

Modelling the Effects of Reaction Temperature and Flow Rate on the Conversion of Ethanol to 1,3-Butadiene

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ABSTRACT

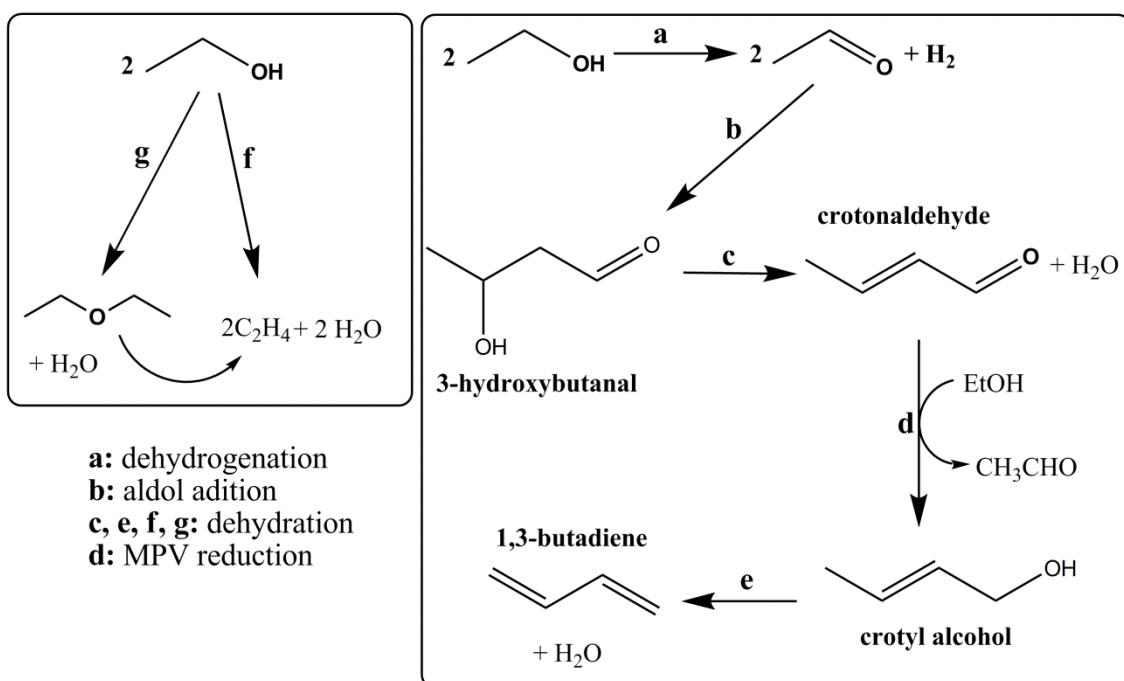
A full factorial experimental design was performed to investigate the conversion of ethanol to 1,3-butadiene (1,3-BD), through manipulation of the reaction temperature and ethanol weight hourly space velocity. Reactions were carried out in presence of the catalyst $K_2O:ZrO_2:ZnO/MgO-SiO_2$, prepared by co-precipitation methods. Mathematical models were developed to correlate observed product selectivities, 1,3-BD yields and productivities with the manipulated reaction variables, allowing for quantification of variable effects on catalyst activity and assessment of the kinetic mechanism. Obtained 1,3-BD productivities were as high as $0.5 \text{ g}_{\text{BD}}/\text{g}_{\text{cat.h}}$, with 1,3-BD yields of 27 %. Results suggest that acetaldehyde condensation is the rate determining step.

Keywords: Ethanol; 1,3-butadiene; silica-magnesia catalyst; experimental design; kinetics.

1. Introduction

1,3-Butadiene (1,3-BD) is a valuable conjugated diene used for production of synthetic polymers [1,2]. The conversion of ethanol into 1,3-BD constitutes a promising alternative for the current conventional 1,3-BD production process, where 1,3-BD is obtained as a co-product during ethene manufacture in steam crackers [3-5], responsible for significant amounts of CO₂ emissions [6]. Recently, Patel and co-workers have compared the performances of biobased and conventional routes for 1,3-BD production, providing positive evidences regarding the technical and economical sustainability of the new proposed process [4,5].

The conversion of ethanol to 1,3-BD has been known since the beginning of the 20th century, although only in the last decade has it received more significant attention from the academic community [7,8]. In particular, it has been established that the ideal catalyst for ethanol to 1,3-BD must present suitable amounts, in terms of strength and distribution, of acid and basic sites, since the widely accepted reaction pathway, Scheme 1 (a-e), involves different consecutive steps: (i) ethanol dehydrogenation, (ii) aldol condensation, (iii) Meerwein-Ponndorf-Verley (MPV) reduction of crotonaldehyde and (iv) crotyl alcohol dehydration [9-15]. However, ethanol dehydration to ethene and diethyl ether (DEE) at acid sites leads to undesired reaction products, Scheme 1 (f-g). As a consequence, much work has been dedicated to catalyst optimisation with the aim of minimising the extent of the parallel reactions and enhancing 1,3-BD yield [1,8].



Scheme 1: Scheme illustrating of the potential reaction network.

Among the many catalysts studied, the use of MgO-SiO₂ systems containing different metals and/or metal oxides based on Cu, Zr, Zn and Ag have shown encouraging performances, related mainly to the distinct Lewis acid and Brønsted basic sites of such materials [12,16-19]. In particular, the method and the Mg-to-Si molar ratio employed to synthesise the MgO-SiO₂ precursor can be very important, as it can significantly modify the catalyst properties [17,18,20,21]. Besides, as reported previously, the optimum Mg-to-Si ratio may depend on the employed preparation procedure [12,18,20]. It is important to note that the co-precipitation method has been used successfully in the synthesis of Zr,Zn-containing MgO-SiO₂ systems that are able to deliver high 1,3-BD productivities (in g_{BD}/g_{cat}·h) [2]. Whereas ZrO₂ and ZnO introduce synergic effects that facilitate the ethanol dehydrogenation, aldol condensation and the MPV reduction [11,13,18,19,22,23], the co-precipitation method allows for efficient formation of Mg-O-Si bonds and homogeneous distribution of the distinct elements on the catalyst surfaces [2,11,18]. Besides, modification of catalyst acidity with alkali metals constitutes an attractive solution to minimise the undesired ethene and DEE by-products [2,23]. For instance, the combined selectivity for 1,3-BD and acetaldehyde (AcH) has been reported as 72 %, at ethanol conversion of 26 % and 1,3-BD yields of 27 % over a K₂O:ZrO₂:ZnO/MgO-SiO₂ system [2].

On the other hand, the effects of common reaction variables, such as temperature and contact time, on ethanol to 1,3-BD reaction performance have received much less attention [7]. Some previous studies have investigated the kinetic aspects of this reaction, without sufficient support of statistical analyses [24,25]. However, in order to improve the catalyst properties it may be necessary to understand the kinetic mechanism and identify the rate-limiting step of the ethanol to 1,3-BD reaction, since both ethanol dehydrogenation [9,17,19,26] and aldol condensation [13,19,26,27] steps have been described as the rate-limiting step. Some previous studies point out that crotyl alcohol might not be an intermediate in the reaction [14]; others, however, suggest that 1,3-BD is produced from the dehydration of crotyl alcohol, which is formed through the reaction between an activated form of ethanol and acetaldehyde, not involving acetaldehyde condensation [20,28]. Thus, it can be said that the ethanol to 1,3-BD reaction mechanism is still subject to debate.

