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Kinetic and process modeling of Guerbet coupling chemistry over Cu–Mg–Al mixed oxides†

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Guerbet coupling chemistry is a route to oligomerize ethanol into C₄₊ alcohols. Long chain ethers can be obtained through bimolecular dehydration of these alcohols. Ethers generated from the dehydration of C₆₊ alcohols produce a fuel that satisfies diesel engine requirements, therefore selective production of C₆₊ alcohols is of particular interest. The desired hexanol is synthesized through ethanol and butanol coupling, accompanied by the formation of undesired products through several reaction pathways. In this work the coupling of ethanol and butanol has been studied over Cu_{0.01}Mg_{2.99}AlO_x to produce C₆₊ alcohols through Guerbet coupling reactions. Two series of catalytic tests were performed at 325 °C and 300 psig by using either pure ethanol feed or a cofeed ethanol–butanol 70–30 mole%. A kinetic model was developed to predict the product distribution over a wide range of contact times. Kinetic parameters were regressed by coding a routine that included a solution of differential mole balances embedded in an optimization problem. The herein developed kinetic model was integrated in a process simulation flowsheet that models the upgrading of ethanol into C₆₊ oxygenates. The butanol cofeeding strategy in the simulations was approached by recycling the produced butanol into the coupling reactor. The simulation results reveal that cofeeding butanol into the Guerbet reactor enhances initial production rates of C₆₊ alcohols, at the expense of fostering production of byproducts from butanol self-coupling. A maximum carbon yield of 82.2% for C₆₊ diesel fuel precursors can be obtained by minimizing the byproduct production after introduction of a hydrogenation reactor.

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Broader context

Diesel fuel demand is forecasted to increase in the upcoming decades driven by transportation of commodities and materials. Electrification of the heavy-duty transportation sector is not a promising solution to supply power to the heavy-duty fleet due to technical limitations. Thereby, investigation on synthesis of sustainable diesel fuel is critical to support economic development with low environmental burden. A technology based on catalytic upgrading of ethanol has been demonstrated as viable to produce diesel fuel. Nevertheless, the current state of this technology requires a better description of the reaction units which is achieved through development of kinetic models. These models not only provide an accurate assessment of the process's feasibility but also identify operational regimes that enhance performance. By guiding process optimization and improving efficiency, the insights from these models contribute to the development of a more sustainable diesel production route.

1. Introduction

Ethanol is the most produced liquid biofuel in the world, with an estimated global production of 29.6 billion gallons in 2023.¹ In the US, ethanol is typically blended with gasoline at ~10% v/v, to produce a mixture that exhibits a higher octane number than gasoline, improving the engine energy efficiency.^{2,3} Other factors like energy security and climate protection have also served as motivation for ethanol blending.⁴ The electrification of vehicles has been promoted as an approach to decarbonize the fuel industry. By 2030, 11 to 63% of the total passenger car sales in the world are forecasted to be electric vehicles, with percentages

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climbing up to 100% by 2050.⁵ Unfortunately, electrification of the heavy-duty transportation fleet is not expected to occur to the same extent due to technical barriers related to the size and weight of batteries. Moreover, regulations to reduce accelerated wear and tear of roads due to transportation of heavy loads hinder the electrification of trucks in the mid-term.⁵ Therefore, it is imperative to find ways to produce sustainable diesel fuel surrogates to stay aligned with decarbonization of the transportation sector. The demand for gasoline is projected to decrease due to the increase of personal electric vehicles,⁶ reducing the consumption of bioethanol for blending and potentially creating an ethanol surplus in the market. Synthesizing heavy distillate-range fuels through ethanol upgrading represents a market opportunity that the bioethanol industry needs to maintain its participation in the energy sector.⁷

Recently, our group developed a three-stage process to produce diesel fuel^{8,9} through catalytic upgrading of ethanol that encompasses the: (1) catalytic oligomerization of ethanol into longer chain alcohols through Guerbet coupling chemistry,^{10,11} (2) hydrogenation of byproducts generated in the first reactor (mainly esters, aldehydes, and ketones) into alcohols,¹² and (3) bimolecular dehydration of the alcohols mixture over an acid catalyst to produce a blend of long-chain ethers.^{13,14} We initially used hydroxyapatite (HAP) as the catalyst for ethanol conversion,^{15,16} but later switched to a Cu/MgAlO_x catalyst.¹⁰ The advantages of the Cu–Mg–Al catalyst compared to the HAP catalyst are higher product selectivity to C₆₊ alcohols and lower inhibition to water. Process modeling, life cycle assessment, techno-economic analysis, and fuel performance demonstrations were carried out by our collaborators to show the competitiveness of this technology.⁸ Despite the meaningful progress made in studying this diesel synthesis technology, there exists strong assumptions in the modeling of the process that constrains its global optimization. For instance, all our studies have contemplated the use of yield rather than kinetic reactors for each catalytic stage due to the inexistence of models applicable to our catalysts. The development of kinetic models under relevant reaction conditions for our reactors will reduce the stiffness imposed by yield reactors, allowing us to perform more realistic modeling of our synthesis process. Process design follows the hierarchical diagram depicted in Scheme S1 (ESI[†]), where the reaction units have the highest priority, with upstream being more important. The Guerbet coupling reactor has the highest priority in our process as it is the first reaction unit. Therefore, making efforts to describe it more accurately is imperative to continue validating the economic feasibility of our technology. As such, derivation of a kinetic model for the Guerbet coupling reactor is one of the goals of this report.

To the best of our knowledge, six previous attempts to model Guerbet coupling chemistry over HAP^{17–20} and mixed metal oxides^{21,22} have been published in the literature. There are two kinetic models where the kinetic constants were derived using the initial rates method to fit experimental information to rate expressions. Ho *et al.*¹⁷ studied and modeled the ethanol coupling chemistry over HAP, which follows the aldol condensation mechanism. Initial rates were determined by performing

reactions in the kinetic regime between 300–340 °C in flow reactors. Their rate function follows a Langmuir adsorption, where ethanol is specified as the most predominant surface species. The model presented by the authors exhibits an excellent ability to explain the observed consumption and formation rates of ethanol, acetaldehyde, and butanol within the studied range of temperature. In another study, Young *et al.*¹⁸ investigated the global ethanol coupling reaction to butanol over TiO₂, HAP, and MgO. However, their work developed solely rate expressions for acetaldehyde consumption in the aldol condensation reaction (2 acetaldehyde → crotonaldehyde + H₂O). While their model accurately describes the acetaldehyde consumption rate as a function of its partial pressure, the formulation of a rate expression for a single elementary reaction step does not provide sufficient information to model the full pathway from ethanol to butanol. Lastly, Scheid *et al.*²¹ investigated the ethanol coupling reaction over Mg–Al mixed oxides, with the Mg/Al ratio equals 5. This catalyst is the one that most resembles ours, with the caveat that ours also includes copper to facilitate (de)hydrogenation steps.¹⁰ Unlike the aforementioned studies, Scheid *et al.* did not focus only on studying the upgrading of ethanol into butanol, but rather on a more complex reaction network involving alcohol coupling, dehydrogenation, and dehydration reactions for ethanol and butanol. In their approach, their reaction network is composed of six reversible reactions that follow elemental reaction orders, with inclusion of adsorption, reaction, and desorption steps to yield a LHHW model. Since their study involves multiple reactions occurring simultaneously over the surface of the catalyst, Scheid *et al.* have approached the kinetic parameters regression problem by solving a differential-algebraic optimization problem to predict the observed mole fraction distribution of the products in the outlet of their flow reactor. Although all these three kinetic models are limited to C₂–C₄ chemistry, useful insights about the shape of the rate expression can be obtained for derivation of more complicated models.

Three more studies modeling a more complicated reaction network for ethanol coupling chemistry are available in the literature. In these papers the authors performed initial experiments to establish a reaction mechanism as a basis for their kinetic modeling. For instance, Tsuchida *et al.*²⁰ performed studies in continuous flow to investigate the alcohol coupling chemistry of ethanol in conversions up to 80% over HAP. Their catalytic observations yielded a reaction network composed of 13 reactions and served as a basis to derive their kinetic model. In their simulations, they predict the outlet concentration of C₂–C₁₀ linear and branched alcohols and olefines by solving a differential model as a function of the residence time for temperatures ranging between 300–450 °C. Recently, Eagan *et al.*¹⁹ made a great effort modeling the Guerbet coupling chemistry over HAP, explaining coupling, interrupted coupling, and dehydration reactions. One of the strengths of this model is its ability to simulate the catalyst inhibition as a function of the amount of water present in the system. The authors' model employs only nine kinetic parameters to describe 647 reactions involving 185 alcohols. Eagan *et al.* considered the existence of acid and basic sites following what is observed from a mechanistic



the pH remained at 10 during the synthesis. Temperature and pH were monitored continuously by employing a temperature/pH probe (Ohaus ST20). After complete addition of solution 1, the mixed solution was transferred into a Schott bottle, capped, and aged over 24 h at 60 °C with magnetic stirring. The aged solution was then vacuum filtered and washed with hot deionized water to remove residual sodium and nitrate ions. Subsequently, the filtered cake was redissolved in 300 mL of a 1 M solution of Na₂CO₃, filtered, and washed with 500 mL of DI. Filter cake was dried overnight at 110 °C in an oven (Lab-line, 3511), crushed into a powder, and calcined for 2 h at 600 °C under a heating ramp of 4 °C min⁻¹ in a muffle furnace under a static air atmosphere.

