Development of Membrane Reactors for Heterogeneously Catalysed Aerobic Oxidation of Alcohols

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Declaration of Authorship

I, Gaowei Wu, declare that the work presented in this thesis is my own. Where information has been derived from other sources, I confirm that this has been indicated in the thesis.
Abstract

Heterogeneously catalysed aerobic oxidation of alcohols has great potential in chemical synthesis, but its wide application is still limited by safety issues with the combination of gaseous oxygen and flammable organics. The aim of this thesis is to develop Teflon AF-2400 membrane reactors for the intrinsically safe use of oxygen in oxidation of alcohols.

Initially, oxidation of benzyl alcohol and cinnamyl alcohol on Au-Pd/TiO$_2$ catalyst was studied in a trickle bed microreactor. The catalyst deactivation in cinnamyl alcohol oxidation, rather than benzyl alcohol oxidation, was attributed to Pd leaching and a complex role of oxygen. Then, a Teflon AF-2400 packed tube-in-tube membrane microreactor was investigated for benzyl alcohol oxidation, which allowed continuous oxygen supply during the reaction and presented higher conversion and selectivity as compared to a reactor with oxygen pre-saturated feed. A novel approach using the tube-in-tube membrane contactor was demonstrated for measuring gas solubility in liquids.

To simplify the reactor scale-up, a Teflon AF-2400 flat membrane microreactor was developed for benzyl alcohol oxidation, and the mass transfer and reaction in the reactor were experimentally and theoretically investigated with different catalysts. The oxygen transverse mass transfer in the catalyst bed, rather than oxygen permeation through membrane or oxygen internal/external transfer in the catalyst particles, was indicated to be the controlling process. An effectiveness factor analysis akin to internal/external mass transfer and reaction in a catalytic particle was provided to guide the catalyst choice and the membrane reactor design.
For direct usage of small catalyst particles in continuous flow reactors, a stirred membrane reactor with a sintered metal filter and an external membrane contactor was experimentally demonstrated and mathematically simulated for benzyl alcohol oxidation. The reactant conversion and the catalyst utilization were indicated to be affected by various operation parameters, which were correlated to guide the reactor design and operation.
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contactor; $X_{\text{exp}}$, the conversion from the experiment; $c_{O_2,\text{sat}}$, oxygen saturation concentration at given oxygen pressure; $c_{O_2,\text{in}}$ and $c_{O_2,\text{out}}$, oxygen concentration at the inlet and outlet of the reactor; $X_{\text{sim}}$, the conversion from the simulation................186
Nomenclature

$A_b$  active specific surface area in the 1D reactive catalyst bed [1/m]

$Bi$  Biot number [-]

$CCT$  catalyst contact time [$g_{\text{cat}} \cdot \text{min}/g_{\text{alcohol}}$]

$CR$  concentration ratio [-]

$c$  concentration [mol/m$^3$]

$d$  depth [mm]

$D$  diffusion/dispersion coefficient [m$^2$/s]

$Da$  Damköhler number [-]

$F$  molar flow rate of alcohol [mol/min]

$H$  dimensionless Henry solubility [-]

$h$  mass transfer coefficient [m/s]

$J$  surface flux [mol/(m$^2$·s)]

$k$  reaction rate coefficient [$m^{3(\alpha+\beta-1)}/(\text{mol}^{(\alpha+\beta-1)} \cdot s)$]

$l$  length [mm]

$M$  molecular weight [g/mol]

$m_{\text{cat}}$  mass of catalyst [mg]

$n_{\text{Ru}}$  the moles of Ru contained in the catalyst packed [mol]

$OCR$  oxygen consumption rate [mL/min]

$P$  pressure [bar]
Nomenclature

\( Pe \)  \( \text{Peclet number [-]} \)

\( r \)  \( \text{radius [\( \mu \text{m} \)]} \)

\( r_{\text{dim}} \)  \( \text{spatial radial coordinate in the pellet [\( \mu \text{m} \)]} \)

\( \bar{R} \)  \( \text{ideal gas constant [m}^3\text{-bar/(K-mol)}] \)

\( R \)  \( \text{reaction rate [mol/(m}^3\text{ bed} \cdot \text{s}]} \)

\( \dot{R} \)  \( \text{recycle ratio [-]} \)