Usually, reaction variables are investigated using the “change-one-factor-at-a-time” method; that is, one variable is changed while the other experimental conditions are kept constant [2,10,12,13,26,27,29-31]. Based on this method, it has been found that temperature exerts an important nonlinear effect on 1,3-BD yields, as observed with help of different catalysts [13,19,26,30,31]. Similar studies have shown that the weight hourly space velocity (WHSV) also modifies 1,3-BD yields and selectivities [2,13,19,26,27,30,31]. Usually, enhancement of 1,3-BD selectivities [13,19] and yields [19,27] could be observed when the WHSV was reduced, suggesting the positive effect of contact times on reaction yields, as one might already expect. However, nonlinear effects of ethanol flow rates on 1,3-BD yields were verified over different single and binary metal oxides (such as MgO, ZrO₂, Al₂O₃-MgO, Al₂O₃-Fe₂O₃, Al₂O₃-Cr₂O₃ and ZrO₂-Fe₂O₃) [30,31].

One of the main drawbacks associated with the “change-one-factor-at-a-time” method is the fact that the influence observed for the particularly analyzed variable may not be the same when some of the remaining experimental conditions change [32]. This can occur because variables may interact with each other, resulting in unexpected nonlinear effects [32,33]. Such interaction effects can only be identified when variables are investigated and manipulated simultaneously. Statistical experimental design techniques, such as factorial designs, overcome this drawback, allowing for identification and quantification of the distinct main variable effects and variable interaction effects. Besides, the use of statistical experimental design can lead to maximisation of the information content of the experimental data set, as variable effects can be computed with minimum uncertainty [34,35]. In this case, mathematical models can be developed to correlate with maximum efficiency the independent (such as reaction temperature, feed concentration and contact time) and dependent variables (such as ethanol conversion and 1,3-BD selectivity), making data interpretation easier and more robust.

Although the mechanistic kinetic modelling of the reaction system is usually desirable, requiring the definition of fundamental rate equations and estimation of kinetic parameters and equilibrium constants [36,37], it is well-known that the phenomenological approach usually leads to very large number of model parameters, making their estimation difficult (and many times impossible) [35,37,38]. For this reason, the empirical modelling of reaction data, with help of sound statistical tools, may be much more efficient for analysis and optimisation of complex catalytic

processes, as the use of empirical mathematical tools is relatively simple and much less time consuming [35,39].

In a previous work, preliminary kinetic investigation of the ethanol to 1,3-BD conversion over MgO-SiO₂ catalyst system was performed. Acetaldehyde condensation was identified as the rate-limiting step between 300 and 400 °C, but detailed analyses of variable effects were not performed [40]. In the present study, the effects of temperature and WHSV on process performance are investigated with help of rigorous statistical experimental design tools, which allow for identification and quantification of variables effects on catalytic activity. Catalyst performance was characterised in terms of the product distributions, 1,3-BD yields and productivities. These dependent variables were correlated with reaction temperatures and ethanol flow rates in order to characterize the process performance. Particularly, the experimental study was also performed with the highly active K₂O:ZrO₂:ZnO/MgO-SiO₂ catalyst system [2], reinforcing that acetaldehyde condensation is the rate-limiting step within the investigated temperature range.

2. Materials and Methods

2.1 Catalyst Preparation and Characterisation

The catalyst was prepared by co-precipitation with the Mg-to-Si molar ratio equal to 1.0. In a typical synthesis, 9.01 g of SiO₂ (Sigma-Aldrich, 99.8 %) were dissolved in 100 mL of 1.2 M NaOH (Sigma-Aldrich, 99 %) solution. The mixture was heated at 60–80 °C under vigorous stirring until complete SiO₂ dissolution. The solution was cooled and 42.4 g of Na₂CO₃ (Sigma-Aldrich, 99.9 %) were added. A Mg(NO₃)₂•6H₂O (Sigma-Aldrich, 99 %) solution was added drop-wise into this mixture whilst stirring at 25 °C (38.85 g of Mg(NO₃)₂•6H₂O in 200 mL). The pH was maintained at 10.5 by adding appropriate quantities of 1.2 M NaOH solution and, at the end of the process, the solution volume was adjusted to 600 mL with deionized water. The resulting mixture was stirred for 2 h and aged for 22 h at 25 °C. Finally, the mixture was filtered and washed with 7.5 L of hot water. The precipitate was dried in static air at 80 °C for 24 h before grinding.

In order to produce materials with 1.5 wt% of Zr(IV) and 0.5 wt% of Zn(II), 0.57 g of ZrO(NO₃)₂•H₂O (Sigma-Aldrich, 99 %) and 0.24 g of Zn(NO₃)₂•6H₂O (Sigma-Aldrich, 98 %) were dissolved in 50 mL of water. The

solution was then added to 10 g of the MgO-SiO₂ system, dried under stirring and calcined in air at 500 °C for 5 h (5 °C/min). Finally, the appropriate volume of 0.4 M KOH (Sigma-Aldrich, 90 %) solution was added to the calcined material drop-wise to generate the final catalysts with 1.2 % weight of potassium. The mixture was stirred for 1 h at 25 °C before drying at 80 °C for 5.5 h. The catalyst preparation procedure is illustrated in Figure SI1 in the Supporting Information (SI).

Catalyst samples were characterised by static N₂ adsorption at -196 °C, scanning electron microscopy with energy dispersive X-rays (SEM-EDX), powder X-ray diffraction (pXRD), ²⁹Si solid-state MAS NMR and temperature programmed desorption of ammonia (NH₃-TPD) as described elsewhere [2]. Bulk loadings were confirmed by inductively coupled plasma optical emission spectroscopy (ICP-OES), leading to K, Zn and Zr loadings of 1.31, 0.57 and 1.68 wt%, respectively. Thermogravimetric analyses of used catalysts were carried out in a Setsys Evolution TGA Setaram system. Samples (20 mg) were heated from room temperature to 1000 °C under air flow (100 mL/min), using a heating rate of 20 °C/min. (The interested reader should refer to reference [2] for further characterisation of this family of catalysts, including characterisation of the acid/base properties.)

2.2 Catalytic Tests

Catalytic tests were carried out in a flow quartz packed-bed reactor at atmospheric pressure. Argon was used as carrier gas (8 mL/min). The ethanol WHSV was varied within 0.3-2.5 h⁻¹ through modification of the ethanol flow rate, keeping catalyst mass and carrier gas flow rate fixed. The investigated WHSV range corresponded to ethanol molar fractions between 0.41 and 0.85. The contact time (calculated as the ratio between the catalyst volume and the total gas flow at the reaction temperature) ranged from 1.3 to 5.3 s. Reaction temperature ranged from 300 to 400 °C. Both WHSV and temperature ranges are consistent with the majority of catalysts disclosed in the literature. The exhaust gases were analysed after 3 h of time on stream (TOS) via GC-MS on an Agilent 7890A instrument, equipped with a HP-PLOT/Q column of length 30 m and diameter 0.53 mm and FID/MS detectors. The GC was calibrated as detailed elsewhere [11]. Carbon balances were typically better than 85 %.