2.2 Alcohol coupling reactions

Gas-phase alcohol coupling reactions were carried out in stainless-steel packed bed reactors (316 SS, 1/4" O.D., 16" length), where the catalyst mass ranged between 40–1500 mg, depending on the studied contact time. Ten catalytic runs per feed were carried out at contact times (τ) ranging between 0.21–21.6 h kg_{cat} kmol_{gas}⁻¹. The synthesized Cu_{0.01}Mg_{2.99}AlO_x catalyst was sieved to ensure a uniform particle size distribution (standard sieve mesh 30–80), packed, and held at the center of the tubing by glass wool (Acros organics) and silica beads (Sigma-Aldrich) at both ends. The tubular reactor was then confined within an aluminum heating block, equipped with a K-type thermocouple, to keep an isothermal profile along the reaction bed. An electric furnace (Thermo Fisher, Lindberg blue M Mini-Mite) was employed to heat up the reactor and control the temperature program. Prior to catalytic tests, the catalyst bed was reduced *in situ* at 325 °C by flowing H₂ (Airgas, UHP grade) at 50 mL min⁻¹ overnight (more than 12 h), with a heating ramp of 4 °C min⁻¹.¹⁰ Posterior to the reduction of the bed, the tubular reactor was pressurized up to 300 psig of hydrogen atmosphere, and pressure was maintained constant by using a back pressure regulator (Equilibar, ZF0SVN8) with nitrogen as the pilot fluid. Alcohol coupling reactions were performed at 325 °C and 300 psig by pumping either pure ethanol or a blend of ethanol–butanol of composition 70–30% in mole basis at 65 μL min⁻¹ into the reactor by using a syringe pump (Teledyne ISCO). Hydrogen was used as a carrier and co-fed into the reactor at an adjusted rate so that the mole ratio of ethanol + butanol to hydrogen in the inlet of the reactor was 4 : 1. Hydrogen is cofeed into the reactor along with the alcohols to partially suppress the formation of esters and aldehydes inherent to the alcohols coupling chemistry.¹¹ Before contacting the catalyst, the liquid feed was passed through a preheated section maintained at 260 °C to ensure complete vaporization of the alcohols. Posterior to the reaction, reaction products were crashed on 10 mL of 1-propanol (96566, Sigma-Aldrich), pre-loaded in a removable glass condenser (110 mL, Ace glass), submerged in a dry ice bath (dry ice/acetone). Uncondensed products left the collection vessel and were directed towards a three-way valve for online analysis in a GC (Shimadzu 2010), equipped with flame ionization and thermal conductivity detectors, or venting. Condensed products were withdrawn every 40 minutes (~2 g of reaction product) up to the collection of

2.1 Catalyst synthesis

A 0.3 wt% copper magnesium aluminum mixed metal oxide catalyst ($\text{Cu}_{0.01}\text{Mg}_{2.99}\text{AlO}$) identified previously¹⁰ as a promising catalyst for ethanol oligomerization reactions has been derived from a hydrotalcite type precursor through the co-precipitation method.^{10,23} Initially, three aqueous solutions were prepared as follows: (1) 240 mL of a solution containing Mg^{2+} 0.843 M ($\text{Mg}(\text{NO}_3)_2 \cdot 6\text{H}_2\text{O}$, Sigma-Aldrich 237175), Al^{3+} 0.282 M ($\text{Al}(\text{NO}_3)_3 \cdot 9\text{H}_2\text{O}$, Sigma-Aldrich 237973), and Cu^{2+} 0.002 M ($\text{Cu}(\text{NO}_3)_2 \cdot 3\text{H}_2\text{O}$, Sigma-Aldrich 61194); (2) 300 mL of a solution 0.225 M of Na_2CO_3 (Sigma-Aldrich 223530); and (3) 500 mL of a solution 1.35 M of NaOH (Sigma-Aldrich 795429). Solutions were prepared in such a way that in the whole synthesis the ratio $(\text{Cu} + \text{Mg})^{2+} : \text{Al}^{3+} : \text{CO}_3^{2-} : \text{OH}^-$ was 3:1:1:10. First, solution 2 containing Na_2CO_3 was transferred into a 1 L beaker and heated up to 60 °C on a stir plate, while being stirred with a magnetic stir bar. Next, solution 1 and 3 were added in a dropwise fashion to the 1 L beaker by using syringe pumps. Flowrates in the syringe pumps were controlled so that the addition occurred in 1 h and

four samples per catalytic run. Liquid products were then prepared for gas chromatography analysis by diluting the samples in 1-propanol and adding a known amount of 1-pentanol (~1 wt%, Sigma-Aldrich, 138975) as the internal standard. Samples' analysis was conducted in a GC-FID (Shimadzu 2014), equipped with an RTX-VMS (Restek) column. Identification of reaction products was helped by GC-MS (Shimadzu GCMS-QP2010) and quantification of product concentration was done by using external standards when available.

Liquid phase concentration of product i was calculated with eqn (1), where $C_{\text{pentOH}}^{\text{vial}}$ is the concentration of the internal standard in the analysis vial, $\text{RF}_{x,i}$ stands for the relative response factor of compound i with respect to the internal standard, Area_i corresponds to the area of peak i recorded in the gas chromatograph, $m_{\text{propOH}}^{\text{condenser}}$ and ρ_{propOH} are the mass and density of 1-propanol employed as the crashing solvent and preloaded in the collection condenser, $m_{\text{product}}^{\text{condenser}}$ and $\rho_{\text{product}}^{\text{condenser}}$ represent the mass and liquid phase density of the reaction effluent crashed over the 1-propanol, $m_{\text{propOH}}^{\text{vial}}$ is the mass of propanol used to dilute the samples for the GC analysis, and $m_{\text{product}}^{\text{vial}}$ and $\rho_{\text{sample}}^{\text{vial}}$ represent the mass and density of the sample (reaction effluent + crashing solvent) used in the GC analysis.

$$C_i^{\text{outlet}} = C_{\text{pentOH}}^{\text{vial}} \cdot \text{RF}_{x,i} \cdot \frac{\text{Area}_i}{\text{Area}_{\text{pentOH}}} \left(1 + \frac{m_{\text{propOH}}^{\text{condenser}}}{m_{\text{product}}^{\text{condenser}}} \cdot \frac{\rho_{\text{product}}^{\text{condenser}}}{\rho_{\text{propOH}}} \right) \times \left(1 + \frac{m_{\text{propOH}}^{\text{vial}}}{m_{\text{product}}^{\text{vial}}} \cdot \frac{\rho_{\text{sample}}^{\text{condenser}}}{\rho_{\text{propOH}}} \right) \quad (1)$$

Ethanol conversion was calculated for each experiment by using eqn (2).

$$X_{\text{Ethanol}} = 1 - \frac{C_{\text{Ethanol}}^{\text{outlet}}}{C_{\text{Ethanol}}^{\text{inlet}}} \quad (2)$$

The carbon balance was calculated as the ratio of the carbon concentration detected in the outlet and the carbon concentration fed to the reaction setup as presented in eqn (3).

$$\text{CB} = \frac{\sum_i \text{CN}_i \cdot C_i^{\text{outlet}}}{2C_{\text{Ethanol}}^{\text{inlet}} + 4C_{\text{Butanol}}^{\text{inlet}}} \quad (3)$$

2.3 Kinetic modeling

The change in flow rate of component i (F_i) participating in j reactions in a packed-bed reactor is expressed by eqn (4).

$$\frac{dF_i}{d\tau} = F_T^0 \sum_j \nu_{i,j} r_j \quad (4)$$

where, τ represents the contact time of the entire gas, calculated as w_{cat}/F_T^0 , where w_{cat} is the mass of catalyst and F_T^0 is the mole flow rate in the inlet of the reactor, $\nu_{i,j}$ is the stoichiometric coefficient of species i in reaction j , and r_j is the rate of reaction j . Based on the results obtained from the catalytic results described in the previous section, we have identified 60 species, from which we have taken the 31 most abundant to generate our kinetic model. Such species are labelled with the

following nomenclature throughout our text: ethanol (E), butanol (B), hexanol (H), octanol (O), decanol (D), 2-ethylbutanol (2EB), 2-ethylhexanol (2EH), 2-butyloctanol (2BO), acetaldehyde (A), butyraldehyde (Ba), hexanal (He), octanal (Oc), 2-ethylbutanal (2EBa), 2-ethylhexanal (2EHa), ethyl acetate (EA), butyl acetate (BA), ethyl butyrate (EB), butyl butyrate (BB), ethyl hexanoate (EH), hexyl acetate (HA), butyl hexanoate (BH), butyl octanoate (BO), 2ethylbutyl acetate (2EBA), acetone (Ac), 2-pentanone (2Po), 2-heptanone (2Ho), 4-heptanone (4Ho), 4-nonanone (4No), water (H_2O), carbon monoxide (CO), and hydrogen (H_2). In addition, nine unobserved but expected intermediates, namely, 2-butenal ($\text{Ba}_=$), 2-hexenal ($\text{He}_=$), 2-ethyl-2-butenal ($2\text{EBa}_=$), 2-octenal ($\text{Oc}_=$), 2-ethyl-2-hexenal ($2\text{EHa}_=$), 2-decenal ($\text{De}_=$), decanal (De), 2-butyl-2-octenal (2BOc), and 2-butyloctanal (2BOc) were included in the model.