\( Re \)  \( \text{Reynolds number [-]} \)

\( S_B \)  \( \text{selectivity to benzaldehyde [%]} \)

\( Sc \)  \( \text{Schmidt number [-]} \)

\( Sh \)  \( \text{Sherwood number [-]} \)

\( T \)  \( \text{temperature [K]} \)

\( TOF \)  \( \text{turnover frequency [h}^{-1}] \)

\( u_x \)  \( \text{longitudinal liquid superficial velocity [cm/s]} \)

\( U \)  \( \text{catalyst utilization factor [-]} \)

\( \tilde{V} \)  \( \text{molar volume [mL/mol]} \)

\( w \)  \( \text{width [mm]} \)

\( X \)  \( \text{conversion of benzyl alcohol [%]} \)

\( x \)  \( \text{x-coordinate [mm]} \)

\( y \)  \( \text{y-coordinate [mm]} \)
Greek Symbols

$\alpha$  reaction order in benzyl alcohol [-]

$\beta$  reaction order in oxygen [-]

$\varepsilon$  porosity or void fraction [-]

$\eta$  effectiveness factor [-]

$\mu$  viscosity [Pa·s]

$\rho$  density [g/mL]

$\tau$  tortuosity [-] or residence time [h]

$v$  volumetric flow rate of benzyl alcohol [$\mu$L/min]

$\phi$  Thiele modulus [-]

$\chi$  association factor [-]

Subscripts

$A$  axial direction

$atm$  atmospheric pressure

$ave$  average

$B$  benzaldehyde

$BnOH$  benzyl alcohol

$b$  catalyst bed

$cat$  catalyst
Nomenclature

\( CD \) cinnamaldehyde

\( D \) disproportionation reaction

\( eff \) effective diffusion coefficient in the catalyst

\( g \) gas phase

\( i \) benzyl alcohol, oxygen

\( in \) inlet of the reactor

\( IN \) inlet of the reactor system

\( m \) membrane

\( O \) oxidation reaction

\( O_2 \) oxygen

\( out \) outlet of the reactor

\( OUT \) outlet of the reactor system

\( p \) catalyst particle

\( PP \) 3-phenyl-1-propanol

\( ps \) catalyst particle surface

\( R \) recycle flow

\( sat \) saturation

\( T \) transverse direction

\( TR \) trans-\( \beta \)-methylstyrene

\( total \) overall reaction
CHAPTER 1. Introduction

This chapter introduces the development of aerobic oxidation of alcohols in fine chemicals and pharmaceuticals industries, and presents the motivation and objectives of this thesis. The structure of the thesis is also outlined.
1.1 Motivation and Objectives

Oxidation of alcohols is one of the most fundamental organic transformations and has great importance in chemical synthesis.\textsuperscript{1-5} The desired products, such as aldehyde, ketone, and acid, are valuable intermediates for fine chemicals and pharmaceuticals industries. Traditionally, this type of reactions involves stoichiometric inorganic oxidants (e.g. chromium (VI), permanganate or nitric acid), which are often not chemoselective and could raise seriously environmental problems.\textsuperscript{5,6}

Molecular oxygen can be considered as the ideal oxidant for oxidation reactions, since it is readily abundant and commonly forms water as the only by-product.\textsuperscript{7,8} This is a big advantage for the production of fine chemicals and pharmaceuticals industries, since the impurities allowed in the products are tightly regulated. Currently, there are 109 industrial oxidation processes with a capacity of > 1000 tonnes/annum listed in \textit{Ullmann's Encyclopedia of Industrial Chemistry}, and 67 out of these 109 processes are using oxygen as an oxidant.\textsuperscript{7,9}

However, a further examination would find that the majority of these processes using oxygen are performed either in gas phase or in liquid phase with aqueous solutions. The lack of aerobic oxidation methodologies for organic solvents is indicated to be more often related to the safety concerns with the combination of gaseous oxygen with flammable organics under high temperature and pressure.\textsuperscript{7,8,10} This becomes more remarkable when it comes to the highly exothermic oxidation of alcohols (e.g. -187 kJ/mol for oxidation of benzyl alcohol to benzaldehyde by oxygen\textsuperscript{11}). To address safety concerns, diluted oxygen (oxygen concentration below 10\%) is most commonly used to ensure the operational conditions outside the
flammability limits. However, this could result to sacrificed process efficiency, since higher oxygen concentration can often enhance reaction rates.\textsuperscript{9} Temperature in the reactor is also preferably below the autoignition temperature. Given that the total mass of explosive mixture determines the power of an explosion, minimization of the reactor volume seems to be advantageous for safe operation.\textsuperscript{12} Better mass and heat transfer is also favoured to avoid local hot spot formation.

