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Liquid phase hydrogenation in a structured multichannel reactor

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ABSTRACT

A compact, structured, multichannel reactor was tested for two exemplary reactions: the selective hydrogenation of aromatic aldehyde and the tandem C–C coupling–hydrogenation. In the case of hydrogenation reaction up to 50% yield with ca. 96% selectivity was attained in a single pass (single channel of 10 cm length of catalytic bed) at a liquid phase residence time of ca. 10 s, proving the effectiveness of the designed reactor. By controlling the point of injection of hydrogen into the reactor an increase in the yield of hydrogenation up to 73% was achieved using two channels in a consecutive mode. The designed compact reactor was also proven to be an excellent tool for kinetic studies: the kinetics of the three-phase hydrogenation was evaluated showing: (i) reaction limitation for applied reaction conditions, (ii) Langmuir–Hinshelwood mechanism of hydrogenation, and finally (iii) the dominating role of adsorption of reactant and absorption of hydrogen in the mechanism of hydrogenation at higher temperatures. Furthermore, the compact reactor was also successfully used for conducting sequential coupling Heck reaction with consecutive hydrogenation of double C–C bond. A stepwise conversion of the substrates to the final product was achieved with ca. 6 min residence time at relatively low temperature and pressure. In conclusion, the developed structured compact reactor was demonstrated as a promising alternative to replace conventional batch reactors and establish continuous synthesis of pharmaceutical intermediates and specialty chemicals.

1. Introduction

Advances in the field of process intensification have led to the development of micro-heat exchanger and compact heat exchanger and reactors. These consist of multichannel structures that may have individual channels with hydraulic diameters in the range of 0.05–5 mm [1–3]. In comparison with the conventional multiphase reactors, these compact structures provide more intensive conditions for promoting catalytic reactions due to the significant reduction of the length scales across which mass and heat transfer should occur. For example, compared with the values reported for the traditional large-scale multiphase packed-bed reactors, more than a 100-fold increase in the gas/liquid mass transfer rate was measured for a microreactor with 0.625 mm reaction channel, which in return significantly accelerated the catalytic reactions [2]. In our previous studies, we demonstrated the use of a compact multifunctional reactor as an effective tool for performing partial oxidation of organic feedstocks. The reactor was integrated with a static mixer and a micro-heat exchanger which enable excellent mixing of reagents prior to entering the catalytic channel, and good heat transfer efficiency for safe running of reactions with significant heat effects. Selective oxidation of aromatic alcohols by molecular oxygen was demonstrated at up to 24 atm undiluted oxygen pressure, which was enabled by (i) highly efficient heat removal in an integral micro-heat exchangers and (ii) the absence of gas-phase volume in which an explosive mixture of hydrocarbon vapour and oxygen may be produced [3].

Continuous operation is another major advantage offered by micro/compact reactors. Adopting this reactor technology in chemistry as a potential solution to circumvent limitations of batch processes is often called ‘flow chemistry’. A prominent benefit of microreactor technology is high surface-to-volume ratio which enables a narrow temperature profile along the reactor and efficient cooling. This feature offers the opportunity to successfully execute highly exothermic/endothermic chemical transformations which are usually inhibited in larger batch reactors. Gross et al. [4] demonstrated the use of a flow reactor in the synthesis of NBI-75043 (an anti-insomnia drug) to promote the key halogen–metal exchange step whose highly exothermic nature limited this total synthesis route in a large-scale batch. Small reaction volume also endows microreaction technology with the inherent safety, which allows us to perform hazardous reactions and reactions using toxic reagents, for instance, ring-expansion reaction involved diazo compound as a reactant [5].