Regression of kinetic parameters for the system of 37 ordinary differential equations generated from eqn (4) has been achieved after implementation of the mathematical model in Matlab[®] R2022b, by following the algorithm depicted in Fig. 1. As observed, the algorithm is fed with estimated values for the kinetic constants (k^0) and liquid phase concentrations of the species quantified in our experiments ($C_i^{\text{exp}}(\tau)$) to minimize the objective function. In the algorithm, differential balances for the gas phase flows are established and solved by using the ODE23s function. For solution of the ordinary differential equations the initial conditions were assumed to be the mole flow of each species in the inlet of the reactor. Subsequently, gas phase mole fractions are computed to be employed in the proposed rate model. A more detailed derivation of the rate expression is presented in Section S2 of the ESI.[†] Estimation of the liquid phase concentrations is carried out by renormalizing the gas phase mole fractions (y_i) in a H_2 - and CO-free basis. Ultimately, profiles of liquid phase carbon concentrations are obtained and compared to experimental data to proceed with the optimization. A hybrid optimization approach was carried out to minimize the objective function, initializing the routine with the particle swarm algorithm by invoking the 'particleswarm' function with a swarm size of 100 particles and typically a maximum of 50 iterations and subsequent refinement of the best solution by using the 'lsqcurvefit' function. Determination of the parameters' confidence intervals was carried out with the 'nlparci' function at 95% confidence interval, and estimation of the confidence band around the regressed model with the 'nlpredci' function. For clarification, experimental data obtained feeding ethanol, and the blend ethanol-butanol were used simultaneously to fit the model.

The objective function to be minimized was formulated as a weighted combination of squared error and logarithmic squared error. While the sum of the absolute squared error tends to bias towards species with high carbon concentrations, the sum of the logarithmic squared error assigns greater weight to species with low concentrations. In our exploratory study, we found that combining both approaches helped balance the performance of our optimization problem. The values of the weights α and β were determined based on the objective function value after the optimization procedure, considering only the absolute squared error and the logarithmic squared



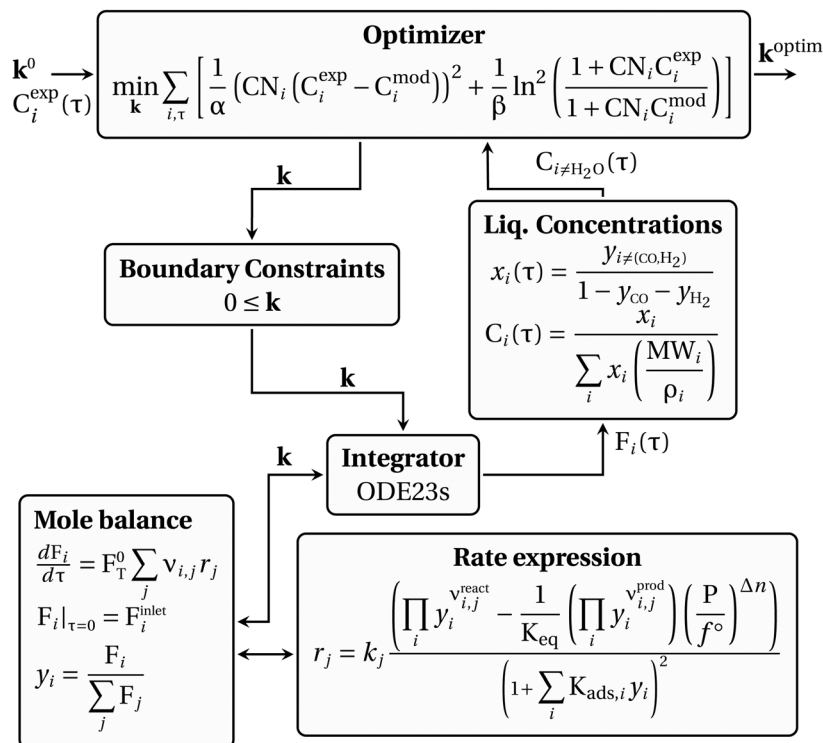


Fig. 1 Flowchart of the optimization algorithm employed for regression of the kinetic constants that rule the chemistry of alcohols coupling.

error, respectively. In this study, $\alpha = 4.44$ and $\beta = 47.18$. This approach ensures that the hybrid objective function has equal contribution from both error types, converging towards a value of two as the optimization progresses.

The rate model proposed in this work is based on a LHHW adsorption model given its proven good performance for modeling the Guerbet coupling chemistry over hydroxyapatite,^{17–19} Ni-La₂O₃/γ-Al₂O₃,²² and Mg_xAl_yO.²¹ We have assumed that dehydrogenation and dehydrogenative coupling reactions between alcohols, aldehydes, hydrogen, and esters are ruled by reversible reactions due to the presence of copper.^{24,25} In this respect, the general rate expression for the reaction $aA + bB \rightleftharpoons cC + dD$ is ruled by eqn (5). Such an expression has been applied to the set of 37 chemical reactions that model the chemistry in the reactor and the final functional shape is shown in Table 1. Our previous work elucidating the reaction chemistry running over the catalyst surface¹¹ served as the basis for the derivation of the reaction network that rules our model.

$$r_j = k_{f,j} \left(y_A^a y_B^b - \frac{y_C^c y_D^d}{K_{\text{eq},j}} \left(\frac{P}{f^\circ} \right)^{\Delta n_j} \right) C^2 \quad (5)$$

where, $k_{f,j}$ and $K_{\text{eq},j}$ are the forward rate constant and equilibrium constant of the reaction j . Equilibrium constants have been calculated as $K_{\text{eq},j} = e^{-\frac{\Delta G_j(T)}{RT}}$ by using thermochemical data at 325 °C. The thermochemical data used for equilibrium constant calculations were sourced from Aspen Properties. The data were validated, when possible, against thermochemical information available in the open literature, including the NIST Webbook,^{26,27} JANAF tables,²⁸ and chemical properties handbooks.^{29,30} Details

of the validation process, including supporting calculations, are provided in the ESI† (.xls file). y_i is the mole fraction of species i in the gas phase, a , b , c , and d are stoichiometric coefficients and Δn_j is the change in moles of the chemical reaction j , P represents the total pressure of the system, and f° is the fugacity of an ideal gas at 1 bar. The value of C is calculated by means of eqn (6).

$$C = \left(1 + \sum_{i,j} \left(K_j^{\text{ads}} y_{i,j} \right) \right)^{-1} \quad (6)$$

In eqn (6) j might be $j = (\text{alcohol, aldehyde, ketone, ester, water})$ and i represents the species i that belongs to group j . In our approach, we have assumed that all species belonging to a given group j share the same adsorption constant. For example, all alcohols have the same adsorption constant. Also, we have assumed that the adsorption constant of any ketone is the same as the adsorption constant of an aldehyde.

2.4 Process modeling simulations

With the aim of demonstrating the feasibility of integrating our kinetic model into a process simulation to develop routes for selective production of diesel fuel precursors (e.g., hexanol, octanol, decanol), we have embedded the kinetic model here developed for the Guerbet coupling reactor in an Aspen Plus V14 flowsheet simulation. A process flow diagram of the simulation is depicted in Fig. 2.