For fine chemicals and pharmaceuticals industries, multiple operations with limited scales also hinder the development of aerobic oxidation reactions in such sectors.\textsuperscript{7, 8} As compared to dedicated and highly integrated facilities for bulk chemical processes (due to the economy of scale), multi-purpose batch reactors are typically used for pharmaceutical and fine chemicals to achieve process flexibility at the expense of efficiency. Due to the considerable headspace in batch reactors, aerobic oxidations cannot be simply implemented and scaled in such reactors, making the reactions fall into a unique category.

Homogeneous catalysts are commonly utilised in oxidation of alcohols, highlighting excellent selectivity and no internal diffusion resistance.\textsuperscript{5} However, an additional separation step is required to remove the homogeneous catalysts from the reaction media, and the recovery and reuse of the homogeneous catalysts are still difficult and expensive. Comparatively, heterogeneous catalysts seem to be optimal due to the ease of processing.\textsuperscript{6} Nevertheless, more efficient oxidation chemistry is still required to realize different oxidative reactions chemoselectively under mild conditions.\textsuperscript{7} It has been identified as a key bottleneck in an analysis of reactions used for the preparation of 128 drug candidate molecules within GlaxoSmithKline, AstraZeneca and Pfizer.\textsuperscript{13} Specific design of reactors would boost the screening and
identification of suitable catalysts for aerobic oxidation reactions.

During the past decade, continuous processing techniques provide an excellent chance to address the above issues.\textsuperscript{14-18} This thesis is aimed to develop new and improved continuous flow membrane reactors and generate fundamental process understanding, in order to efficiently and safely facilitate the application of heterogeneously catalysed aerobic oxidation of alcohols. To achieve this objective, catalyst activity and stability in oxidation of alcohols were studied, and data on oxygen solubility in organic solvents were measured. Various types of membrane reactors (e.g. packed bed membrane reactor, stirred membrane reactor) were also experimentally and theoretically investigated for aerobic oxidation of alcohols.

1.2 Outline of Thesis

In Chapter 1, the background and challenges faced by aerobic oxidation of alcohols are introduced. The motivation and objectives are given.

In Chapter 2, the development in oxidation of benzyl alcohol and cinnamyl alcohol is reviewed, followed by an introduction of microreactors, especially their application in three-phase reactions. The progress of membrane reactors/microreactors in three-phase reactions is also presented, highlighting the potential of Teflon AF-2400 membrane reactors.

In Chapter 3, the stability of Au-Pd/TiO\textsubscript{2} catalyst was initially examined in a packed bed capillary microreactor for oxidation of benzyl alcohol and cinnamyl alcohol. Different catalyst stabilities were observed for these two reactions, and the reason was identified by changing the reaction conditions in the microreactor.

In Chapter 4, a Teflon AF-2400 tube-in-tube membrane microreactor was
investigated for heterogeneous oxidation of benzyl alcohol with pure oxygen. The inner Teflon AF-2400 tube was packed with Au-Pd/TiO$_2$ catalyst particles, as an attempt to continuously supply oxygen during the reaction. Oxygen permeation through the Teflon AF-2400 membrane was experimentally studied and the performance of different reactors was compared. The effect of oxygen pressure, catalyst contact time, and catalyst dilution was also studied in the packed bed membrane microreactor.

In Chapter 5, a novel approach using a Teflon AF-2400 tube-in-tube membrane contactor was developed for the measurement of gas solubility in organic solvents. The approach was validated by comparing the obtained oxygen solubility in toluene with literature data. Then, oxygen solubility in benzyl alcohol was measured at pressures up to 10 bar and temperatures 298 - 393 K. An empirical correlation of the Henry’s law constants was also determined as a function of temperature.