In this study, we expanded the application scope of the compact reactor [3,6] to the three-phase catalytic hydrogenation (hydrogenation of benzaldehyde as a model reaction) and continuous tandem C–C Heck coupling with consecutive hydrogenation

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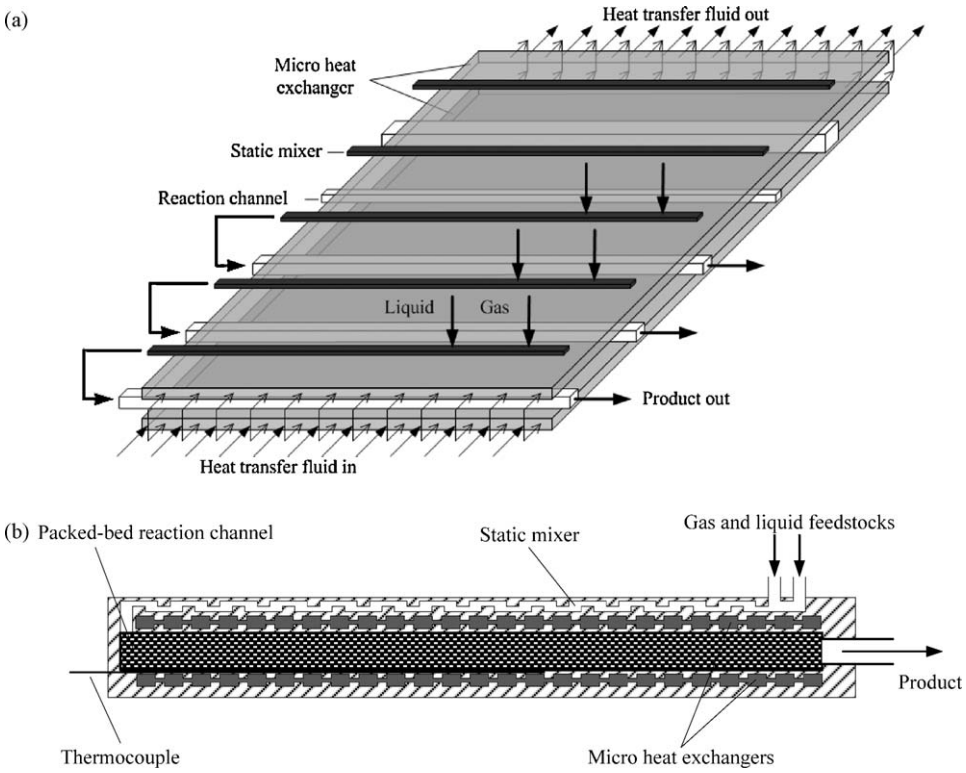


Fig. 1. Schematic diagram of the structured compact reactor–microchannel heat exchanger: (a) channel arrangements and (b) cross-sectional view.

reaction. The effects of the operating conditions and staged injection of hydrogen on the reactor performance were studied in detail for the hydrogenation reaction. The possibility of using compact multichannel reactor for performing continuous sequential synthesis was also examined.

2. Experimental

2.1. Pt/C and Pd/C catalysts and reagents

The catalysts used in this study were 3 wt.% Pt/C (prepared by impregnating the synthetic carbon using H_2PtCl_6 as precursor as described by van Dam and van Bekkum [7] and 5 wt.% Pd/C prepared by using deposition–reduction method using H_2PdCl_4 as precursor. Both catalysts were produced by using microspherical mesoporous carbon supports (MAST Carbon International) with mean particle size ca. 150 μm, manufactured from phenolic resins. All chemicals used in this study were purchased from Sigma–Aldrich with either ReagentPlus® grade (for reactants) or HPLC grade (for solvents) and used without further purification.

2.2. Multifunctional compact reactor and experiment procedure

The schematic diagram of the compact reactor assembly is shown in Fig. 1. Static mixers were embedded prior to reaction channels (on the top of the reaction channels, see Fig. 1(b)) to pre-saturate the liquid phase with gaseous reactant before introducing the mixed stream into the reaction channels. Five packed-bed channels (100 mm length) were imbedded in parallel between the two micro-heat exchangers. The reaction channels have square cross-section with three different sizes: 2 mm × 2 mm (one channel), 3 mm × 3 mm (three channels) and 5 mm × 5 mm (one channel). The two micro-heat exchangers were located underneath and above all the reaction channels (Fig. 1(b), arranged in cross-flow with respect to the reaction channels). Therefore, all reaction channels have an even temperature field during operation. Details about the design concepts and hydrodynamic characteristics of the compact reactor and details of the rig were specified elsewhere [3,6]. A HPLC pump (Kontron Instruments) was used to flow liquid feed through packed-bed reaction channels. A low dead-volume six-way valve (Valco Instruments) equipped with a 250 μL sample loop allows sampling at elevated pressures. Temperature was controlled by using a re-circulating bath (Haake) to circulate heat transfer fluid (glycerol) through the micro-heat exchangers. The operating pressure was controlled by a back pressure regulator (Brooks) and the pressure drop across the reactor was monitored using a differential pressure transducer (Bronkhorst). Catalyst beds were washed with solvents after each run for 30 min under flow conditions.