As displayed, ethanol is pumped (P-101) from a feed-tank into a heat exchanger for its vaporization before feeding it into the Guerbet reactor (R-101). Pressure is increased in this stage from atmospheric up to 20 barg. The Guerbet reactor is simulated



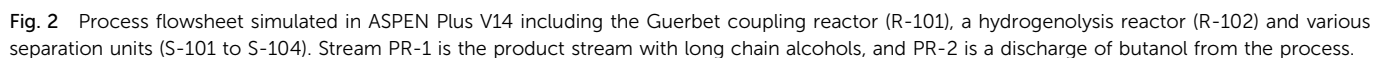
Table 1 Reactions (Guerbet and dehydrogenative coupling) and rate functions used for the kinetic model. The exact expression for C (adsorption term in the heterogeneous model) is presented in eqn (6)

ID	Group	Reactions	Rate function (r_j)
1	Alcohol $\rightleftharpoons$ aldehyde + H_2	Ethanol $\rightleftharpoons$ acetaldehyde + H_2	$r_1 = k_1 \left(y_E - \frac{y_A y_{H_2}}{K_{eq,1}} \left(\frac{P}{f^\circ} \right) \right) C^2$
2		Butanol $\rightleftharpoons$ butyraldehyde + H_2	$r_2 = k_2 \left(y_B - \frac{y_{Ba} y_{H_2}}{K_{eq,2}} \left(\frac{P}{f^\circ} \right) \right) C^2$
3		Hexanol $\rightleftharpoons$ hexanal + H_2	$r_3 = k_3 \left(y_H - \frac{y_{He} y_{H_2}}{K_{eq,3}} \left(\frac{P}{f^\circ} \right) \right) C^2$
4		Octanol $\rightleftharpoons$ octanal + H_2	$r_4 = k_4 \left(y_O - \frac{y_{Oc} y_{H_2}}{K_{eq,4}} \left(\frac{P}{f^\circ} \right) \right) C^2$
5		Decanol $\rightleftharpoons$ decanal + H_2	$r_5 = k_5 \left(y_D - \frac{y_{De} y_{H_2}}{K_{eq,5}} \left(\frac{P}{f^\circ} \right) \right) C^2$
6		2-Ethylbutanol $\rightleftharpoons$ 2-ethylbutanal + H_2	$r_6 = k_6 \left(y_{2EB} - \frac{y_{2EBa} y_{H_2}}{K_{eq,6}} \left(\frac{P}{f^\circ} \right) \right) C^2$
7		2-Ethylhexanol $\rightleftharpoons$ 2-ethylhexanal + H_2	$r_7 = k_7 \left(y_{2EH} - \frac{y_{2EHa} y_{H_2}}{K_{eq,7}} \left(\frac{P}{f^\circ} \right) \right) C^2$
8		2-Butyloctanol $\rightleftharpoons$ 2-butyloctanal + H_2	$r_8 = k_8 \left(y_{2BO} - \frac{y_{2BOc} y_{H_2}}{K_{eq,8}} \left(\frac{P}{f^\circ} \right) \right) C^2$
9	2Aldehyde $\rightarrow$ alkenal + H_2O	2 Acetaldehyde $\rightarrow$ 2-butenal + H_2O	$r_9 = k_9 y_A^2 C^2$
10		Acetaldehyde + butyraldehyde $\rightarrow$ 2-hexenal + H_2O	$r_{10} = k_{10} y_A y_{Ba} C^2$
11		Acetaldehyde + butyraldehyde $\rightarrow$ 2-ethyl-2-butenal + H_2O	$r_{11} = k_{11} y_A y_{Ba} C^2$
12		Acetaldehyde + hexanal $\rightarrow$ 2-octenal + H_2O	$r_{12} = k_{12} y_A y_{He} C^2$
13		Acetaldehyde + hexanal $\rightarrow$ 2-ethyl-2-hexenal + H_2O	$r_{13} = k_{13} y_A y_{He} C^2$
14		Acetaldehyde + octanal $\rightarrow$ 2-decenal + H_2O	$r_{14} = k_{14} y_A y_{Oc} C^2$
15		2Butyraldehyde $\rightarrow$ 2-ethyl-2-hexenal + H_2O	$r_{15} = k_{15} y_{Ba}^2 C^2$
16		2Hexanal $\rightarrow$ 2-butyl-2-octenal + H_2O	$r_{16} = k_{16} y_{He}^2 C^2$
17	Alkenal + $H_2 \rightarrow$ aldehyde	2-Butenal + $H_2 \rightarrow$ butyraldehyde	$r_{17} = k_{17} y_{Ba} y_{H_2} C^2$
18		2-Hexenal + $H_2 \rightarrow$ hexanal	$r_{18} = k_{18} y_{He} y_{H_2} C^2$
19		2-Ethyl-2-butenal + $H_2 \rightarrow$ 2-ethylbutanal	$r_{19} = k_{19} y_{2EBa} y_{H_2} C^2$
20		2-Octenal + $H_2 \rightarrow$ octanal	$r_{20} = k_{20} y_{Oc} y_{H_2} C^2$
21		2-Ethyl-2-hexenal + $H_2 \rightarrow$ 2-ethylhexanal	$r_{21} = k_{21} y_{2EHa} y_{H_2} C^2$
22		2-Decenal + $H_2 \rightarrow$ decanal	$r_{22} = k_{22} y_{De} y_{H_2} C^2$
23		2-Butyl-2-octenal + $H_2 \rightarrow$ 2-butyloctanal	$r_{23} = k_{23} y_{2BOc} y_{H_2} C^2$
24	Alcohol + aldehyde $\rightleftharpoons$ ester + H_2	Ethanol + acetaldehyde $\rightleftharpoons$ ethyl acetate + H_2	$r_{24} = k_{24} \left(y_E y_A - \frac{y_{EA} y_{H_2}}{K_{eq,24}} \right) C^2$
25		Butanol + acetaldehyde $\rightleftharpoons$ butyl acetate + H_2	$r_{25} = k_{25} \left(y_B y_A - \frac{y_{BA} y_{H_2}}{K_{eq,25}} \right) C^2$
26		Ethanol + butyraldehyde $\rightleftharpoons$ ethyl butyrate + H_2	$r_{26} = k_{26} \left(y_E y_{Ba} - \frac{y_{EB} y_{H_2}}{K_{eq,26}} \right) C^2$
27		Butanol + butyraldehyde $\rightleftharpoons$ butyl butyrate + H_2	$r_{27} = k_{27} \left(y_B y_{Ba} - \frac{y_{BB} y_{H_2}}{K_{eq,27}} \right) C^2$
28		Butanol + hexanal $\rightleftharpoons$ butyl hexanoate + H_2	$r_{28} = k_{28} \left(y_B y_{He} - \frac{y_{BH} y_{H_2}}{K_{eq,28}} \right) C^2$
29		Ethanol + hexanal $\rightleftharpoons$ ethyl hexanoate + H_2	$r_{29} = k_{29} \left(y_E y_{He} - \frac{y_{EH} y_{H_2}}{K_{eq,29}} \right) C^2$
30		Hexanol + acetaldehyde $\rightleftharpoons$ hexyl acetate + H_2	$r_{30} = k_{30} \left(y_H y_A - \frac{y_{HA} y_{H_2}}{K_{eq,30}} \right) C^2$
31		Butanol + octanal $\rightleftharpoons$ butyl octanoate + H_2	$r_{31} = k_{31} \left(y_B y_{Oc} - \frac{y_{BO} y_{H_2}}{K_{eq,31}} \right) C^2$
32		2-Ethylbutanol + acetaldehyde $\rightarrow$ 2ethylbutyl acetate + H_2	$r_{32} = k_{32} \left(y_{2EB} y_A - \frac{y_{2EBAc} y_{H_2}}{K_{eq,32}} \right) C^2$
33	2 Aldehyde $\rightarrow$ ketone + $CO + H_2$	2Acetaldehyde $\rightarrow$ acetone + $CO + H_2$	$r_{33} = k_{33} y_A^2 C^2$
34		Acetaldehyde + butyraldehyde $\rightarrow$ 2-pentanone + $CO + H_2$	$r_{34} = k_{34} y_A y_{Ba} C^2$
35		Acetaldehyde + hexanal $\rightarrow$ 2-heptanone + $CO + H_2$	$r_{35} = k_{35} y_A y_{He} C^2$
36		2Butyraldehyde $\rightarrow$ 4-heptanone + $CO + H_2$	$r_{36} = k_{36} y_{Ba}^2 C^2$
37		Butyraldehyde + hexanal $\rightarrow$ 4-nonanone + $CO + H_2$	$r_{37} = k_{37} y_{Ba} y_{He} C^2$

as a plug flow reactor (PFR) in which the set of reactions and their rate expressions derived from our kinetic model are implemented.

Reaction contact time is specified by varying the mass of catalyst loaded in the reactor. The PFR is simulated as isothermal and





In this work we have investigated the influence of the following three process variables: (1) the residence time τ in the Guerbet coupling reactor, (2) the H_2 :ester ratio on the hydrogenolysis reactor, and (3) the discharge ratio of butanol in stream PR-2 leaving the separation column S-104 to understand its impact on the overall production of C_{6+} alcohols. Quantification of our results was performed by calculating carbon yields in the stream PR-1. For example, in eqn (7) we present the carbon yield to C_{6+} alcohols, which is defined as the ratio of the concentration of the carbon atoms contained in the C_{6+} alcohols in the stream PR-1 to the carbon concentration in the ethanol feed stream. Carbon yields for other species like esters, ketones, and aldehydes have been calculated in a similar way and reported as cumulated fractions.