Ethanol conversion (X), selectivity (S), 1,3-BD yield (Y_{BD}) and 1,3-BD productivity (P_{BD} , in $\text{g}_{\text{BD}}/\text{g}_{\text{cat}} \cdot \text{h}$) were computed as described in Equations (1), (2), (3) and (4), respectively. $N_{EtOH,in}$ and $N_{EtOH,out}$ represent the number of mols of ethanol that were added and collected, respectively. N_i represents the number of mols of the product i , while NP is the total number of products, m_{cat} is the catalyst mass and t is the total reaction time.

$$X(\%) = \frac{(N_{EtOH,in} - N_{EtOH,out}) \cdot 100}{N_{EtOH,in}} \quad (1)$$

$$S_i(\%) = \frac{N_i}{\sum_{i=1}^{NP} N_i} \cdot 100 \quad (2)$$

$$Y_{BD}(\%) = \frac{2 \cdot N_{BD}}{N_{EtOH,in}} \cdot 100 \quad (3)$$

$$P_{BD} = \frac{N_{BD} \cdot 54}{(m_{cat} \cdot t)} \quad (4)$$

2.3 Experimental Design

The effect of the experimental reaction variables, temperature and WHSV, on the catalyst performances were investigated with help of a two-level factorial design, with four central point experiments. The statistical design approach was selected to allow for the simultaneous and precise quantification of the main effects, quadratic effects and interaction effects of temperature and WHSV on catalyst activity [32,33,35,41]. The experimental variables, z_i , were normalised within the $[-1,+1]$ interval, according to Equation (5). z_i represents the actual value of variable i , z_{ic} denotes the actual value of variable i at the central condition (equal to $350\text{ }^{\circ}\text{C}$ and 0.93 h^{-1} for temperature and WHSV, respectively), Δz_i is equal to $25\text{ }^{\circ}\text{C}$ and 0.31 h^{-1} for temperature and WHSV, respectively, and x_i is the normalised value of variable i .

$$x_i = \frac{z_i - z_{ic}}{\Delta z_i} \quad (5)$$

Table 1 shows the experimental design matrix, with normalised and actual values of reaction conditions. Four experiments at central condition (Exps. 5-8, Table 1) were carried out in order to evaluate the experimental error and to test for the evidence of

non-linear effects. Two additional axial experiments, -2 and +2, were performed for each variable (Exps. 9-12), to allow for improved quantification of nonlinear effects [32]. Besides, Experiments 13 and 14, at conditions [-1,0] and [+1,0], were performed to evaluate the prediction capability of the proposed models. Finally, additional experiments were performed to assess the system behaviour at higher WHSV values (Exps. 15-18). As a consequence, the performances of all proposed models were validated statistically in the range of interest, as illustrated by Tables SI2-SI4 in the Supporting Information.

Table 1. Matrix of experimental conditions: actual and normalised variable values.

Experiment	Temperature (°C)		WHSV (h ⁻¹)
	$z_1 (x_1)$	$z_2 (x_2)$	
1	325 (-1)	0.62 (-1)	
2	325 (-1)	1.24 (+1)	
3	375 (+1)	0.62 (-1)	
4	375 (+1)	1.24 (+1)	
5	350 (0)	0.93 (0)	
6	350 (0)	0.93 (0)	
7	350 (0)	0.93 (0)	
8	350 (0)	0.93 (0)	
9	300 (-2)	0.93 (0)	
10	400 (+2)	0.93 (0)	
11	350 (0)	0.31 (-2)	
12	350 (0)	1.55 (+2)	
13	325 (-1)	0.93 (0)	
14	375 (+1)	0.93 (0)	
15	325 (-1)	2.49 (+5)	
16	350 (0)	2.49 (+5)	
17	375 (+1)	2.49 (+5)	
18	400(+2)	2.49 (+5)	

Ethanol conversions, product selectivities, molar fractions of reaction products, 1,3-BD yields and 1,3-BD productivities were selected as response or dependent variables to assess reaction temperature and WHSV effects. Models with the general form of Equation (6) - a classic model structure used to analyze experimental data collected in factorial designs [32,33,35,41] - were then applied to correlate dependent variables, y_i , with the reactions conditions, using the independent normalised variables, x_i . At this point, it must be noted that WHSV values were modified by varying ethanol

flow rate only. This means that contact time and ethanol composition were modified simultaneously, as WHSV values were modified. Thus, in the following discussion it is necessary to keep in mind that WHSV influence is related to both contact time and ethanol composition effects.

$$y_i = b_0 + b_1 x_1 + b_2 x_2 + b_{12} x_1 x_2 + b_{11} (x_1^2 - \lambda_1) + b_{22} (x_2^2 - \lambda_2) \quad (6)$$

The parameters of Equation (6), b_1 and b_2 , b_{12} , b_{11} and b_{22} , are related to the linear, interaction and quadratic effects of temperature and WHSV, respectively, while b_0 is the independent bias parameter. Finally, λ_i is a constant used to guarantee the orthogonality of the design matrix, calculated as shown in Equation (7), where NE is the total number of experiments. Parameters from Equation (6) were estimated with help of the well-known least-squares estimation procedure. Statistical significance of estimated model parameters was evaluated with the standard t-test. Whenever parameter significance was lower than 5%, the parameter and respective variable effect were regarded as statistically insignificant and were removed from Equation (6). Additionally, the quality of the model fit was always validated by comparing experimental variances with model prediction variances using the standard F-test [42], in order to avoid overfitting. For all models obtained, experimental variances were always statistically similar to model prediction variances, supporting the satisfactory statistical quality of the models.

$$\lambda_i = \frac{1}{NE} \sum_{j=1}^{NE} x_{ij}^2 \quad (7)$$

3. Results and Discussion

3.1 Catalysis Characterisation

The BET surface area of the catalyst was equal to 305 m²/g. In order to minimise internal pore diffusion limitations, catalyst particles were grounded until sizes smaller than 200 µm were achieved. The elements were uniformly distributed on the surface of the catalyst, as indicated by the SEM-EDX elemental mapping, Figure 1, and the elemental dispersions at specific locations of the catalyst particle, Table 2.

Table 2 – Surface elemental dispersion of catalyst sample in wt %^[a].

Mg	Si	Zr	Zn	K
37.42 ± 0.50	52.79 ± 0.49	5.44 ± 0.55	1.69 ± 0.31	2.66 ± 0.20

^[a] Elemental values were normalised to 100 and represent a dispersion measure only.

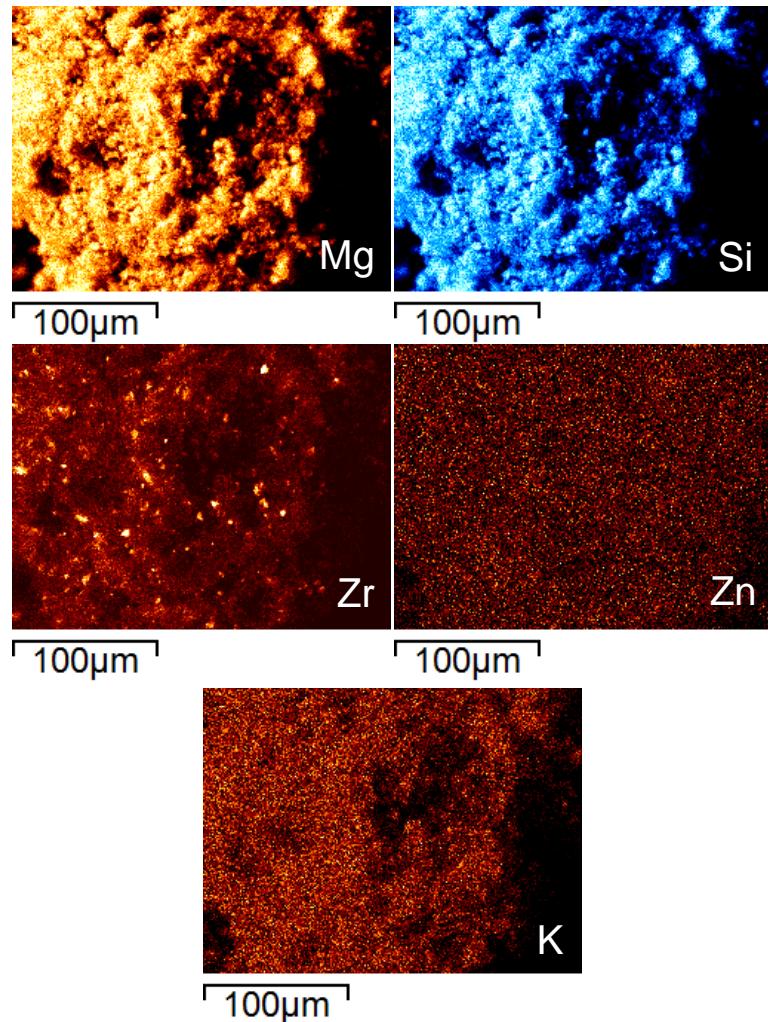


Figure 1 – Elemental mapping of the $\text{K}_2\text{O}:\text{ZrO}_2:\text{ZnO}/\text{MgO}-\text{SiO}_2$ catalyst.