2.3. Analysis techniques

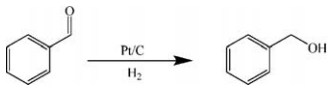
Gas chromatography (GC) analysis was carried out to determine the product distribution and conversion during catalytic experiments. A Varian chromatograph (CP-3800, equipped with FID detector and split injector) and a non-polar capillary column (CP Sil 5 CB) were employed in this study.

MS analysis (microTOF, Bruker) and 1H NMR analysis (NMR) spectrometer (300 MHz, Bruker) were employed for the determination of the chemical structures of by-products.

3. Results and discussion

3.1. Performance of the compact flow reactor

The hydrogenation of benzaldehyde to benzyl alcohol with molecular hydrogen (Scheme 1) was first investigated in the



Scheme 1. Model reaction of hydrogenation of benzaldehyde.

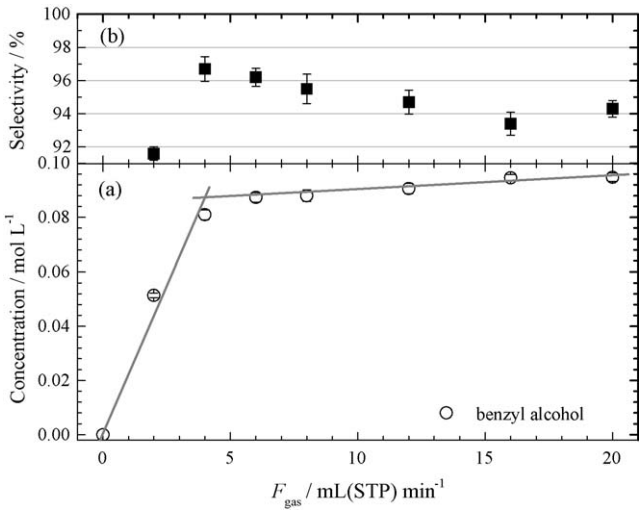


Fig. 2. Product yield and the rate of reaction (a), and product selectivity (b) as a function of hydrogen flow rate. Reaction conditions: $T = 317$ K; $P = 8$ barg; $F_{liquid} = 1$ mL min^{-1}; $C_{substrate,\ 0} = 0.2$ mol L^{-1}.

compact flow reactor. For the substrate concentration used in this study (0.2 mol L^{-1}), 100% conversion of the substrate was achieved in a single pass through a packed 10 cm × 3 mm × 3 mm channel under mild operation conditions, i.e. 320 K, 8 barg, and 10 s liquid phase residence time, and using catalyst mass of 0.45 g. In order to study the effectiveness of the reactor and the catalyst, a catalyst bed diluted with glass beads was employed in the consecutive experiments.

Fig. 2 shows the product yield, reaction rate (steady-state turnover frequency, TOF), and selectivity in a single pass channel as a function of hydrogen flow rate at 317 K and 8 barg. At low gas flow rates, the outlet concentration of benzyl alcohol depended linearly on the hydrogen flow rate. Product yield correlated well with the maximum stoichiometric yield based on the hydrogen amount supplied to the reactor, similarly as it was observed for the oxidation reaction [3,6]. This indicated that for the low range of the hydrogen flow rate (up to 6 mL(STP) min^{-1}), the reaction was limited by the supply of hydrogen. Product yield was found almost independent on the gas flow rate for higher gas flow rates (>6 mL(STP) min^{-1}).