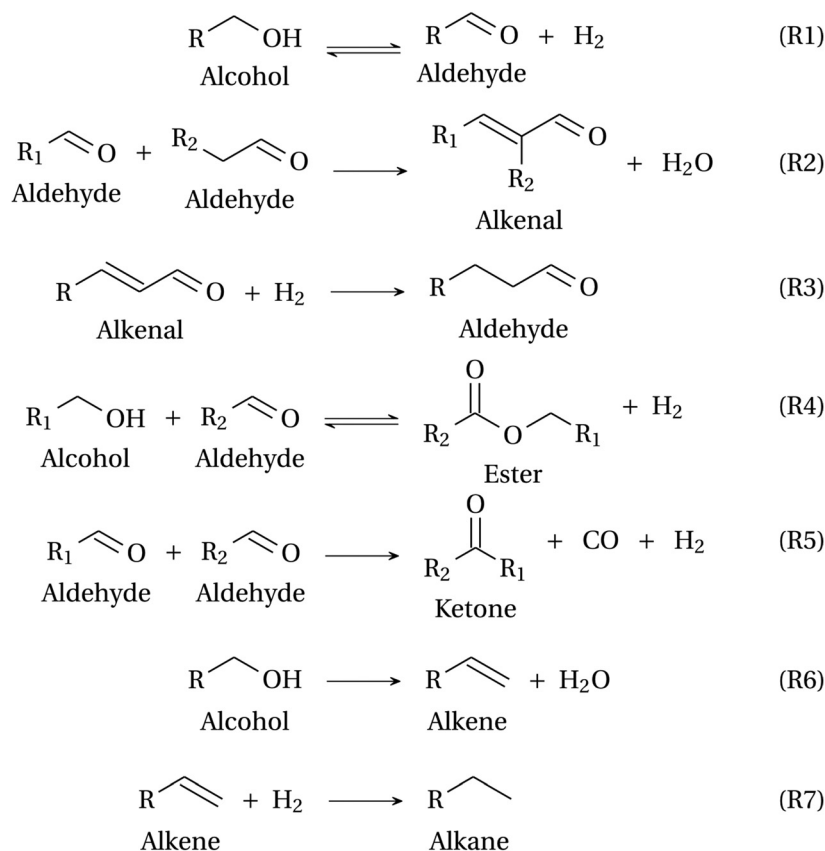
$$Y_{C_{6+}} = \frac{6(C_{\text{Hexanol}}^{\text{Outlet}} + C_{2\text{-Ethylbutanol}}^{\text{Outlet}}) + 8(C_{\text{Octanol}}^{\text{Outlet}} + C_{2\text{-Ethylhexanol}}^{\text{Outlet}}) + 10C_{\text{Decanol}}^{\text{Outlet}}}{2C_{\text{Ethanol}}^{\text{Inlet}}} \quad (7)$$

In the following paragraphs, we present the kinetic modeling for the Guerbet coupling chemistry using a fresh catalyst. The stability of this catalyst has been previously tested under

For our model, we selected a set of 31 identified species plus 9 expected intermediates labelled as alcohols, aldehydes, esters,

Fig. 4 presents the ethanol conversion as a function of the contact time for both pure ethanol and cofeed experiments. Note that since butanol is a product and a reactant for the cofeed experiments, calculation of its conversion extent is

Property	Value
Cu loading (wt%)	0.3
Mg:Al molar ratio	2.9
BET surface area ($\text{m}^2 \text{g}^{-1}$)	172
Pore volume (mL g^{-1})	0.5
Basic sites ($\mu\text{mol g}^{-1}$)	325
Acid sites ($\mu\text{mol g}^{-1}$)	861



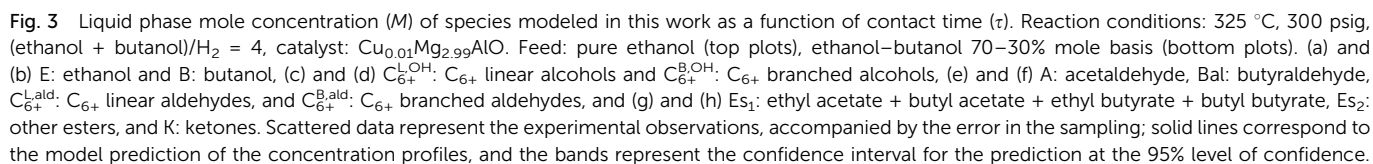
Scheme 1 Chemical reactions that span the reaction network of the Guerbet coupling chemistry over copper-based catalysts. (R1): alcohol dehydrogenation, (R2): aldol condensation, (R3): alkenal hydrogenation, (R4): dehydrogenative coupling, (R5): retro-aldol CO elimination for ketone formation, (R6): monomolecular dehydration, and (R7): alkene hydrogenation.

challenging, and no attempts were made for this purpose. As displayed, experiments indicate that ethanol conversions in cofeed experiments are slightly higher than those in the control experiments within the first 5 h $\text{kg}_{\text{cat}} \text{ kmol}_{\text{gas}}^{-1}$ of contact time. This initial behavior is because the addition of butanol readily produces large amounts of butyraldehyde, which form a butanol/butyraldehyde-rich pool that enables new reaction pathways for coupling ethanol and acetaldehyde moieties. Interestingly, in both tested scenarios, ethanol conversion slows at higher contact times, reaching a maximum of $\sim 70\%$, consistent with previous observations from our group.¹¹ This deceleration is attributed to the competitive adsorption of water, which impacts the adsorption frequency of reactive aldehydes required for the Guerbet coupling mechanism.³⁸ Since aldehyde partial pressure is governed by its equilibrium with the parent alcohol, a reduced aldehyde consumption rate directly limits alcohol conversion.

Acetaldehyde and butyraldehyde concentration profiles are presented in Fig. 3(e) and (f). As depicted, acetaldehyde is readily formed from ethanol dehydrogenation over copper active sites. This dehydrogenation is observed to occur fast due to the rapid increase of acetaldehyde concentration in a short contact time. In contrast, the butyraldehyde concentration exhibits two different initial increasing rates for the cofeed and control experiments. In the control experiments, butyraldehyde is formed from aldol condensation followed by a hydrogenation, while in

the cofeed experiments this aldehyde is formed from butanol dehydrogenation. Given that the apparent formation rate of butyraldehyde (the slope of the profile) in the control experiments at early contact times is lower than that of the cofeed experiments, it is concluded that butanol dehydrogenation is faster than acetaldehyde self-condensation. In our system, once butyraldehyde is formed, acetaldehyde and butyraldehyde react on the surface yielding C_{4-8} alcohols (c) and (d), and ketones and esters (g) and (h). The intermediate role of the linear aldehydes in the reaction network is also observed in the diagram, showing an increasing concentration profile at low contact times that peaks between 2–3 h $kg_{cat} \text{ kmol}_{gas}^{-1}$ and decreases afterwards. The exact position of the aldehyde peak depends on whether the precursor alcohol was fed into the reactor or not. Although other species such as butanol or hexanol can also be considered intermediates due to participation in dehydrogenative couplings to produce esters and ketones, their intermediate fashion is not evident within the tested contact times.

Panels (a) and (b) present the butanol concentration profiles for our experiments, exhibiting two different trends. In the control experiments it is observed that butanol concentration is continuously increasing, while in the cofeed experiments butanol concentration decreases from 4.4 M with an apparent convergence to a steady value of 3.4 M. The concentration of other alcohols is plotted in Fig. 3(c) and (d). As shown, the



ethanol feed (control) experiments, leveling off at a ratio of 3. This finding suggests that feeding butanol does not affect nucleophilicity and/or electrophilicity roles in the condensation of acetaldehyde with butyraldehyde, and that at high contact times $\sim 75\%$ of C_6 alcohols will be hexanol. Nonetheless, a similar behavior is not observed for C_8 alcohols in the control and cofeed experiments. As shown, the ratio of linear to branched alcohols in the control experiments converges to ~ 2 , while in the cofeed experiments the convergence is towards 1. This disparity in relative ratios of C_8 alcohols is due to the fact that feeding butanol increases the production of 2-ethylhexanol, since this alcohol can also be formed from butyraldehyde self-condensation. The plot demonstrates that alcohols coupling over our catalysts form predominantly linear species rather than branched. This selectivity of linear over branched alcohols can also be regarded in the instantaneous selectivity calculated from the kinetic constants of our model. In Table 1 it is seen that reactions (10) and (11) modeling the formation of 2-hexenal and 2-ethyl-2-butenal, respectively, share the same reactants, which allows us to compute the instantaneous selectivity of the linear over the branched unsaturated aldehydes as the ratio of their kinetic constants, yielding a value of $S_{\text{He-2EBa-}} = 5.4$. In further steps of the reaction network, the alkenals are hydrogenated into their respective aldehydes, which react to form new species like alcohols at rates that yield the results presented in Fig. 5.