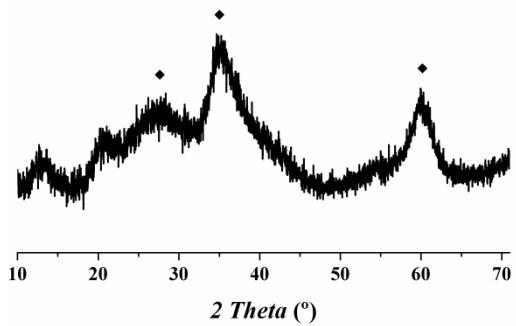


Figure 2 – XRD pattern of the $\text{K}_2\text{O}:\text{ZrO}_2:\text{ZnO}/\text{MgO}-\text{SiO}_2$ catalyst sample. (♦) denotes magnesium silicate hydrate structure [43,44].

Moreover, the catalyst sample was amorphous, as characterised by pXRD as shown in Figure 2. The broad diffraction bands (at 25-30, 33-39 and 58-62 °) can be assigned to the magnesium silicate hydrate structure [43,44]. ZrO_2 , ZnO and K_2O were efficiently dispersed into the -Mg-O-Si- network and could not be detected through the pXRD analysis.

In a previous study [2], the effects of the Mg-to-Si molar ratio of $\text{MgO}-\text{SiO}_2$ and ZrZn -containing $\text{MgO}-\text{SiO}_2$ systems prepared by co-precipitation on the ethanol to 1,3-BD conversion were investigated. Zr and Zn contents were kept fixed since it was shown that loadings of 1.5 and 0.5 wt% for Zr and Zn, respectively, produced higher selectivities towards 1,3-BD [11]. A beneficial effect on 1,3-BD yield and selectivity was observed as Zr and Zn were added on the $\text{MgO}-\text{SiO}_2$ precursor [2]. However, the ethene and DEE selectivities were also high.

IR measurements performed after NH_3 adsorption experiments suggested the increase of the catalyst Lewis acidity in ZrZn -containing $\text{MgO}-\text{SiO}_2$ samples [2]. Catalyst doping with alkali metals, especially potassium, was shown to reduce this effect, leading to suppression of ethanol dehydration to ethene and DEE. It was rationalised that the addition of alkali metal neutralised the strong acid sites of the catalyst, responsible for ethanol dehydration [2,23]. The alkali metal loading of 1.2 wt.% was shown to be the most suitable for this system: whereas higher alkali metal loadings resulted in lower ethanol conversion and 1,3-BD yields, lower loadings produced more ethene and DEE [2]. Moreover, ^{29}Si NMR experiments indicated that addition of alkali metal and Zr and Zn did not change the silicon environments [2], supporting the hypothesis that the Lewis acid sites participate in the overall reaction pathway.

In the present work work, we have also characterized the silicon environments of the K₂O:ZrO₂:ZnO/MgO-SiO₂ material by ²⁹Si NMR analysis, as shown in Fig SI2. The two broad resonances with maxima around -85 and -94 ppm suggest the presence of -Mg-O-Si- linkages, in line with the pXRD pattern. Chemical shifts between -85 and -89 ppm and between -92 and -99 ppm were already reported for magnesium silicate systems and they were attributed to Q² and Q³ species, respectively, as Si*(OMg)(OSi)₂(OH) and Si*(OMg)(OSi)₃ [26,43,45].

The role of potassium in the catalyst system has been evaluated through NH₃-TPD experiments, Fig SI3, which confirmed a reduction in the total catalyst acidity as the alkali metal was added, as expected. It should be emphasised, however, that to determine the active sites of the employed catalyst is beyond the scope of this work, which pursue to rigorously quantify reaction variables temperature and WHSV effects on the catalyst performance. Thus, the K₂O:ZrO₂:ZnO/MgO-SiO₂ system was selected for this study because of its high selectivity to 1,3-BD. Nevertheless, active sites could be related to weak Lewis acid-Brønsted basic pairs distributed throughout catalyst surface, which could involve, for instance, Mg-O, Zn-O, Zr-O pairs with different metal coordination environments.

3.2 Catalytic Tests

Catalyst deactivation was evaluated through a 48 h reaction test. The selectivities for the main carbon containing products are presented in Figure 3(a). The slight reduction of the 1,3-BD selectivity with the time on stream was observed, which may be related to catalyst deactivation, potentially via coking. However, the observed variability of product selectivities during the initial reaction hours were similar to the experimental fluctuations observed at the central point experiments, indicating that catalyst performances were not affected significantly by catalyst deactivation.

The recyclability of the catalyst was also investigated. In order to assess this point, the catalyst used during the 48 h reaction test was regenerated (re-calcined in air) and re-tested for an additional period of 25 h, as shown in Figure 3(b). Catalyst recycling did not restore the full catalyst activity; in spite of that, the recycled catalyst showed constant and high selectivity to 1,3-BD during the test procedure.

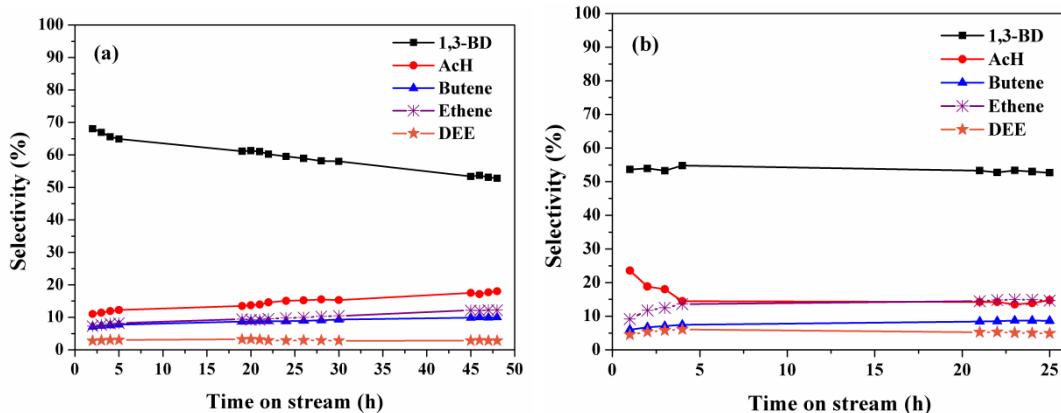


Figure 3 - Effect of time on stream on the selectivity of the main carbon containing products: (a) fresh catalyst; (b) recycled catalyst. ($T = 350\text{ }^{\circ}\text{C}$, WHSV = 0.62 h^{-1} , contact time = 3.8 s).

The designed experimental conditions afforded ethanol conversions ranging from 7 to 44 %. The main carbon containing products were 1,3-BD, acetaldehyde (AcH), ethene, diethyl ether (DEE) and butene (1-butene, *cis*- and *trans*-2-butene). Other minor products were propene, propane, ethane, acetone, with combined selectivities below 5 %. Besides, traces of C5 and C6 compounds could also be detected. Table 3 shows ethanol conversions, selectivities of the main carbon containing products, 1,3-BD yields and productivities obtained in the designed experiments. Additional information regarding contact time, ethanol molar fractions in the feed, molar fractions of the main products and carbon balances (typically greater than 85 %) are reported in Table SI1 as Supporting Information.