In Chapter 6, a flat membrane microreactor was developed with a Teflon AF-2400 film, in order to provide wider operating pressure range and facile reactor scale-up. The working pressure of the mesh supported Teflon AF-2400 flat membrane was experimentally studied, together with the organic pervaporation through the membrane. The various mass transfer processes in the membrane reactor were investigated by changing the oxygen pressure, the membrane thickness, the liquid flow rate and the catalyst particle size. An effectiveness factor analysis was also performed to assess the relative importance of these various mass transfer processes on the reactor performance. At the end, scale-up of the flat membrane microreactor was demonstrated by increasing the width of the catalyst bed channel.

In Chapter 7, an experimental and modelling study of the flat membrane
microreactor was further carried out with Ru/Al$_2$O$_3$ catalyst, in an attempt to better understand the mass transfer and catalytic reaction in the reactor and provide guidance on its improvement. A 2D reactor model was developed and validated with the experimental results of benzyl alcohol oxidation under different oxygen pressures and liquid flow rates. Then, the effect of various parameters (membrane thickness, liquid channel depth and reaction rate coefficient) was theoretically studied in the model. The controlling process was identified and the guidance on the reactor improvement was provided.

In Chapter 8, a continuous stirred membrane reactor system was developed and evaluated for the direct usage of small catalyst particles in continuous flow aerobic oxidation of alcohols. A sintered metal disk was used as a filter to retain the Ru/Al$_2$O$_3$ within the reactor, and a Teflon AF-2400 tube-in-tube membrane contactor was employed externally to the stirred reactor within a recycle loop, to allow more flexibility in adjusting the oxygen supply through the membrane. A mathematical simulation was also performed to better understand the effect of the various reactor operating parameters on the catalyst utilization and the reactant conversion, in order to provide guidance on its design and operation.

In Chapter 9, the conclusions of the thesis are summarized, and the guidelines for future work are provided.

An overview of the thesis is shown in Figure 1-1.
Development of membrane reactors for heterogeneously catalysed aerobic oxidation of alcohols

Catalyst deactivation study in packed bed capillary microreactor (Chapter 3)

Teflon AF-2400 tube-in-tube membrane microreactor (Chapter 4)

Measurement of gas solubility using a tube-in-tube membrane contactor (Chapter 5)

Teflon AF-2400 flat membrane microreactor

Experimental study with Au-Pd/TiO₂ (Chapter 6)

Continuous stirred membrane reactor (Chapter 8)

Experimental/Modelling study with Ru/Al₂O₃ (Chapter 7)

Figure 1-1 Thesis overview.
CHAPTER 2. Literature Review

Membrane microreactors, which combine the benefits of both microreactors and membrane reactors, have recently drawn growing interest. In this chapter, the development in oxidation of benzyl alcohol and cinnamyl alcohol is presented in detail. The status of microreactors and membrane reactors is also reviewed, highlighting the application in three-phase reactions. Finally, advancing aerobic oxidation of alcohols by continuous flow Teflon AF-2400 membrane reactors are proposed.
2.1 Aerobic Oxidation of Alcohols

2.1.1 Introduction

Up to now, several books and reviews about oxidation of alcohols on supported metal catalysts have been published, and different reaction mechanisms have been proposed.\textsuperscript{3, 4, 6, 19-21} One of the reaction mechanisms is suggested to follow the classical dehydrogenation mechanism (shown in Figure 2-1).\textsuperscript{4, 22} It involves a first dehydrogenation of alcohols to produce adsorbed metal alkoxide and metal hydride. The metal alkoxide is subsequently dehydrogenated into the corresponding aldehyde through β-hydride elimination. Last, the metal hydride reacts with dissociatively adsorbed oxygen to regenerate the metal surface. In this chapter, oxidation of benzyl alcohol and cinnamyl alcohol, two of the most intensively investigated alcohols, is specifically discussed.