Table 3 Kinetic constants for ethanol + butanol coupling over $\text{Cu}_{0.01}\text{Mg}_{2.99}\text{AlO}$ at 325 °C and 300 psig. Rate constants are reported in units of $\text{kmol}_i \text{h}^{-1} \text{kg}_{\text{cat}}^{-1}$, while equilibrium and adsorption constants are unitless

Reaction	Parameter	$k_f \pm \text{CI}_{\alpha=0.05}$	K_{eq}
1	k_1	$12\,236 \pm 1706$	3.19
2	k_2	$29\,219 \pm 6334$	0.88
3	k_3	$1.035 \times 10^6 \pm 1.4028 \times 10^6$	1.24
4	k_4	$10\,252 \pm 10\,851$	0.39
5	k_5	123 ± 108	0.41
6	k_6	$29\,986 \pm 30\,720$	1.00
7	k_7	$30\,996 \pm 43\,001$	0.87
8	k_8	123 ± 108	0.87
9	k_9	$1\,014\,500 \pm 192\,330$	
10	k_{10}	$759\,300 \pm 138\,900$	
11	k_{11}	$140\,520 \pm 33\,472$	
12	k_{12}	$1\,732\,700 \pm 477\,480$	
13	k_{13}	$116\,090 \pm 172\,470$	
14	k_{14}	$1\,761\,100 \pm 779\,760$	
15	k_{15}	$53\,227 \pm 36\,027$	
16	k_{16}	$22\,582 \pm 17\,638$	
17–23	$k_{17}-k_{23}$	$1.5531 \times 10^7 \pm 2.0014 \times 10^7$	
24	k_{24}	7209 ± 1688	5.41
25	k_{25}	$14\,955 \pm 4083$	9.11
26	k_{26}	9016 ± 2252	30.47
27	k_{27}	$13\,175 \pm 3594$	25.15
28	k_{28}	$64\,958 \pm 22\,391$	12.91
29	k_{29}	$17\,895 \pm 10\,171$	10.45
30	k_{30}	$24\,300 \pm 20\,908$	7.97
31	k_{31}	$149\,400 \pm 128\,300$	54.20
32	k_{32}	$22\,582 \pm 17\,638$	5.71
33	k_{33}	$21\,031 \pm 14\,301$	
34	k_{34}	$46\,745 \pm 23\,823$	
35–37	$k_{35}-k_{37}$	$22\,582 \pm 17\,638$	
38	$K_{\text{alcohols}}^{\text{ads}}$	274 ± 33	
39	$K_{\text{water}}^{\text{ads}}$	616 ± 230	
40	$K_{\text{aldehydes + ketones}}^{\text{ads}}$	36 ± 56	
41	$K_{\text{esters}}^{\text{ads}}$	7807 ± 1900	

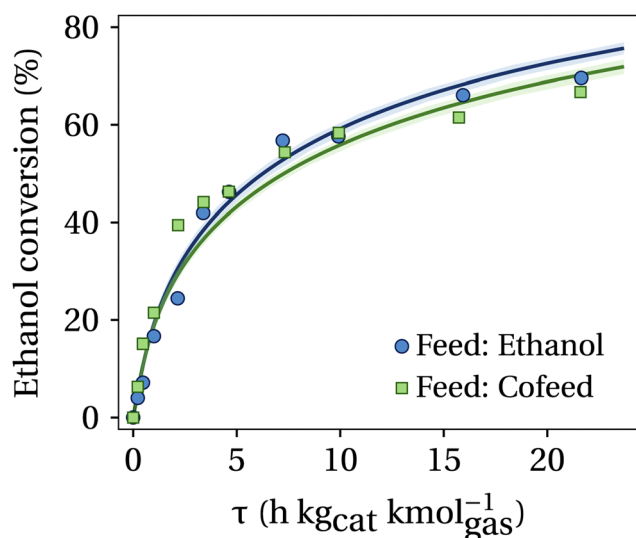


Fig. 4 Ethanol conversion as a function of contact time (τ). Reaction conditions: 325 °C, 300 psig, (ethanol + butanol)/ H_2 = 4, catalyst: $\text{Cu}_{0.01}\text{Mg}_{2.99}\text{AlO}$. Cofeed: ethanol–butanol 70–30% mole basis. Solid lines represent the prediction of the model.

As previously commented, the motivation of feeding butanol into the alcohol coupling reactor is to increase the production rates

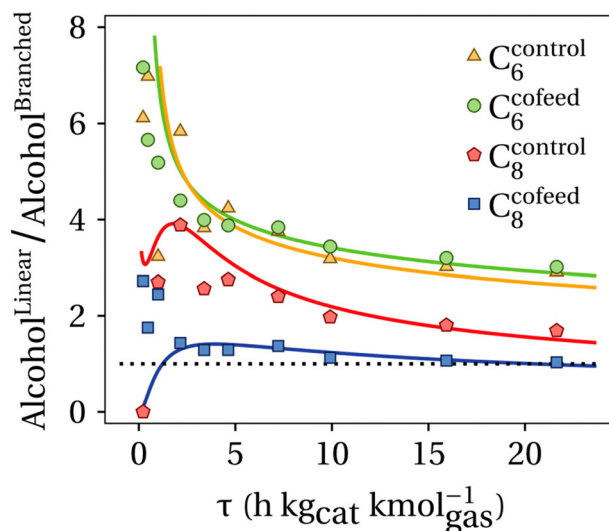


Fig. 5 Relative concentration ratio of linear to branched alcohols for cofeed and control (pure ethanol feed) experiments as a function of contact time. C_6 and C_8 ratios refer to hexanol/2-ethylbutanol and octanol/2-ethylhexanol, respectively. Reaction conditions: 325 °C, 300 psig, (ethanol + butanol)/ H_2 = 4, catalyst: $\text{Cu}_{0.01}\text{Mg}_{2.99}\text{AlO}$. Cofeed: ethanol–butanol 70–30% mole basis. Solid lines represent the prediction of the model. A dotted line at ratio = 1 has been added as a visual aid.

of longer chain alcohols, primarily hexanol produced from the net reaction ethanol + butanol $\rightarrow$ hexanol + H_2O . Nevertheless, the coupling of the same reactants also yields 2-ethylbutanol, which is a less desirable product. Linear alcohols undergo preferably bimolecular dehydrations to ethers while branched alcohols dehydrate monomolecularly into olefins.^{13,15} As a reminder, the product of our global process is a blend of ethers produced through bimolecular dehydration of alcohols. Fig. 6 presents the ratio of the concentration of the linear and branched alcohols in the cofeed experiments with respect to the control experiments. As observed, the concentration of all alcohols is higher in the cofeed experiments at early contact times, as compared to the control ones, with a continuous decrease that converges asymptotically to values around 1. These results indicate then that cofeeding butanol fosters the condensation of linear alcohols or aldehydes at higher rates to produce longer chain alcohols. Noticeably, at high contact times the experiments show few increases in hexanol concentration if butanol is cofed into the reactor. For longer chain alcohols like octanol and decanol the experimental observations and the model lines fall below the dotted reference line, indicating that other reactions are consuming hexanol/octanol and acetaldehyde, which causes a decreased concentration of octanol and decanol in the final product compared to the control case. It is also observed that at long contact times the production of 2-ethylbutanol is virtually unaffected by the addition of butanol, however the final concentration of 2-ethylhexanol increased by 50% when 30 mol% butanol is fed into the reactor.

One of the consequences of cofeeding butanol is the promotion of dehydrogenative coupling reactions between butanol/butyraldehyde and aldehydes/alcohols. Fig. S3 (ESI[†]) presents a

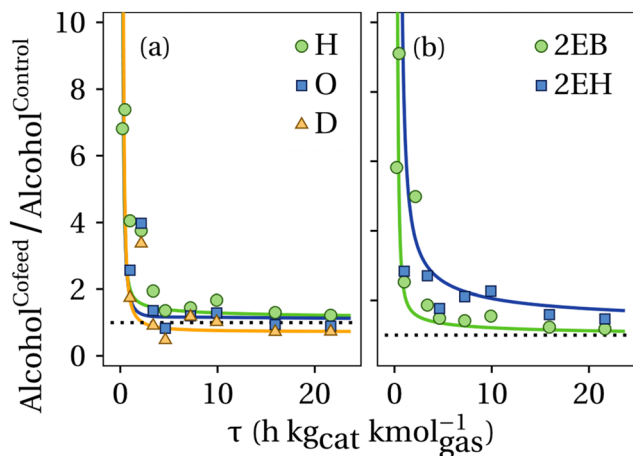


Fig. 6 Relative concentration ratio of (a) linear and (b) branched alcohols in cofeed to control (pure ethanol feed) experiments. Reaction conditions: 325 °C, 300 psig, (ethanol + butanol)/H₂ = 4, catalyst: Cu_{0.01}Mg_{2.99}AlO. Cofeed: ethanol–butanol 70–30% mole basis. Solid lines represent the prediction of the model. The dotted line at ratio = 1 has been added as a visual aid. H: hexanol, O: octanol, D: decanol, 2EB: 2-ethylbutanol, and 2EH: 2-ethylhexanol.

comparison of the concentration of the esters produced from ethanol and butanol coupling for the studied cofeed scenario relative to the pure ethanol feed. As observed, feeding less ethanol into the system caused a reduction of ~50% in the mole concentration of ethyl acetate compared to the reference scenario. This is due to the lower amount of acetaldehyde generated in the cofeed experiments that limited the coupling of ethanol with acetaldehyde. In contrast, the opposite trend is observed for butyl butyrate, whose concentration in the final product increased by at least 150% due to the presence of butanol in the feed. The concentration ratio of unsymmetrical esters like butyl acetate and ethyl butyrate indicates that at early contact times there exist promotion of the dehydrogenative coupling reaction that forms additional C₆ esters in the cofeed scenario. Experiments also reveal that the relative butyl acetate concentration converges towards a value slightly higher than one, which indicates that even at high contact times the production of butyl acetate is more promoted than the production of ethyl butyrate when butanol is cofed. The relative concentration of ethyl butyrate to butyl acetate in the cofeed and control experiments is presented in Fig. S4 (ESI†). As observed, ethyl butyrate is formed in higher concentration than butyl acetate in the control experiments. Nonetheless, in the cofeed experiments both esters are observed to have virtually the same concentration at most of the contact times. The explanation for this resides in the fact that butyl acetate is formed more easily in cofeed than in the control experiments. Formation of butyl acetate requires the dehydrogenative coupling of butanol and acetaldehyde, which are easily furnished in the cofeed experiments at low contact times, while for the control experiments butanol formation requires previous butyraldehyde synthesis through acetaldehyde condensation. As such, in cofeed experiments the production of butyl acetate is

observed to occur at comparable rates as ethyl butyrate. Fig. S5 (ESI†) presents the ratio of the reaction quotient (Q) to the equilibrium constant (K_{eq}) for alcohol dehydrogenation and dehydrogenative coupling, as predicted by our kinetic model. As observed, C₄₊ alcohols rapidly establish equilibrium with their respective aldehydes and hydrogen. In contrast, ethanol remains far from equilibrium, maintaining the driving force that favors acetaldehyde formation. Similarly, the model indicates that dehydrogenative coupling reactions of alcohols and aldehydes are also far from equilibrium, with a Q/K_{eq} ratio lower than 0.1. This suggests that the reaction system still retains significant chemical potential for ester production.