Table 3 - Catalytic results over the K₂O:ZrO₂:ZnO/MgO-SiO₂ catalyst, for 3 h of time on stream.

Exp.	T (°C)	WHSV (h ⁻¹)	X (%)	Selectivity (mol%)				1,3-BD yield ^[a] (mol%)	1,3-BD productivity (g _{BD} /g _{cat} ·h)
				1,3- BD	AcH	Ethene	DEE		
1	325	0.62	17.0	65.9	15.0	5.6	3.3	7.3	14.8 0.06
2	325	1.24	16.4	49.4	37.1	4.3	2.8	4.2	9.0 0.08
3	375	0.62	25.8	54.5	21.0	8.8	2.8	8.4	28.5 0.13
4	375	1.24	26.2	48.8	32.7	7.4	2.5	5.5	22.4 0.20
5	350	0.93	23.6	51.3	31.2	6.1	2.9	5.6	18.1 0.12
6	350	0.93	26.2	52.1	30.5	5.9	2.9	5.7	18.3 0.12
7	350	0.93	32.9	49.6	33.6	5.8	2.9	5.2	19.5 0.13
8	350	0.93	29.0	58.7	22.2	7.0	2.8	6.2	17.9 0.12
9	300	0.93	13.4	55.2	31.5	3.9	3.0	4.2	4.9 0.03
10	400	0.93	41.8	50.1	22.0	11.2	2.4	9.1	31.4 0.21
11	350	0.31	43.6	63.4	14.6	7.2	2.6	8.0	26.1 0.05
12	350	1.55	13.6	51.7	32.4	5.9	2.6	4.9	14.6 0.16
13	325	0.93	21.7	48.2	38.6	3.7	3.5	4.0	8.2 0.05
14	375	0.93	39.1	54.3	25.9	7.2	2.8	6.4	25.3 0.16
15	325	2.49	6.5	33.3	57.3	3.5	2.3	2.3	5.5 0.10
16	350	2.49	9.9	41.4	45.5	5.1	2.5	3.6	13.5 0.25
17	375	2.49	12.3	43.4	40.6	7.0	2.2	4.4	20.0 0.37
18	400	2.49	19.1	46.4	33.0	9.3	2.3	5.7	26.8 0.49

^[a] Calculated as Eq. (3).

The possible existence of mass transfer limitations was discarded from the analysis through estimation of the apparent activation energy [2,46], as shown in Figure SI2 of the Supporting Information. Since total flow changes with WHSV, the assessment of mass transfer limitations was performed using the central condition (0.93 h⁻¹) for this variable, in order to obtain such evaluation at an average WHSV. Catalytic results were also shown to be far from equilibrium conditions, as supported by calculated equilibrium compositions at the analyzed reaction conditions [47], Figure SI3. Moreover, catalysts used in Experiments 7 and 10 (Table 3) were analyzed after catalysis by thermogravimetric analysis in order to characterise possible carbon formation on the catalyst surface. Weight losses of 10.1 and 9.5 % were attributed to

carbon formation during reactions performed at 400 and 350 °C, respectively, as shown in Figure 4.

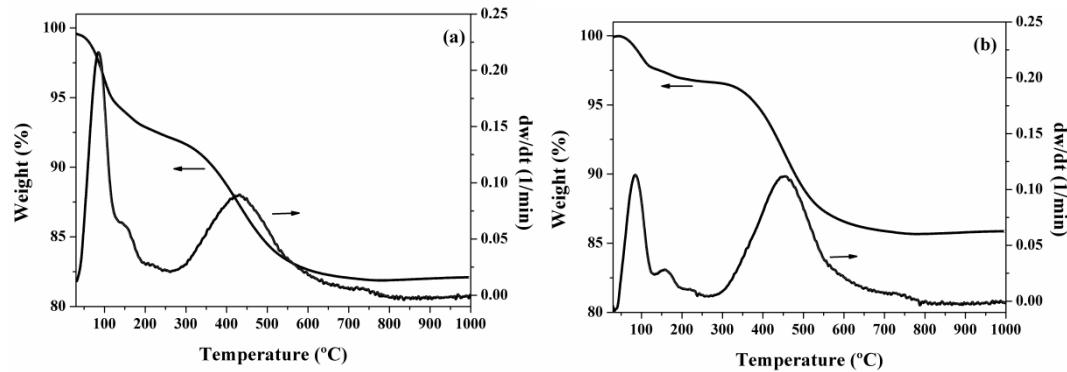


Figure 4 -Thermogravimetric analysis of used catalysts in Experiments 7 (a) and 10 (b). Only weight loss above 280 °C was attributed as carbon formation.

Ethanol conversions, product selectivities, 1,3-BD yields and productivities were correlated to reaction variables utilising Equation (6). For ethanol conversions, Figure 5 shows experimental data and values calculated with the empirical model described in Equation (8), where lines denote constant WHSV values. The experimental standard deviation is equal to 4.0 % and the linear correlation coefficient is equal to 0.89. As expected, ethanol conversion increased with reaction temperature and decreased with ethanol WHSV [2]. It is important to emphasise that higher WHSV values correspond to higher ethanol molar fractions in the feed and lower contact times.

$$X = (26.68 \pm 1.41) + (6.02 \pm 1.23) \cdot x_1 - (3.68 \pm 0.57) \cdot x_2 \quad (8)$$

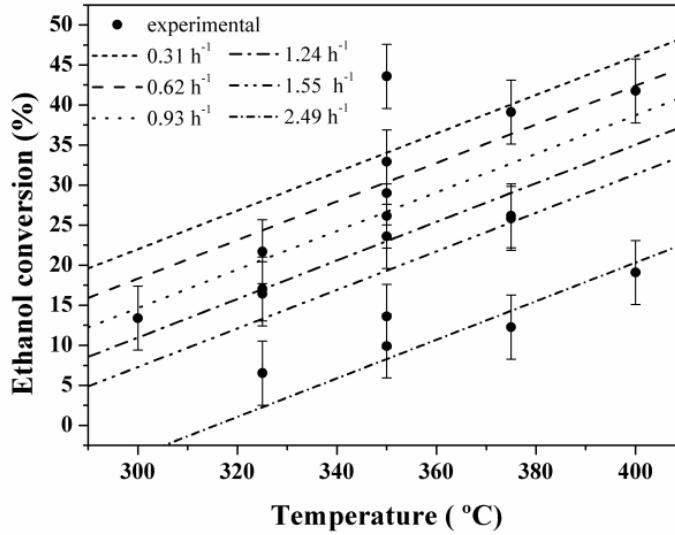


Figure 5 - Experimental values (●) and empirical model for ethanol conversion ($r = 0.89$). Lines represent constant ethanol WHSV.

Figure 6(a) shows the experimental ethene selectivities and the values calculated with the model described in Equation (9), which has a linear correlation coefficient of 0.97. Ethene selectivity ranged from 3.5 to 11.2 % and has a standard deviation of 0.57 %. Temperature was the most influential variable on ethene selectivity, presenting a linear and quadratic influence within the investigated experimental region. WHSV, in turn, showed a linear and negative effect on the ethene selectivity, suggesting that reduction of contact time (or richer ethanol molar fraction in the feed) can cause the decrease of the ethene selectivity.

$$S_{Ethene} = (6.44 \pm 0.15) + (1.72 \pm 0.13) \cdot x_1 - (0.26 \pm 0.06) \cdot x_2 + (0.29 \pm 0.09) \cdot (x_1^2 - 14/15) \quad (9)$$

For DEE selectivity, obtained data varied in a narrow range, from 2.2 to 3.5 %, with a standard deviation of 0.06 %. The empirical model described in Equation (10) indicated that the linear effect of temperature is the most important, although the WHSV exerted a nonlinear influence on DEE selectivity. Thus, the DEE selectivity decreased with the increase of temperature, showing a point of maximum as a function of WHSV. This behaviour is illustrated in Figure 6(b), where the lines represent the isotherms.

$$S_{DEE} = (2.85 \pm 0.06) - (0.15 \pm 0.05) \cdot x_1 - (0.02 \pm 0.005) \cdot (x_2^2 - 12/15) \quad (10)$$

Using a ZrO₂:ZnO containing MgO-SiO₂ system, similar effects of reaction temperature and WHSV on selectivities of ethanol dehydration products were observed [2]. It can be rationalised that the increase of ethanol molar fraction in the feed can promote DEE formation due to the higher concentration of surface ethoxide species, which can suppress the formation of ethene [2]. However, when the WHSV reached higher values, selectivity to DEE was reduced due to the increase of AcH selectivity, as shown in Figure 6(c).

