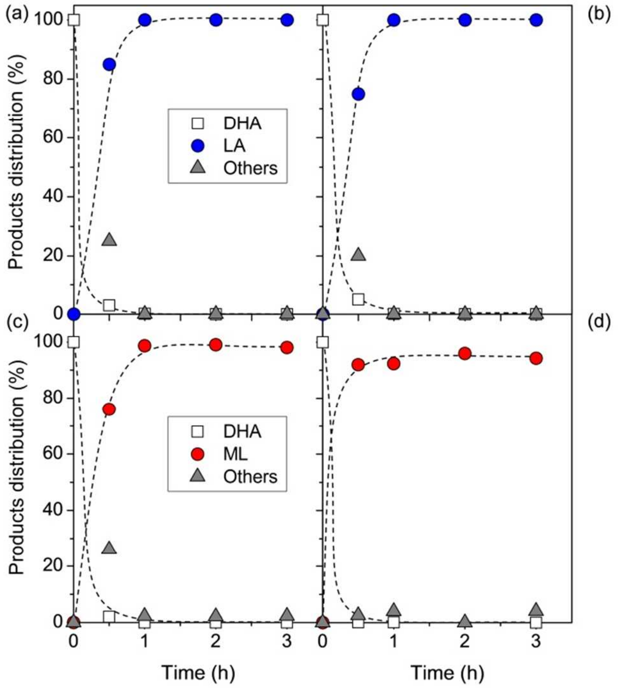


Fig. 3 Concentration profiles of reactant and products during the conversion of 15 wt.% DHA at 413 K over (a) Sn-MFI and (b) MFI-ATSn in water and over (c) Sn-MFI and (d) MFI-ATSn in methanol. "Others" include glyceraldehyde and pyruvaldehyde as well as acetic acid or pyruvaldehyde dimethyl acetal for the reaction in water or methanol, respectively.

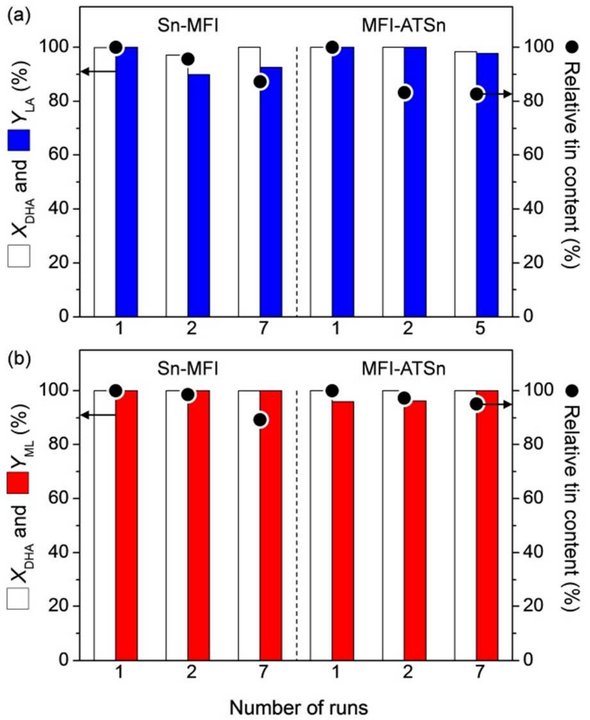


Fig. 4 Reusability tests for Sn-MFI and MFI-ATSn in (a) water and (b) methanol. The white bars refer to the DHA conversion while the blue/green bars identify the LA/ML yields.

Table 4 Porous properties of fresh, used and regenerated zeolite catalysts

Catalyst	State	S_{meso} ^{a} (m^2 g^{-1})	V_{micro} ^{a} (cm^3 g^{-1})	V_{pore} ^{b} (cm^3 g^{-1})
Sn-MFI	Fresh	49	0.17	0.24
	Used, H_2O	87	0.11	0.27
	Reg., H_2O	47	0.16	0.26
	Used, MeOH	39	0.12	0.23
	Reg., MeOH	52	0.16	0.28
MFI-ATSn	Fresh	50	0.11	0.22
	Used, H_2O	77	0.13	0.27
	Reg., H_2O	101	0.12	0.27
	Used, MeOH	62	0.11	0.22
	Reg., MeOH	63	0.13	0.25

^{a} Determined by the *t*-plot method. ^{b} Volume adsorbed at $p/p_0 = 0.99$.