Of particular interest is understanding the effect of an ethanol–butanol cofeed for the linear alcohols and the respective aldehyde carbon yields as described by eqn (8). Fig. 7 presents a heat map of the carbon yield of linear alcohols + aldehydes as a function of the contact time and the mole fraction of butanol in the feed of the reactor. As observed, the model indicates within the analyzed range of contact times, the maximum amount of carbon atoms for the linear alcohols/aldehydes is ~14% of the carbon fed into the system. Remarkably, it is observed that at a constant contact time the carbon yield to linear species decreases after $x_{ButOH}^{Feed} \sim 25\%$, which comes from a reduction in the concentration of acetaldehyde needed to proceed with the aldol-condensation towards linear species. Fig. 7 also suggests that there exists a region between $x_{ButOH}^{Feed} = 0\text{--}30\%$ and high contact times that maximizes the production of linear alcohols + aldehydes. Over the heat map a white line has been drawn and represents $(\partial Y_C / \partial x_{ButOH}^{Feed})_{\tau} = 0$. This line indicates the maximum concentration of diesel fuel precursors that can be achieved at constant contact time. The trajectory of the drawn line over the heatmap indicates that cofeeding between 23–28 mole% of butanol is the optimal blend ratio to maximize the advantageous effect of butanol

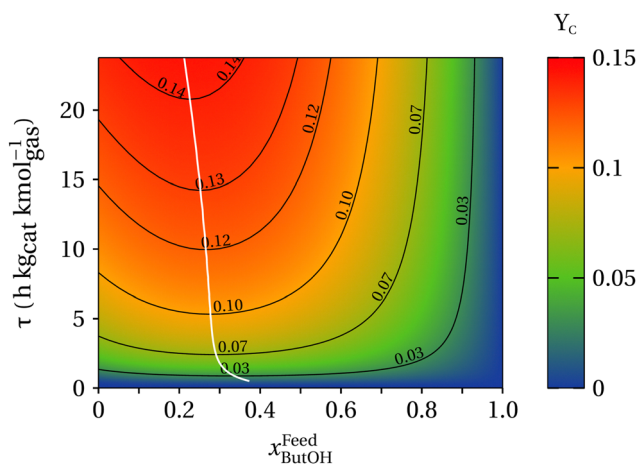


Fig. 7 Heat map of model predicted carbon yield of diesel fuel precursors as a function of contact time (τ) and mole fraction of butanol in the feed composed of a blend of ethanol and butanol, reacting over Cu_{0.01}–Mg_{2.99}AlO at 325 °C, 300 psig, and (ethanol + butanol)/H₂ = 4 in the feed. The white line represents the couple (x_{ButOH}^{Feed}, τ) that maximizes the carbon concentration of diesel fuel precursors at a determined contact time.



cofeeding, particularly in the production of linear and branched C₆₊ alcohols and aldehydes.

$$Y_C = \frac{6(C_{\text{Hexanol}}^{\text{Outlet}} + C_{\text{Hexanal}}^{\text{Outlet}}) + 8(C_{\text{Octanol}}^{\text{Outlet}} + C_{\text{Octanal}}^{\text{Outlet}}) + 10C_{\text{Decanol}}^{\text{Outlet}}}{2C_{\text{Ethanol}}^{\text{Inlet}} + 4C_{\text{Butanol}}^{\text{Inlet}}} \quad (8)$$

3.2 Process modeling

The process simulation for diesel fuel production has been carried out in previous works,^{8,9} where a conversion reactor was used to calculate the composition in the outlet stream of the Guerbet coupling reactor. The implementation of a conversion reactor implies that experiments must be carried out at a fixed conversion for the respective feed composition to obtain an appropriate product composition. The influence of the catalyst quantity and residence time must therefore always be determined separately and cannot be predicted by simple variations in Aspen Plus. Nevertheless, introduction of a kinetic model to simulate the reactor's behavior allows the investigation with more flexibility of the influence of variables such as residence time and feed composition on the performance of the overall system (product quantities, energy demands, *etc.*).

In preliminary simulations, the Guerbet coupling kinetic model was implemented in a plug-flow reactor in Aspen Plus V14 as a custom model. Converged simulations were obtained for simulations feeding pure ethanol and the 70–30% ethanol–butanol cofeed, achieving the same product composition as in the model presented in the kinetic modeling section. This demonstrates the feasibility of integrating the Guerbet coupling kinetic model in a process simulation. Our kinetic model was then embedded in the simulation process presented in Fig. 2, which consists of the Guerbet reaction, a downstream hydrogenolysis reactor, and the product separation units. This simulation elucidated the impact of process variables such as the resident time and the ethanol–butanol cofeed (varying the butanol recycling ratio) in the Guerbet coupling reactor, as well

as the H₂:ester ratio in the hydrogenolysis reactor. The influence of the contact time in the coupling reactor was assessed by changing the amount of catalyst in the reactor, while the impact of the H₂:ester ratio fed into the hydrogenolysis reactor over the product yields is assessed by studying specific ratios of 400:1, 200:1, and 100:1.

Fig. 8 shows the simulation results of the process depicted in Fig. 2. As observed in panel (a), a maximum yield of 82.2% (green circles) for C₆₊ alcohols in the final product of the process can be achieved when operating at a contact time of 5.23 h kg_{cat} kmol_{gas}^{−1} and a H₂:ester ratio of 400:1 in the Guerbet and hydrogenolysis reactors, respectively. The ester yield under such conditions is calculated as 3.6%. In our process, the ester yield needs to be lower than 5% to decrease their negative effect in the downstream etherification reactor.^{12,13} In Fig. 8(b) it is observed that reduction of the H₂:ester ratio to 200:1 in the hydrogenolysis reactor increases the ester yield up to 4.7% at the assessed contact time, going even higher than 5% at the 100:1 ratio regardless of the contact time as presented in panel (c). Remarkably, the maximum yield of C₆₊ alcohols also decreases with reduction of the H₂:ester ratio, showing a maximum value of 75.6% at 6 h kg_{cat} kmol_{gas}^{−1} of contact time in the alcohol coupling reactor and a H₂:ester ratio of 100:1 in the hydrogenation unit.

Additional evaluation of the process performance is obtained by assessing the relative selectivity of linear to branched alcohols in the outlet stream of the presented technology. Fig. 9 displays the relative selectivity of linear to branched C₆ and C₈ alcohols. As observed in all cases, short residence times in the coupling reactor favors the production of hexanol over 2-ethylbutanol, whilst the opposite effect is seen for 1-octanol and 2-ethylhexanol. Increasing the contact time in the coupling reactor causes a decrease in the ratio of the C₆ alcohols and increases the one of their C₈ analogue. In general, these relative selectivities change within the first 4.5 h kg_{cat} kmol_{gas}^{−1} regardless of the assessed scenario, plateauing between 4.3–3.5 and 1.1–1.5 for the C₆ and C₈ ratios, respectively.