Figure 6(c) shows the experimental data and calculated values of AcH selectivities as functions of WHSV. AcH selectivities ranged from 14.6 to 57.3 % and with a standard experimental deviation of 4.9 %. AcH selectivities increased with the linear increase of WHSV and decreased with the interaction effect observed between WHSV and temperature, as described in Equation (11).

$$S_{AcH} = (27.47 \pm 1.33) + (4.30 \pm 0.56) \cdot x_2 - (1.65 \pm 0.44) \cdot x_1 \cdot x_2 \quad (11)$$

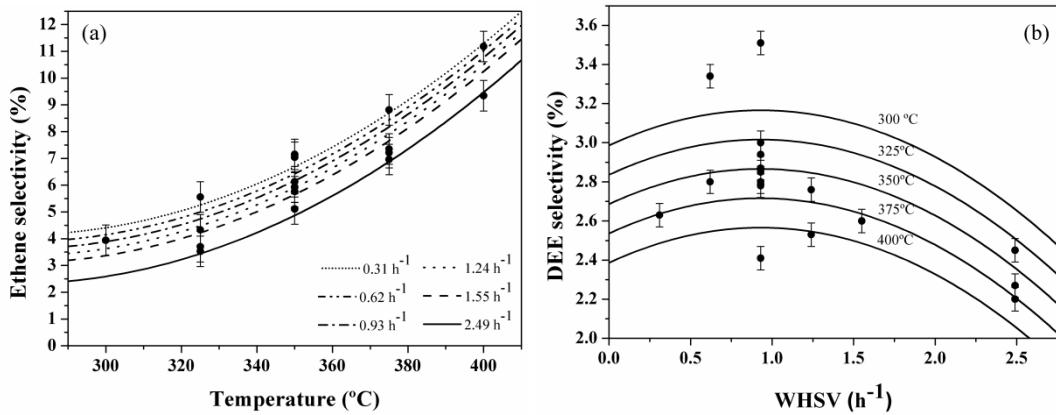
Thus, results suggest that ethanol dehydrogenation can be facilitated at the catalyst surface, since the lower contact times associated with higher WHSV values cause the increase of the AcH selectivities. Besides, results presented in Figure 6(c) also suggest that AcH was the primary product, since selectivities were high even at low contact times. These high selectivity values also suppressed DEE formation, as discussed in the previous paragraph. These results are also in agreement with previously observed empirical data [2]. It is more complicated to understand the interaction effect between WHSV and temperature on AcH selectivity, but it may be associated with the higher AcH to 1,3-BD conversions as temperature increases [7,11].

The effect of reaction variables on 1,3-BD selectivities, however, indicates a clear relationship between AcH and 1,3-BD. 1,3-BD selectivities ranged from 33 to 66 % and the standard experimental deviation was equal to 3.99 %. The empirical model of Equation (12) presented a linear correlation coefficient of 0.88, as illustrated in Figure 6(d). This graph is similar to that obtained for AcH selectivities; however, opposite effects were observed (that is, WHSV affected linearly and negatively 1,3-BD selectivities, whereas the interaction effect was positive). Thus, these results suggest that the AcH condensation can be the slowest reaction step, since selectivities of AcH and 1,3-BD present opposite trends.

$$S_{BD} = (53.92 \pm 0.99) - (3.08 \pm 0.42) \cdot x_2 + (0.91 \pm 0.33) \cdot x_1 \cdot x_2 \quad (12)$$

However, in order to increase 1,3-BD selectivities it is not only necessary to decrease the WHSV, as it results in the increase of the butene selectivities. Butene selectivities ranged from 2.3 to 9.1 % and the standard deviation was equal to 0.38 %. The empirical model described by Equation (13) led to a linear correlation coefficient of 0.94 and is illustrated in Figure 6(e). The WHSV exerts a fundamental role on the evolution of butene selectivities, effecting these values linearly and non-linearly. Temperature, in turn, exerted only a weak linear influence. Therefore, it is possible to identify a point of minimum for butene selectivities at each reaction temperature, as shown in Figure 6(e). It has been reported that butene formation may involve the deoxygenation of butanal, produced from the isomerisation of crotyl alcohol [7], or a butanol dehydration product, which in turn can be produced from the hydrogenation of butanal or the hydrogenation of the vinyl bond of a crotonaldehyde product [15,48]. However, no traces of butanal or butanol were observed in this work, suggesting that butene may result from the 1,3-BD hydrogenation, in line with previous results [2].

$$S_{Butene} = (6.04 \pm 0.17) + (1.05 \pm 0.15) \cdot x_1 - (1.04 \pm 0.18) \cdot x_2 + (0.11 \pm 0.04) \cdot (x_2^2 - 12/15) \quad (13)$$



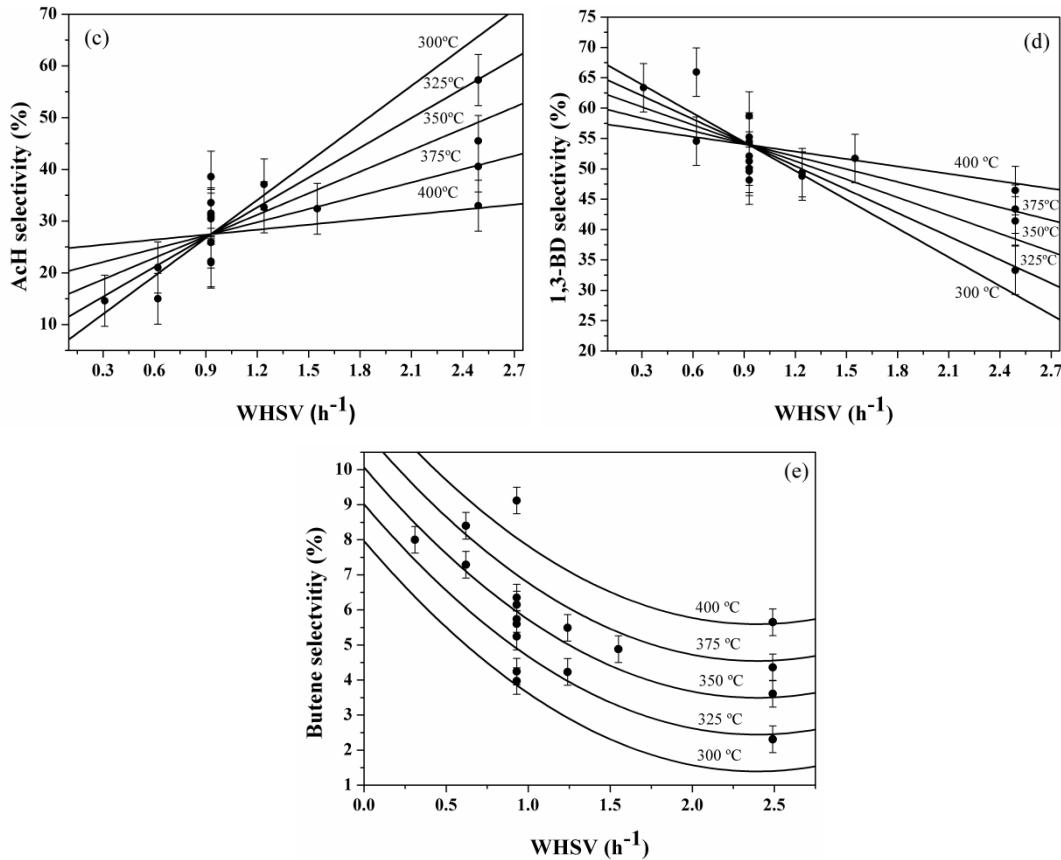


Figure 6 - Experimental values (●) and empirical model selectivities for: (a) ethene ($r = 0.97$); (b) DEE ($r = 0.82$); (c) AcH ($r = 0.89$); (d) 1,3-BD ($r = 0.88$); (e) butene ($r = 0.94$). In (a), lines represent constant WHSV. In (b), (c), (d) and (e), lines represent isotherms.