The origin of tin leaching has not been elucidated at a molecular level so far. Nevertheless, the stability of tin-containing MFI zeolites in the isomerisation of glyoxal in water and methanol and the substantial metal loss observed in the isomerisation of sugars in either of the solvents [32] suggest that the extraction of tin from the materials is better aided by chelating functional groups of the substrate/product than by the acidity of the product. With respect to the stabilisation offered by the matrix in which tin is contained, our MFI zeolites appear strongly superior to bifunctional carbon-silica composites and BEA zeolites.[7c,8] Thus, in light of their performance and of the scalability of their synthesis, they stand as promising candidates for large-scale DHA isomerisation.

3.2 Process assessment

The results of the environmental assessment according to the CED metric are presented in Fig. 5a for the conventional technology and the various alternative hybrid process models of Table 1. The conventional enzymatic process (*LA-*) requires 109 $\mathrm{MJ_{eq}}$ to produce 1 kg of LA from GLU and the biggest environmental burden is created by the gate-to-gate energy utility consumption of the process. With respect to the alternative cascade processes, the single oxidative fermentation of pure GLY to DHA (*DHA-1*) was found to have a high environmental impact (102 $\mathrm{MJ_{eq}\ kg_{LA}}^{-1}$) due to the large CED allocated to the raw material (35 $\mathrm{MJ_{eq}\ kg_{LA}}^{-1}$). This evaluation was refined taking into account that, since crude GLY is a waste, its purification and use as a chemical feedstock effectively competes with its disposal, which is typically conducted *via* incineration or treatment in a waste water treatment plant. Because of the steam and electricity generated upon burning GLY, the CED value was slightly higher in the first case (108 $\mathrm{MJ_{eq}\ kg_{LA}}^{-1}$, *DHA-2*), while preventing the consumption of energy and materials upon treatment in a waste water plant substantially reduced the CED value in the second case (65 $\mathrm{MJ_{eq}\ kg_{LA}}^{-1}$, *DHA-3*). Hence, the most sustainable *DHA-3* model was selected for the assessment of the environmental footprint of the combined bio- and chemocatalytic process. In this respect, the effect of the type of catalyst and of its recyclability was studied besides the nature of the solvent. Accordingly, the use of MFI-AT (*LA-1*) was compared to that of Sn-MFI (*LA-2*), considering a lifetime of 3 catalytic runs in both cases. CED values of 121 and 114 $\mathrm{MJ_{eq}\ kg_{LA}}^{-1}$ were respectively obtained. The incomplete conversion of DHA over the tin-free catalyst penalises the process by 7 $\mathrm{MJ_{eq}\ kg_{LA}}^{-1}$ as it imposes additional unit operations to separate DHA from LA as well as to recover the butanol used in these steps. When the Sn-MFI zeolite was recycled for 7 instead of only 3 times, the CED value decreased by 11 $\mathrm{MJ_{eq}\ kg_{LA}}^{-1}$ due to the reduced materials consumption (*LA-3* and *LA-2,* respectively). Therefore, the reusability of the catalyst has a slightly higher impact than its activity (*i.e.*, DHA conversion) on the environmental footprint in the ranges herein investigated. It is worth noting that the latter process becomes competitive with respect to the traditional technology. When methanol was used as a medium for the DHA conversion (*LA-4*), the CED value of the cascade process was estimated at 96 $\mathrm{MJ_{eq}\ kg_{LA}}^{-1}$, which is even moderately inferior to the case of GLU fermentation. The main reason for the alleviated environmental impact is the lower energy spent to recover methanol after reaction compared to the energy needed to concentrate the aqueous reaction mixture through water evaporation. Similar overall trends were observed for the GWP and EI99 LCA metrics, which are presented and discussed in the ESI. It is expected that the assessment of LA production processes using MFI-ATSn would lead to equivalent results to the case of Sn-MFI in view of the identical performance and the virtually similar cost of preparation of the two catalysts.