The increase in the gas flow rate decreases the external liquid saturation (β_L, $\beta_L = \dot{V}_L/\dot{V}_L + \dot{V}_G$, where $\dot{V}_L$ and $\dot{V}_G$ are volumetric flow rates of liquid and gas phases in the reactor), and increases the superficial velocity of the liquid phase [8]. The independency of the exit concentration of hydrogenation product (benzyl alcohol) on the flow rate of the gas phase excludes the possibility of influence of the external mass transfer on the overall reactor performance. Therefore, hydrogenation in a compact reactor could be limited either by the chemical reaction itself, or by internal diffusion. The calculations of Weisz–Prater criterion (CWP) [9] excluded the possibility of the internal diffusion to be a rate limiting step, since the CWP was much lower than 1. Thus, it could be confirmed unequivocally, that the hydrogenation reaction in a compact reactor was limited by the catalytic reaction itself and external mass transfer.

In order to highlight the effectiveness of the developed multiphase compact flow reactor, a set of activity data from two batch experiments [10,11] are given in Table 1 for comparison. The observed steady-state turnover frequencies (TOF) were much higher in the case of the flow process in comparison with the earlier studied batch reactors [10,11]. However, a lower selectivity to the desired product benzyl alcohol was found for the flow process due to the acid-induced acetal formation, which can be eliminated by catalysts washing. Although the data presented in Table 1 for the 'Bath' compact reactor was attained with only a 46% conversion, the full conversion of the substrate could be achieved with the undiluted catalyst bed under the same steady-state operation conditions, i.e. single reaction channel, $T = 317$ K, $P = 8$ barg, $F_{liquid} = 1$ mL min^{-1}, and $F_{gas} > 10$ mL(STP) min^{-1}.

3.2. Effect of the operating pressure

The product yield was found to be strongly dependent on the operating pressure, which is illustrated in Fig. 3. An increase in the yield of benzyl alcohol was observed with the increase in the pressure of hydrogen over the pressure range used in this study (1–8 barg). The increase in the partial pressure of hydrogen (here equal to total pressure) increases the solubility of hydrogen in propan-2-ol [12] and therefore increases the rate of catalytic reaction. Based on the data provided in reference [12], a ca. 4.8-fold increase in

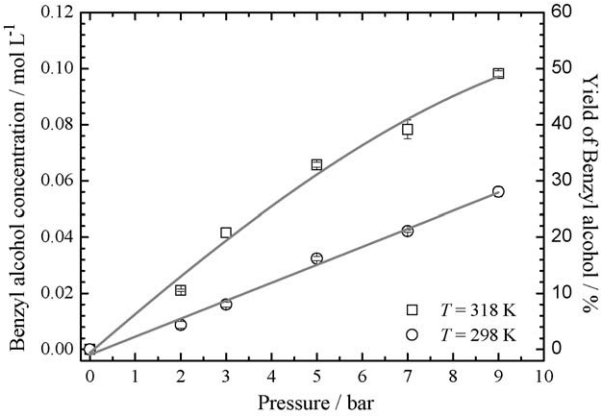


Fig. 3. Effect of the absolute operating pressure on the production of benzyl alcohol. Reaction conditions: $F_{liquid} = 1$ mL min^{-1}; $F_{gas} = 16$ mL(STP) min^{-1}; $C_{substrate,\ 0} = 0.2$ mol L^{-1}.

Table 1
Activities for different reactors for selective hydrogenation of benzaldehyde to benzyl alcohol over supported metal catalysts.

Reactors	Catalyst	Conditions		TOF × 10^{-3} s^{-1}	Selectivity [%]
		T [K]	P [barg]		
'Bath' structured compact reactor[a]	Pt/C catalyst[a]	317	8	95.6	96[a]
Batch reactor 1 [10]	Pt/Al_2O_3[b]	308	1	3.58	100[b]
Batch reactor 2 [11]	Pt/Cc	323	1	24.4	100[c]
	Pd/C[c]			66.5	

[a] This study, 3 wt.% Pt, continuous three-phase catalysis, 46% conversion.
[b] Low-temperature reduced alumina-supported platinum catalyst, 1 wt.% Pt, conversion is not available.
[c] Commercial 5 wt.% Pt/C, alumoborosilicate glass fibers supported palladium catalyst (0.2 wt.% Pd), 100% conversion.