However, analyses of the reaction variables effects on product molar fractions resulted in much more accurate relationships, with linear correlation coefficients of 0.99, 0.96, 0.95, 0.99 and 0.98 for ethene, DEE, AcH, 1,3-BD and butene, respectively, as shown in Figure 7. This is because selectivities are strongly affected by fluctuations of molar fractions of all reaction products. As a consequence, variances of selectivities are higher than variances of molar fractions, contributing to the lower quality of fittings presented in Figure 6.

The molar fractions of ethanol dehydration products, ethene and DEE, were not significantly affected by the WHSV, with temperature being the only variable responsible for changes in these molar fractions. Figure 7(a-b) shows the experimental and empirical model for ethene and DEE molar fraction, as described in Equations (14) and (15).

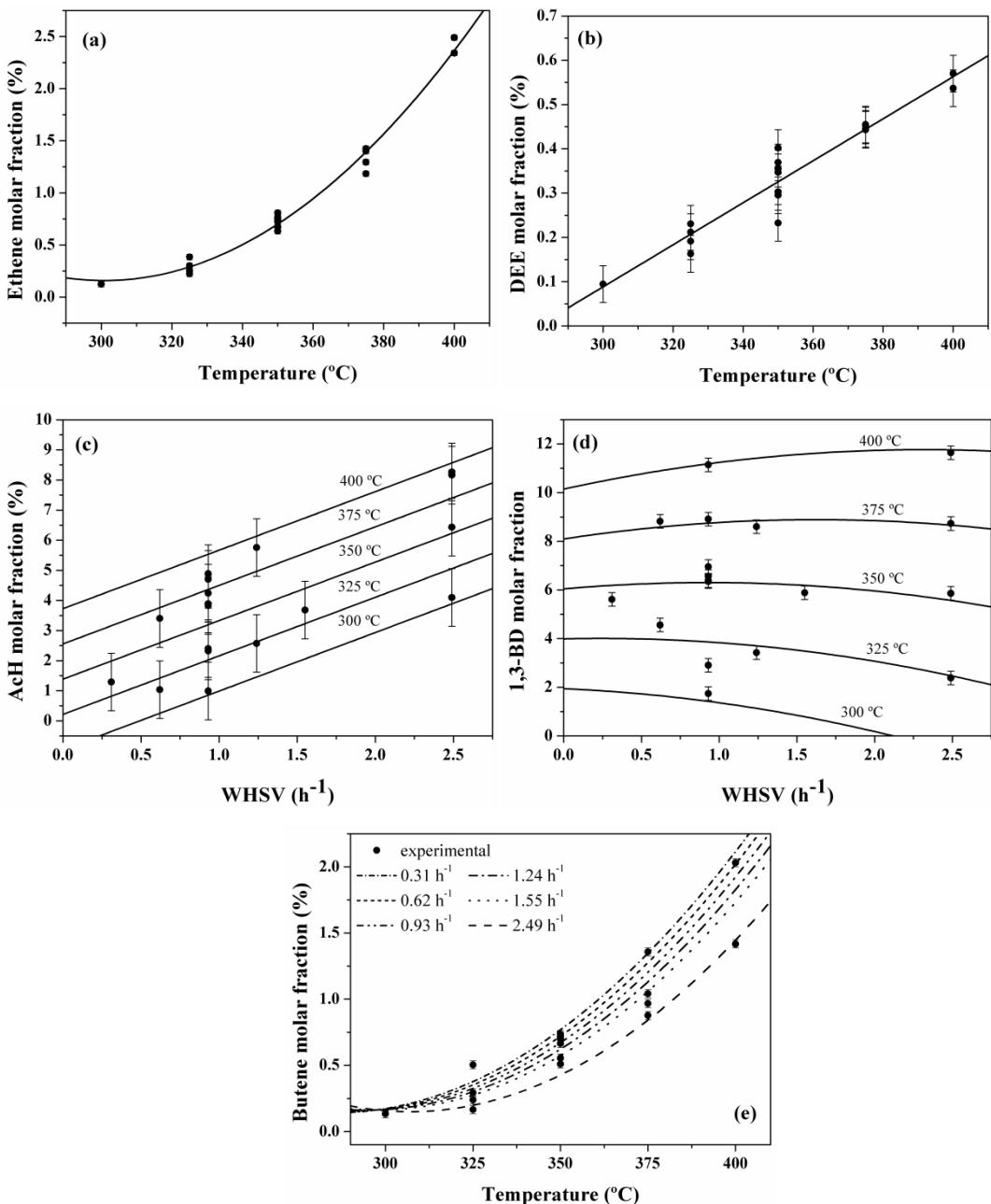


Figure 7 - Experimental values (●) and empirical model molar fraction for: (a) ethene ($r = 0.99$); (b) DEE ($r = 0.96$); (c) AcH ($r = 0.95$); (d) 1,3-BD ($r = 0.99$); (e) butene ($r = 0.98$). Molar fractions do not present their sum next to 100 due to inert gas molar fraction, which was omitted. Experimental molar fractions are also summarised in Table SI1 in the Supporting Information.

$$m_{Ethene} = (0.83 \pm 0.02) + (0.55 \pm 0.02) \cdot x_l + (0.14 \pm 0.02) \cdot (x_l^2 - 14/15) \quad (14)$$

$$m_{DEE} = (0.33 \pm 0.008) + (0.12 \pm 0.008) \cdot x_l \quad (15)$$

$$m_{AcH} = (3.2 \pm 0.19) + (1.17 \pm 0.17) \cdot x_l + (0.60 \pm 0.08) \cdot x_2 \quad (16)$$

$$m_{BD} = (6.28 \pm 0.12) + (2.44 \pm 0.12) \cdot x_1 + (0.13 \pm 0.04) \cdot x_1 \cdot x_2 - (0.029 \pm 0.01) \cdot (x_2^2 - 12/15) \quad (17)$$

$$m_{Butene} = (0.76 \pm 0.03) + (0.44 \pm 0.03) \cdot x_1 - (0.049 \pm 0.01) \cdot x_2 - (0.02 \pm 0.01) \cdot x_1 \cdot x_2 + (0.09 \pm 0.02) \cdot (x_1^2 - 14/15) \quad (18)$$

On the other hand, AcH molar fractions increased as functions of WHSV and temperature, as described by Equation (16) and illustrated in Figure 7(c), highlighting the dehydrogenation capacity of this catalytic system. 1,3-BD molar fractions, in turn, increased with temperature and with the interaction effect between temperature and WHSV, being also affected negatively by the nonlinear WHSV effect. Thus, it is possible to identify a point of maximum in the 1,3-BD molar fraction for each reaction temperature, as shown in Figure 7(d). Finally, butene molar fractions were adjusted by a function containing the linear effects of temperature and WHSV and the quadratic effect of temperature, as shown in Equation (18) and Figure 7(e).