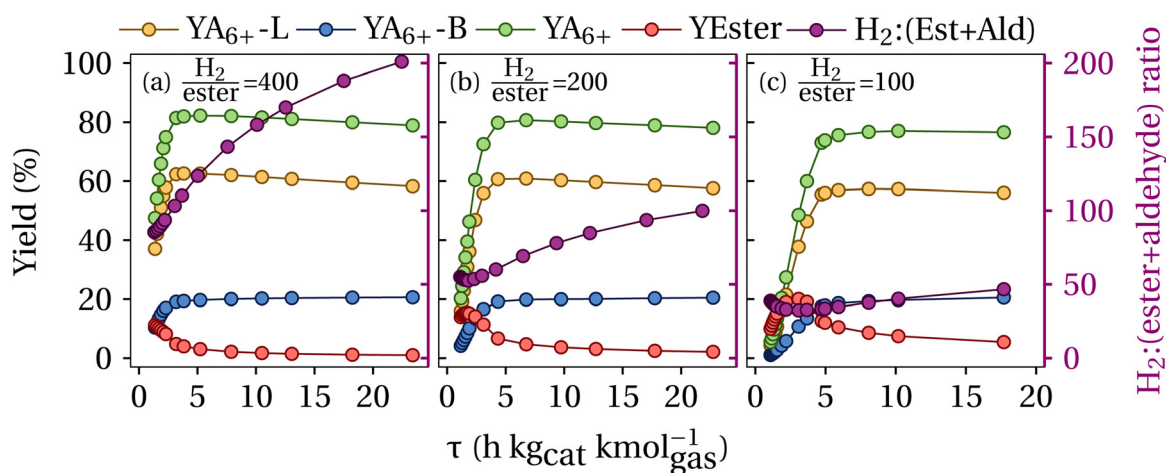


Fig. 8 Product yields (PR-1) as a function of contact time (τ) in the Guerbet coupling reactor and H₂:ester ratios of (a) 400:1, (b) 200:1 and (c) 100:1. YA₆₊-L: carbon yield of C₆₊ linear alcohols, YA₆₊-B: carbon yield of C₆₊ branched alcohols, YA₆₊: carbon yield of C₆₊ alcohols, and YEster: carbon yield of esters. Scattered data represent the results from the simulated scenarios. Continuous lines are drawn to guide the eyes.

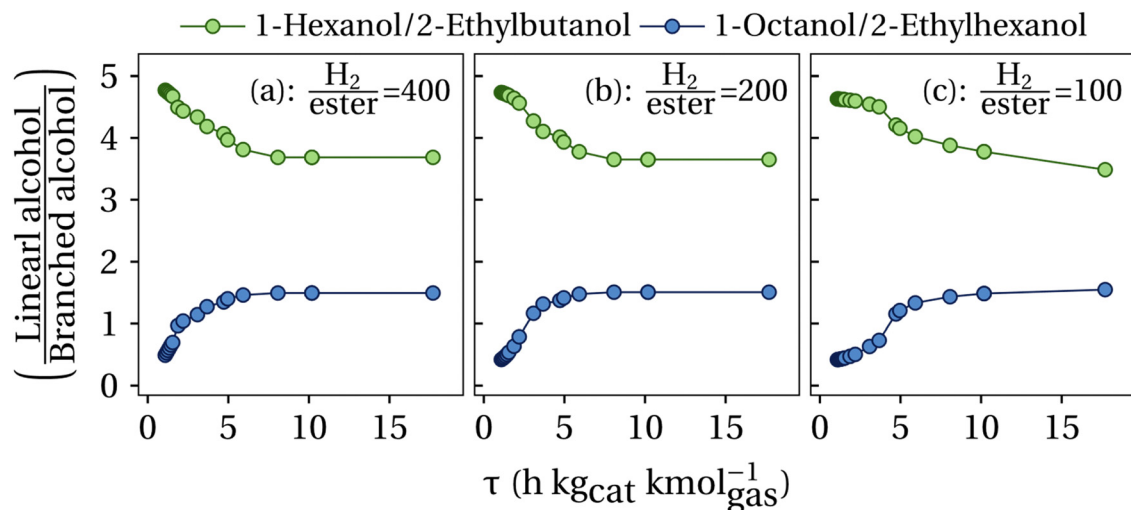


Fig. 9 Ratio of linear to branched alcohols as a function of contact times (τ) in the Guerbet coupling reactor and H_2 : ester ratios of (a) 400 : 1, (b) 200 : 1 and (c) 100 : 1. Scattered data represent the results from the simulated scenarios. Solid lines are drawn to guide the eyes.

In a second process evaluation phase, the influence of the amount of catalyst and the discharging of butanol from the separation column S-104 (see Fig. 2) on the alcohol product yields is considered. The top product stream of the column S-104 is split into the stream PR-2 and the recycling stream fed back to the Guerbet coupling reactor. The split percentage was assessed between 0 and 20%, where 0 represents no flow

through PR-2. $Y_{A4-PR-2}$ is indicated here as the yield of butanol in the discharged stream at the top product of the separation column S-104. The yields of the higher alcohols and esters, as well as the cumulated yield of alcohols higher than carbon number 6 (Y_{A6+}) and the ethanol to butanol ratio upstream of the Guerbet coupling reactor are given for a low (15 tons) and high (60 tons) catalyst load and the discharge ratio of 0 to 20% in Fig. 10, respectively.

As presented in Fig. 10(b), for a low catalyst load of 15 tons the butanol discharge (between 0–20%) in the distillation column S-104 has no significant impact on the overall A_{6+} yield, remaining stable within 45–49% and exhibiting its maximum at a discharge ratio of 2.5%. Nevertheless, stronger variations for the ethanol–butanol feed ratio were observed, with ratios ranging between 65 and 82%. It is also observed in Fig. 10(a), that the formation of esters decreases for a higher butanol discharge ratio and undershoots the desired 5% level at a butanol discharge ratio higher than 5%.

The yields for a higher catalyst load of 60 tons are presented in Fig. 10(c) and (d). It is observed that the butanol discharge has a negative impact on the A_{6+} yield, resulting in a maximum of 81% without butanol discharge, and decreasing to 61% at a discharge ratio of 20%. Remarkably, the requirement of limiting the yield of esters to maximum 5% is achieved for all butanol discharge ratios. In this view, at higher contact times no butanol discharge in S-104 increases the yield of C_{6+} alcohols.

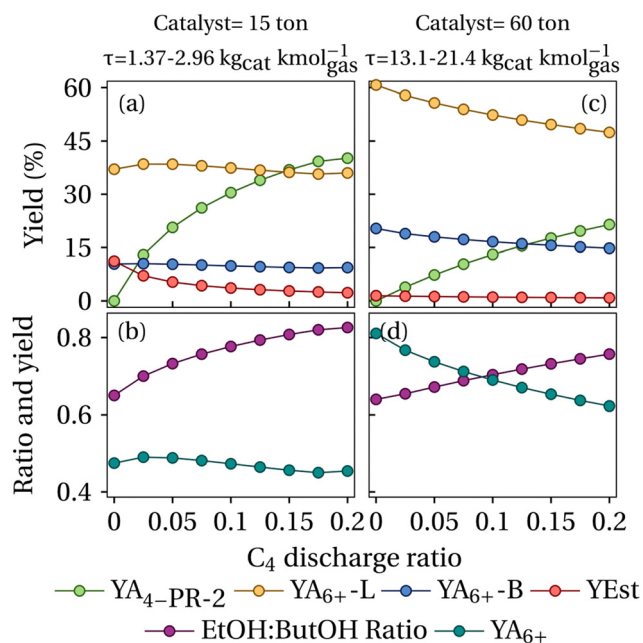


Fig. 10 Alcohols, esters, and aldehydes yields, calculated according to eqn (8), as well as the ethanol to butanol ratio and cumulated C_{6+} yield in the product stream P-103 for a catalyst load of (a) and (b) 15 tons and (c) and (d) 60 tons and a H_2 : ester ratio of 400 : 1. $Y_{A4-PR-2}$: carbon yield of butanol in stream PR2, Y_{A6+-L} : carbon yield of C_{6+} linear alcohols, Y_{A6+-B} : carbon yield of C_{6+} branched alcohols, Y_{Est} : carbon yield of esters, and Y_{A6+} : carbon yield of C_{6+} alcohols. Scattered data represent the results from the simulated scenarios and the continuous line is drawn to guide the eyes.

4. Conclusions

A model for the kinetics of alcohols coupling with Guerbet chemistry over a $Cu_{0.01}Mg_{2.99}AlO_x$ catalyst was developed. Catalytic experiments between 0–21.6 h kg_{cat} kmol_{gas}^{−1} were performed in a gas phase continuous flow reactor at 325 °C and 300 psig by feeding either pure ethanol or an ethanol–butanol blend under a hydrogen atmosphere. Experimentally at least 60 different chemical species were identified in the



The optimized operating conditions of the Guerbet coupling reactor include a contact time of $5.23 \text{ kg}_{\text{cat}} \text{ kmol}_{\text{gas}}^{-1}$, where the C_{6+} alcohols yield is maximized at 82.2%. Partial discharge of the butanol leaving the S-104 separation stage has no positive influence on the C_{6+} alcohols yield, whereby all the butanol is recycled upstream of the Guerbet reactor. A hydrogenolysis reactor needs to be incorporated and operated with a H_2 :ester ratio of 400:1 to achieve an ester yield of 3.6%, which is below the permissible limit of 5%. Lower H_2 :ester ratios lead to a shift in terms of higher contact times, which result in a lower C_{6+} alcohol yield. The kinetic model developed herein provides a good basis for future process optimization and techno-economic evaluations.

JEC: experiments, data interpretation, kinetic modeling programming and fitting, paper conceptualization, writing; CM: simulations, data interpretation, kinetic modeling programming and fitting, paper conceptualization, writing; EV: experiments; ML: paper conceptualization, funding acquisition, writing, project managing, GWH: paper conceptualization, funding acquisition, writing, project managing. The manuscript was written through contributions of all authors. All authors have given approval to the final version of the manuscript.

All the data for this article are available in the main text and the ESI.[†]

There are no conflicts to declare.

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