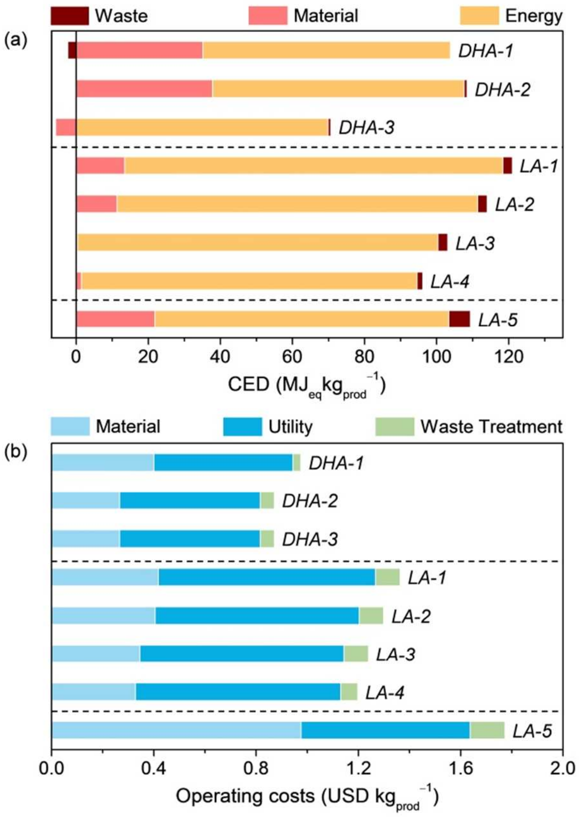


Fig. 5 (a) Cradle-to-gate LCA according to the non-renewable CED metric and (b) operating costs for various process layout scenarios.

The same process layouts and scenarios were evaluated in terms of operating costs as an economic metric. The results of this study are shown in Fig. 5b. The conventional LA production (*LA-5*) was found to have the highest operating cost with 1.77 USD $\mathrm{kg_{LA}}^{-1}$. This originates from the higher price of GLU (400 USD ton^{-1}) compared to GLY (270 USD ton^{-1} for pure GLY and 110 USD ton^{-1} for crude GLY). Therefore, *ca*. 50% of the operating costs are allocated to the material in the conventional process. Considering the cascade bio- and chemocatalytic processes, the use of MFI-AT (*LA-1*) rather than Sn-MFI (*LA-2*) leads to a more costly technology (1.36 *versus* 1.30 USD $\mathrm{kg_{LA}}^{-1}$, respectively, due to the more energy-intensive purification dictated by the incomplete conversion of DHA. The cascade process using the tin-containing catalyst becomes more economical by 0.06 USD $\mathrm{kg_{LA}}^{-1}$ upon its reuse for 7 times (1.24 USD $\mathrm{kg_{LA}}^{-1}$, *LA-3*) as the material cost is contained. Based on these findings, it appears that a 12% higher LA yield produces the same impact as a doubled catalyst reusability on the operating costs. The production of LA through the methanol-based isomerisation (*LA-4*) is more economical by 3% than the best process performed in water resulting in 1.20 USD $\mathrm{kg_{LA}}^{-1}$. This value is *ca*. one third smaller compared to the conventional LA production. Finally, considering an LA price of 1800 USD ton^{-1},[23] the marginal profit of LA production *via* DHA isomerisation in methanol is 15 times higher than that of the conventional process (0.61 *versus* 0.04 USD kg$_{LA}^{-1}$).

Since utilising crude instead of pure GLY is one key factor for improving the economic and environmental performance of the bio/chemocatalytic process, a deeper investigation on the influence of the composition of crude glycerol (Table S6 in the ESI) on the assessment was performed. Accordingly, additional scenarios were evaluated in which the methanol and ash contents were separately increased at the expense of water, keeping the GLY and MONG contents constant. Furthermore, a limit case was considered in which the GLY content corresponds to the lowest value given in ref. [26], *i.e.*, 38 wt.%, and the difference to the standard amount was filled by an increase in methanol. In all cases, crude GLY disposal in waste water treatment plants was still considered as the more significant alternative option. The results of the assessment (Figs. S8-11 in the ESI) reveal that a higher ash or methanol content negligibly impacts the process with respect to both the environmental and economic footprint, while a low GLY-high methanol feedstock determines two interesting effects. Thus, the impact of treating the significantly larger process waste generated is counterbalanced by the avoided impact of treating a low GLY-high methanol stream, which ultimately ensures that the *LA-3* and *LA-4* processes remain environmentally more attractive than the conventional case (net CED values of 108 and 101 *versus* 109 MJ$_{eq}$ kg$_{LA}^{-1}$, Fig. S8). In contrast, since this compensating effect does not apply to the economic assessment, the operating costs are increased by *ca.* 30%. This implies a still less costly process than the enzymatic route, but a reduced margin profit-wise. It should be noted that in this sensitivity analysis it was assumed that the price of crude GLY does not to depend on its composition. Thus, as low-purity GLY will possibly have a lower price, the effective increase in operating costs should be more contained than in the current study. A more detailed discussion of these results can be found in the ESI.