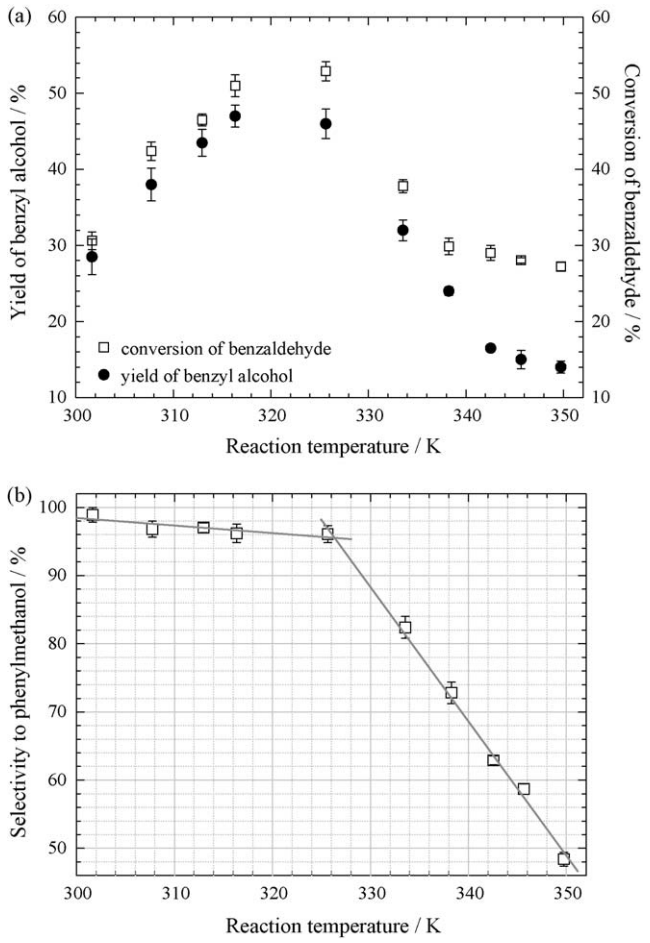


Fig. 4. (a) Yield of benzyl alcohol and conversion of benzaldehyde in the compact reactor as a function of the reaction temperature. (b) Selectivity to benzyl alcohol as a function of the reaction temperature. Reaction conditions: P = 8 barg; F_{liquid} = 1 mL min^{-1}; F_{gas} = 16 mL(STP) min^{-1}; $C_{substrate,\ 0}$ = 0.2 mol L^{-1}.

hydrogen solubility in propan-2-ol was estimated when pressure changes from 1 barg to 8 barg. This increment shows a satisfactory agreement with the experimental results, i.e. the measured product yield at 8 barg is ca. 4.3 times higher than that at 1 barg.

The lack of the influence of the gas flow rate on the overall reactor performance (see analysis in Section 3.1), i.e. the lack of the influence of hydrogen mass transfer on global kinetics, proves the reaction limitation of the process. It can also be seen by analysing the data shown in Fig. 3. For the higher reaction temperature (318 K) the dependency of the exit benzyl alcohol concentration on the hydrogen pressure bends slightly, indicating Langmuir–Hinshelwood-type of hydrogenation kinetics.