\textit{Figure 2-1 Mechanism of oxidation of alcohols on supported metal catalysts.}\textsuperscript{3} Reproduced from ref. 3 with permission from The Royal Society of Chemistry.
2.1.2 Oxidation of Benzyl Alcohol

Oxidation of benzyl alcohol has been intensively investigated as a model reaction, due to the relatively high reactivity of benzyl alcohol and few possible side products.\textsuperscript{23} It is also used as a reference test of the activity of gold catalyst.\textsuperscript{24} Benzaldehyde is mainly used in food industry as a flavouring additive, as well as in pharmaceutical and perfume industries.\textsuperscript{25} Various types of supported metal catalysts, such as Au, Pd, Ru and Ag, have been thoroughly studied, and their activities under different reaction conditions are frequently compared using turnover frequency (TOF).\textsuperscript{3,4}

Notably, Hutchings and co-workers\textsuperscript{26} have developed a highly active Au-Pd/TiO\textsubscript{2} for solvent-free oxidation of alcohols using a conventional impregnation method. The high catalytic efficiency was attributed to the formation of an Au-rich core with a Pd-rich shell, which facilitated the electronic promotion for Au on Pd. In their following studies, the effect of synthesis methods on the bimetallic nanoparticles morphology, particle size distribution, and the nanoparticles composition was also intensively studied.\textsuperscript{27}

Generally, the conventional impregnation method is employed through impregnating a support with a solution of metal salts, followed by heat treatment. Although this is a facile way for catalyst preparation, it could control the particle morphology during alloy formation and synthesize a complex range of nanostructures, which makes this method ideal for catalyst discovery. Au-Pd based catalysts synthesized by the conventional impregnation method were active and selective for oxidation of benzyl alcohol. However, this method cannot control the particle compositions or the particle size distribution. It was supposed that only a
small proportion of the nanostructures in the prepared catalyst played an important role in the reactivity.\textsuperscript{28} To identify and synthesize the active species, a sol-immobilization method has been developed.\textsuperscript{29-31} The typical procedures are preparation of metal salts solution with stabilizer ligands, reduction of metal nanoparticles with reducing agents, impregnation, and drying. A narrow range of particle sizes could be produced in the homogeneous sol, and the following supporting process would not significantly increase the mean particle size. Thus, the prepared catalyst showed higher activity than those prepared by traditional impregnation method. However, a systematic variation for random alloy nanoparticles was also observed in the composition of the nanoparticles, which resulted to much lower selectivity to benzaldehyde and higher amount of benzoic acid and benzyl benzoate. To control the nanoparticle composition, a modified impregnation method was investigated, where additional halide was added to the impregnating solution.\textsuperscript{32,33} A reduction treatment with hydrogen was also introduced to remove the halide after the heat treatment. With the help of excess halide, the dispersion of gold on the support was significantly improved. The control of the nanoparticle size and composition also became possible. The catalysts prepared with modified impregnation were extraordinarily active and stable for solvent-free aerobic oxidation of benzyl alcohol.\textsuperscript{33}

The reaction routes of benzyl alcohol oxidation on Au-Pd/TiO\textsubscript{2} were reported to involve a variety of reactions: oxidation to form benzaldehyde, benzoic acid and benzyl benzoate; disproportionation to form benzaldehyde, toluene, and water; dehydration to form dibenzyl ether; and self-condensation to form anthracene and stilbene.\textsuperscript{34} Among these, oxidation and disproportionation of benzyl alcohol were
considered to be the two main pathways. Two main products were also observed. Benzaldehyde, the target product, could be formed by both the oxidation (equation 2-1) and the disproportionation (equation 2-2) of benzyl alcohol, while another main product, toluene, could be only obtained from the disproportionation.

\[
2\text{PhCH}_2\text{OH} + \text{O}_2 \rightarrow 2\text{PhCHO} + 2\text{H}_2\text{O} \quad (2-1)
\]

\[
2\text{PhCH}_2\text{OH} \rightarrow \text{PhCHO} + \text{PhCH}_3 + \text{H}_2\text{O} \quad (2-2)
\]

The active site for the disproportionation reaction was suggested to be at the interface between the bimetallic nanoparticles and the TiO\textsubscript{2} support. Au sites alone were not active for the disproportionation reaction. Pd sites were more active for both reactions but only after an extended induction period. The addition of Pd to Au created a synergy with the overall activity increasing by 50-fold at the expense of toluene formation. Increasing oxygen pressure could monotonically accelerate the oxidation reaction within the pressure range studied. However, a maximum could be observed for the reaction rate of disproportionation, which suggested different mechanisms for the disproportionation reaction. Additionally, a rise in reaction temperature was shown to promote both oxidation and disproportionation reactions. The promotional effect was stronger on disproportionation than that on oxidation, resulting to the decrease of benzaldehyde selectivity at higher reaction temperature.