3.3 Towards a continuous process

Since a continuous operation is widely preferred for the manufacture of bulk chemicals and would make the alternative LA production process even more advantageous, the possibility to extrapolate the catalytic systems herein studied for DHA isomerisation from a batch system to a continuous-flow reactor was explored. Only one contribution so far reports the attempt to produce LA and ML continuously.[33] Therein, a USY zeolite has been tested at 430 K and with a WHSV of 0.16 kg$_{DHA}$ kg$_{cat}^{-1}$ h^{-1} and has been found to rapidly deactivate in water but to retain *ca.* 90% of its activity after 48 h on stream in methanol. In our case (Fig. 6), Sn-MFI was evaluated under industrially more relevant conditions, *i.e.*, at a lower temperature (383 K) and a nearly 4-fold higher WHSV (0.60 kg$_{DHA}$ kg$_{cat}^{-1}$ h^{-1}). An ML yield of *ca.* 68% was observed in the first few hours of the test, which decreased by *ca.* 20%after 24 h. The apparently more limited stability of our tin-containing catalysts compared to the USY zeolite is likely due to an enhanced adsorption of C-based species (4 wt.% C in the used catalyst) due to the *ca.* 3 times more concentrated DHA solution. Since calcination was effective in restoring the initial catalyst properties (Fig. 4b), a strategy to enable a continuous LA production in the future could comprise the use of series fixed-bed reactors, alternatively in use/under regeneration.

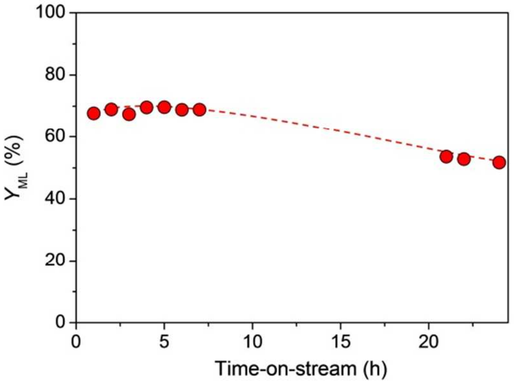


Fig. 6 ML yield during a continuous-flow experiment over Sn-MFI at 383 K.

4. Conclusions

Environmental, energetic and economic aspects of the currently applied enzymatic route for lactic acid production and of an alternative route comprising the biocatalytic oxidation of glycerol to dihydroxyacetone followed by chemocatalytic isomerisation have been herein evaluated by LCA using CED, EI99, GWP and operating costs as indicators. For the triose isomerisation step, the modelling was based on catalytic data herein gathered which uncover Sn-containing MFI zeolites prepared by either hydrothermal synthesis or alkaline-assisted stannation as outstandingly active, selective and reusable Lewis-acid catalysts.

The novel cascade process was shown to have substantially lower operating costs and a comparable till milder environmental and energetic footprint compared to the traditional technology. The economic aspects were found to be mainly determined by the costs associated with the raw materials and the sustainability to be critically influenced by both raw materials and process characteristics. In this respect, the high impact of the large CED of biodiesel-derived crude glycerol can be modulated taking into account the energy flows and material consumptions of the disposal procedures required if it is not used as a chemical feedstock. Additionally, considerable energy savings can be achieved employing a highly performing and recyclable Lewis-acid catalyst in the dihydroxyacetone isomerisation reaction. Accordingly, the Sn-containing MFI zeolites herein studied are preferable to Sn-free MFI zeolites previously reported. Interestingly, the one-step chemocatalytic manufacture of lactic acid in water is slightly energetically and economically more intensive than the two-step process based on methyl lactate production followed by hydrolysis due to the lower energy demands of the different downstream separation and purification procedures. Both process scenarios remain greener and economically more attractive than the conventional route even if the purity of the crude glycerol feedstock is as low as 38 wt.% .

Overall, the cascade bio-/chemocatalytic production of lactic acid from glycerol appears to have a strong potential for industrial implementation. In this direction, it is envisaged that the application of a continuous process for dihydroxyacetone isomerisation, which was herein demonstrated, and the possibility of obtaining the triose intermediate from glycerol through a more productive chemocatalytic route, preferably also operated continuously, would render this alternative two-step process even more appealing.

Finally, this study emphasises the importance and potential of combining experimental results, rigorous process modelling and multi-criteria analysis for assessing the sustainability performance of proposed (bio)chemical production processes.

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Notes and references

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† Electronic Supplementary Information (ESI) available: Catalyst synthesis and characterisation methods; reusability tests of the catalysts in water and methanol and tin content in the solids after every run; X-ray diffraction and infrared spectroscopy of adsorbed pyridine of MFI-ATSn prior to and after LA production; process information for modelling the conventional biocatalytic LA production and the alternative chemocatalytic LA production from DHA; environmental assessment of the process models according to GWP and EI99; economic assessment of the process models according to profit calculations; background data for the environmental and economic assessment; See DOI: 10.1039/b000000x/

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