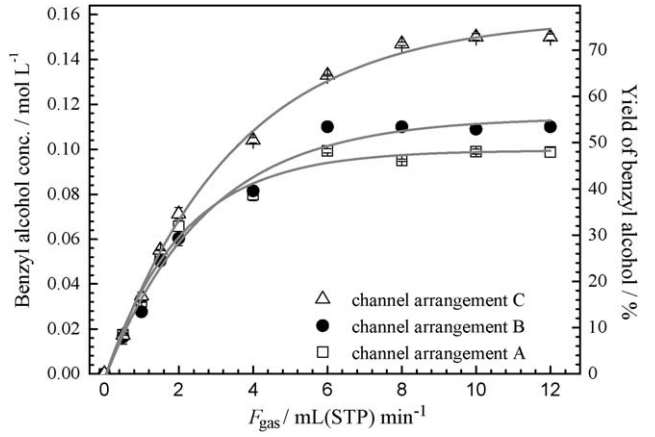


Fig. 6. Benzyl alcohol yield against gas flow rate for different hydrogen injection manners. Reaction conditions: P = 8 barg; T = 318 K; F_{liquid} = 1 mL min^{-1}; $C_{substrate,\ 0}$ (by GC) = 0.2 mol L^{-1}.

3.3. The effect of reaction temperature

The temperature dependence of the conversion of benzaldehyde and yield of phenyl alcohol is shown in Fig. 4(a). Isothermal conditions were kept at all conversion levels, which was proven by temperature measurements inside the reactor. Hydrogenation of benzaldehyde is a mildly exothermic reaction with standard reaction enthalpy of −67 kJ mol^{-1} (thermodynamic properties of compounds involved in this reaction were obtained from [13,14]). However, very effective heat transfer characteristics of the compact reactor secured isothermal mode of operation.

The bell-shape dependencies of the conversion and yield with temperature may result from the following factors: (i) adsorption equilibria, (ii) change of reaction rate constant with temperature (Arrhenius' law), and (iii) change of hydrogen solubility in propan-2-ol with temperature (Henry's law). Adsorption of benzaldehyde and hydrogen on active sites of a catalyst decreases with the increase in temperature, thus decreasing the rate of reaction. Similar, the negative effect of the increase in temperature can be observed in the case of H_2 solubility in propan-2-ol [12], decreasing the concentration of hydrogen at the surface of a catalyst and therefore decreasing the rate of hydrogenation. The only positive effect of temperature on the reaction rate (dominating in the temperature region 303 K < 318 K) is connected with the Arrhenius law. Thus, it follows that for the temperatures higher than ca. 325 K the negative effects of absorption and adsorption prevailed in the overall hydrogenation mechanism.

Additionally, some discrepancies between the changes in conversion and yield with temperature may be attributed to the formation of higher molecular weight products, as proven by mass

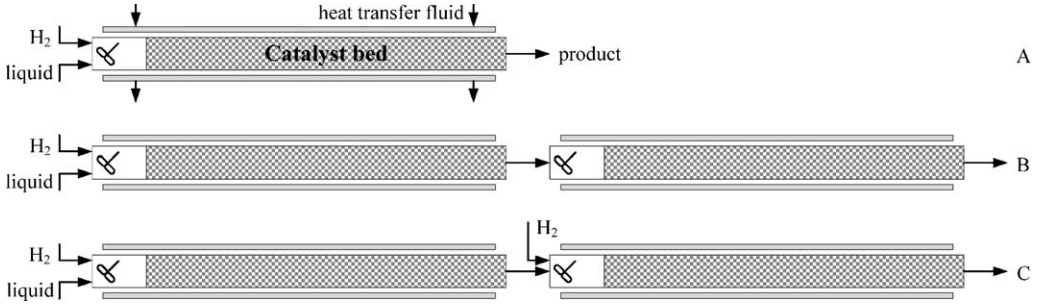
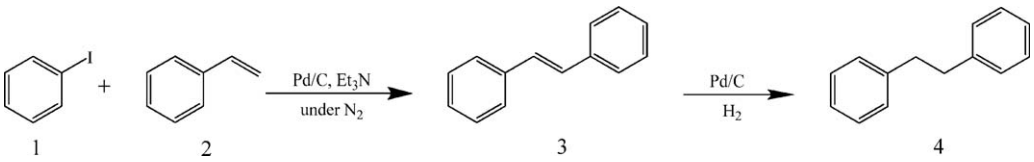


Fig. 5. Channel arrangements for split hydrogen injection experiments. (a) Single channel, injection of hydrogen at length L = 0 cm, (b) double channel, injection of hydrogen at length L = 0 cm, and (c) double channel, injection of hydrogen at lengths L = 0 cm and L = 10 cm.