It is important to note that all effects in Equations (8-18) were found to be statistically significant, within 95 % confidence level, when all data were used to estimate the model parameters. Thus, even though some error bars cross more than one statistical model line (see Figure 5, Figure 6 (a,c,d) and Figure 7(c)), all variable effects are meaningful – and valid – as they were determined with the whole experimental set. For instance, in Figure 7(c) at WHSV equal to 2.5 h^{-1} , although values at 400°C are not significantly different from values predicted at 375°C , the significant increase of the AcH molar fraction can be observed as temperature increases from 300°C to 350°C and from 350°C to 400°C . Consequently, it can be concluded that the increase of temperature leads to the significant increase of the AcH molar fraction.

The results obtained with the molar fractions highlight some of the advantages related to the use of experimental designs for identification and quantification of variable effects. All molar fractions of products increased linearly with temperature, which is directly related to the increase of the ethanol conversion. However, it is possible to observe that this effect was more pronounced for the 1,3-BD molar fractions, which presented a linear temperature effect equal to 2.44, as shown in Equation (17), followed by the effects on AcH, ethene, butene and DEE molar fractions. This suggests that reduction of DEE selectivities with temperature may be related to the faster increase of the molar fractions of the remaining products, when compared to the molar fractions of DEE. Additionally, since the WHSV exerted no significant effect on ethene and DEE molar fractions, the WHSV effect on ethene and DEE selectivities can be

understood as a consequence of the observed WHSV effect on 1,3-BD, AcH and butene molar fractions.

Equations (14-18) also indicate that higher WHSV conditions are beneficial for 1,3-BD production, as AcH and 1,3-BD molar fractions are favoured by this variable. This suggests that the catalyst surface should be rich in active sites for ethanol dehydrogenation, since the increase of the ethanol molar fraction in the feed (using higher WHSV) resulted in higher AcH molar fractions. Furthermore, the behaviour of 1,3-BD and AcH molar fractions supports the hypothesis that the AcH condensation step constitutes the slowest reaction step, as also concluded from product selectivity analysis. Therefore, efforts should be driven to describe how AcH condensation sites depend on the reaction conditions and catalyst preparation conditions, in order to further optimise catalyst properties and maximise 1,3-BD production.

At this point, it is important to emphasise that Equations (8-18) can be used to describe catalytic performance within the analyzed experimental ranges and should be used with care for extrapolations. Nevertheless, the proposed models describe catalyst activity very well inside the selected experimental region, unveiling the kinetic rate-limiting step in the commonest range of experimental conditions reported in the literature and fully justifying the utilizations of the proposed statistical approach.

The effects of reaction variables on 1,3-BD yields and productivities were also evaluated. 1,3-BD yields, as defined in Equation (3), ranged from 4.9 to 31.4 % and had standard deviation of 0.73 %. Data could be fitted using a function containing the linear effect of temperature and the linear and quadratic effect of WHSV, as shown in Equation (19) and Figure 8(a). As pointed previously, catalyst performance is usually assessed at low ethanol flow rates, resulting in too low 1,3-BD productivities to be industrially significant [2,12]. The quantification of reaction variables effects on 1,3-BD yield showed, however, that WHSV affected 1,3-BD yields linearly and nonlinearly, being possible to obtain reasonably high 1,3-BD yields under high WHSV conditions, thus increasing the industrial viability of this system.

$$Y_{BD} = (18.45 \pm 0.26) + (6.96 \pm 0.23) \cdot x_1 - (2.99 \pm 0.27) \cdot x_2 + (0.39 \pm 0.06) \cdot (x_2^2 - 12/15) \quad (19)$$

Figure 8(b) shows experimental 1,3-BD productivities and values calculated with an empirical model, as a function of WHSV. Productivities ranged from 0.06 to

0.49 g_{BD}/g_{cat}·h and the standard deviation was equal to 0.006 g_{BD}/g_{cat}·h. 1,3-BD productivities increased as a linear function of temperature and WHSV, being favoured by the interaction effect, as shown in Equation (20) and Figure 8(b).

$$P_{BD} = (0.116 \pm 0.002) + (0.046 \pm 0.002) \cdot x_1 + (0.024 \pm 0.001) \cdot x_2 + (0.016 \pm 0.001) \cdot x_1 \cdot x_2 \quad (20)$$

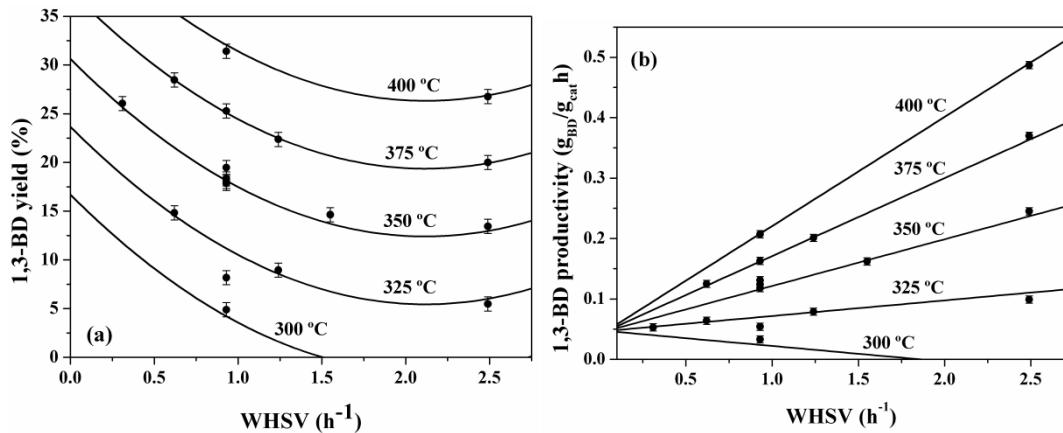


Figure 8 - Experimental and empirical model for: (a) 1,3-BD yield ($r = 0.99$); (b) 1,3-BD productivity ($r = 0.99$).

Finally, as WHSV was modified by varying ethanol flow rate only, it is not possible to distinguish between contact time and ethanol composition effects on the discussed models. Thus, WHSV effects represent the influence of both variables.

4. Conclusions

For the first time the effects of temperature and ethanol WHSV on the performances of ethanol to 1,3-BD reactions were investigated with the aid of a statistical experimental design approach. Catalytic results, ethanol conversions, product selectivities, 1,3-BD yields and productivities were correlated with reaction variables, allowing for identification and quantification of variable effects on the 1,3-BD formation.

The interaction effect between temperature and WHSV was very important for 1,3-BD molar fractions, selectivities and productivities. Thus, evaluation of catalyst performance in terms of the “change-one-factor-at-a-time” method should be thought carefully, so as not to erroneously assign a change in selectivity to one variable and not two acting in a cooperative manner. Moreover, the nonlinear effects of WHSV on 1,3-

BD molar fractions and yields were significant, indicating that high WHSV conditions can clearly benefit the 1,3-BD formation with the investigated catalyst system.

Further, the results indicate the existence of a strong relationship between acetaldehyde and 1,3-BD selectivities, as well as their respective molar fractions, suggesting that conversion of acetaldehyde to 1,3-BD constitutes the rate determining step of the reaction mechanism. As a consequence, efforts should be driven to understand and improve how AcH condensation sites depend on the reaction conditions and catalyst preparation conditions, in order to allow for optimisation of catalyst properties and maximise 1,3-BD production. Finally, given the obtained empirical model responses, the investigation of 1,3-BD production at higher temperatures and WHSV's should be considered.

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Full experimental data and further characterisation is available as supporting information

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