Scheme 2. Consecutive synthesis of 1,2-diphenylethane (4) involving a Heck C–C coupling followed by the hydrogenation.

Table 2
Results of the continuous sequential reaction in the compact reactor.

Pressure [barg]	PhI conversion [%]	Selectivity [%]		
		3 (intermediate)	4 (product)	Benzene (by-product)
4	100	0	83	13
8	100	0	83	12

Reaction conditions: C_{PhI} = 0.4 mol L^{-1}; T = 393 K; F_{liquid} = 0.25 mL min^{-1}; F_{gas} = 8 mL min^{-1}; ethanol as solvent.

spectroscopy (e.g. benzaldehyde dipropyl acetal). High temperatures were found to be favorable for the side reaction, which was also confirmed by GC analysis. The increase in the by-products formation at high temperatures, furthermore, caused deterioration of selectivity to the main product, which is shown in Fig. 4(b). The selectivity remains above 96% at low temperatures ($T < 330$ K), thereafter, a steep decrease in selectivity was measured.

3.4. Possibility of increasing the product yield by staged injection of hydrogen

The possibility of using staged injection of hydrogen to increase the reactor efficiency was also investigated. The reaction channel arrangements, i.e. different gas injection options, are illustrated in Fig. 5. For all configurations, the total amount of hydrogen supplied relative to the catalyst amount used remains the same. Compared with the arrangement A, double amount of hydrogen was used due to the 2-fold increase of the catalyst amount in the arrangements B and C. Therefore, the two-phase flow pattern along the channels may change, as well as the residence time of the liquid phase in the reactor, and both may affect conversion. Exemplary, a ca. 12% increase in the product yield (up to 55%) was measured with the arrangement B in comparison with the arrangement A (Fig. 6). For the arrangement C, an identical total amount of hydrogen as in the case B was split injected at the entrance of the two consecutive channels. As anticipated, the maximum product yield was increased by 52% (up to 73%) due to the variation in the residence time. In all three cases, selectivity to benzyl alcohol remained the same ca. 96%.

The introduction of split injection (case C) in comparison to feeding the entire amount of gaseous phase at the inlet to the reactor (case B) has a strong influence on the liquid saturation (β_L,) in the reactor and therefore on the residence time of the liquid phase in the reactor. For the case A, the liquid saturation (based on the conversion data shown in Fig. 6 and selected flowrates F_{liquid} = 1.0 mL min^{-1}, F_{gas} = 10 mL(STP) min^{-1}) changed from 0.46 to 0.52 for the inlet and outlet from the reactor, respectively. For the case B the corresponding values were 0.30 and 0.36. The decrease in the liquid saturation equates to the decrease in the residence time of the liquid phase in the reactor; the resulting final conversion was only 12% higher in-spite of doubling the reactor length. For the case C the corresponding values were 0.46 and 0.39. The higher and more equal contact time in both channels resulted in much higher conversions. These results confirm the important role of the two-phase flow hydrodynamics through the packed bed in determining the final conversion.

With regard to the simple model reaction investigated in this work, the split gas distribution along the length of reaction channel resulted in a higher yield of the desired product. However, the concept of dosing the required amount of gaseous reactant at different positions along the catalyst bed (at different levels of conversion) [3] might be a flexible method for performing more complex reactions continuously, such as consecutive oxidation/hydrogenation [3] and tandem reactions.

3.5. Possibility of performing continuous tandem reaction in a compact flow reactor

Based on the effectiveness of the staged gas injection experiments, the possibility of extending the functionality of this flow reactor system to perform continuous sequential reactions was also studied. Palladium catalysed C–C coupling and the subsequent hydrogenation of the resulting alkene were used as model reactions to demonstrate this concept (see Scheme 2). In the tandem reaction experiments, four channels were used in series. The first two channels were used for the Heck reaction and the system was pressurized using N_2 to protect the Heck coupling from hydrogen. Hydrogen was introduced into the subsequent two sections to reduce the intermediate 3 to the final product 4.

Results of this investigation were proven to be satisfactory as shown in Table 2. Complete conversion of the substrate 1 was achieved and an adequate product yield (with ca. 80% selectivity to the final product) was measured. This result suggested that stepwise conversion of the substrates/intermediate occurred in the continuous process. In comparison with a conventional batch reactor [15], the flow process in the compact reactor was more intensified: (i) reaction time was significantly reduced (ca. 7 min versus several hours), (ii) no additional ligands were used and (iii) the reaction was run at a lower operating pressure (8 barg versus 20 barg) and temperature (393 K versus 413 K).

4. Conclusions

The effectiveness of a compact multichannel reactor for performing continuous hydrogenation of benzaldehyde was investigated. It was shown that, in comparison with conventional batch reactors [10,11], the compact flow reactor system is a more intensified tool for promoting the model hydrogenation reaction (TOF = 95.6×10^{-3} s^{-1}, at 317 K, and at ca. 50% conversion ca. 96% selectivity for the single pass through the reactor). The observed excellent performance of the reactor is largely due to the integration of heat transfer, mixing and reaction functionalities into a single reactor, as well as the choice of the reaction channel sizes. Staged injection of hydrogen was shown to be beneficial for this model reaction due to a better control over the residence time of liquid phase in reaction channels.

A stepwise conversion of the substrates to a final product was also achieved in the compact multichannel reactor by sequential coupling Heck reaction and hydrogenation. Compared with a batch reactor [15], this flow process in the compact reactor features some advantages, such as: (i) being more intensified (much shorter residence time) due to high catalyst/substrates ratio, and (ii) being operated under more benign conditions, i.e. lower temperature and pressure.

In conclusion, it was shown that the developed structured multichannel reactor shows considerable promise for the catalytic hydrogenation of organic feedstocks as well as for continuous multi-step organic syntheses. Thus, together with the earlier results on catalytic oxidation [3,6], it shows unequivocally the great potential of structured compact reactors for transforming conventional batch catalytic processes into continuous process.

Acknowledgements

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References

[1] W. Ehrfeld, V. Hessel, H. Löwe, Microreactors, Wiley–VCH, Weinheim, 2000.
[2] M.W. Losey, M.A. Schmidt, K.F. Jensen, Ind. Eng. Chem. Res. 40 (2001) 2555.
[3] D.V. Bavykin, A.A. Lapkin, S.T. Kolaczkowski, P.K. Plucinski, Appl. Catal. A: Gen. 288 (2005) 175.
[4] T.D. Gross, S. Chou, D. Bonneville, R.S. Gross, P. Wang, O. Campopiano, M.A. Ouellette, S.E. Zook, J.P. Reddy, W.J. Moree, F. Jovic, S. Chopade, Org. Process. Res. Dev. 12 (2008) 929.
[5] X. Zhang, S. Stefanick, F.J. Villani, Org. Process. Res. Dev. 8 (2004) 455.
[6] P.K. Plucinski, D.V. Bavykin, S.T. Kolaczkowski, A.A. Lapkin, Ind. Eng. Chem. Res. 44 (2005) 9683.
[7] H.E. van Dam, H. van Bekkum, J. Catal. 131 (1991) 335.
[8] M.P. Dudukovic, F. Larachi, P.L. Mills, Catal. Rev. 44 (2002) 123.
[9] P.B. Weisz, C.D. Prater, Adv. Catal. 6 (1954) 143.
[10] M. Arai, A. Obata, Y. Nishiyama, React. Kinet. Catal. Lett. 61 (1997) 275.
[11] L. Kiwi-Minsker, I. Yuranov, V. Holler, A. Renken, Chem. Eng. Sci. 54 (1999) 4785.
[12] Hydrogen and Deuterium, Pergamon, Oxford, 1981.
[13] S.P. Verevkin, T.V. Vasiltsova, J. Chem. Eng. Data 49 (2004) 1717.
[14] A. Vatani, M. Mehrpooya, F. Gharagheizi, J. Mol. Sci. Int. 8 (2007) 407.
[15] M. Gruber, S. Chouzier, K. Koehler, L. Djakovitch, Appl. Catal. A: Gen. 265 (2004) 161.