Ru-based catalysts have also been investigated for oxidation of benzyl alcohol, featured by low activity but high selectivity. The selectivity to benzaldehyde was often reported to be >99\% under moderate reaction conditions. Yamaguchi et al. studied the kinetics and mechanism of alcohol oxidation on Ru/Al\textsubscript{2}O\textsubscript{3} catalyst in a batch reactor. The reaction rate was indicated to be of fractional order with the
benzyl alcohol (first-order at low concentrations and zero-order at high concentrations; alcohol concentration range: 0-2 M) and a zero order on the oxygen pressure (0.2-3.0 bar). Using a multichannel packed-bed reactor, Bavykin et al.\textsuperscript{41} observed that the yield of benzaldehyde in oxidation of benzyl alcohol (1 M) on Ru/Al\textsubscript{2}O\textsubscript{3} catalyst was affected by the oxygen pressure at pressure lower than 9 bar and independent of the oxygen pressure at pressures above 9 bar. Zotova et al.\textsuperscript{11} studied aerobic oxidation of alcohols to aldehydes and ketones on Ru/Al\textsubscript{2}O\textsubscript{3} catalyst in a commercially-available XCube\textsuperscript{TM} reactor by pre-mixing and saturating the liquid with the gaseous reactant before reaching the catalyst bed. The oxidation of benzyl alcohol was indicated to be a pseudo-first-order with respect to benzyl alcohol (up to 0.2 M), and oxygen was found to be integral to the kinetics of the reaction, since the reaction rate was significantly enhanced by increasing the oxygen pressure from 6 to 26 bar. In a following study, the reaction rate was re-examined in a plug flow differential reactor with an inline FTIR, and was found to be of zero order with benzyl alcohol (0.1 M) under reaction conditions not subject to mass transfer limitations, and a partial positive order in oxygen (oxygen saturation pressure up to 26 bar).\textsuperscript{42}

An initially fast deactivation process was also observed on Ru/Al\textsubscript{2}O\textsubscript{3} catalyst, followed by a substantially stable state.\textsuperscript{38, 41} This was attributed to the poisoning of the active sites by benzoic acid formed from the over-oxidation of benzyl alcohol. The regeneration of the used catalyst was attempted through washing with base; however, full recovery was not obtained.\textsuperscript{41} Only minor improvements were attained when using different solvents or basic additives.\textsuperscript{38} Brazier et al.\textsuperscript{42} further derived an equation for the effect of oxygen on Ru/Al\textsubscript{2}O\textsubscript{3} catalyst deactivation, which accounted
for a fast and reversible inhibition by the formed benzoic acid, combined with a slower and irreversible loss of active sites caused by Ru reduction.

### 2.1.3 Oxidation of Cinnamyl Alcohol

![Reaction network proposed from Baiker and co-workers.](image)

**Figure 2-2** Reaction network proposed from Baiker and co-workers. Reproduced from ref. 43 & 44 with permission from Elsevier.

Oxidation of cinnamyl alcohol is also often used as a model reaction for alcohol oxidations, since cinnamyl alcohol is a simple aromatic allylic alcohol. The target product, cinnamaldehyde, is an important chemical intermediate in organic transformations and has wide applications in the food and perfume industries.

Up to now, studies of cinnamyl alcohol oxidation have mainly focused on monometallic catalysts, and platinum-group metals (particularly Pd) have been intensively investigated. Due to the co-existence of reactive hydroxyl and alkenyl functional groups in cinnamyl alcohol, complex reaction pathways are possible in this reaction, depending on the catalyst used and reaction conditions (Figure 2-2). Apart from cinnamaldehyde arising from dehydrogenation of cinnamyl alcohol, 3-phenyl-1-propanol and trans-β-methylstyrene can also be
produced through hydrogenation and hydrogenolysis reactions, respectively. These three products could potentially further lead to 1-phenylpropane, styrene, 3-phenyl-1-propanal, ethylbenzene, and benzaldehyde.

Marked catalyst deactivation has been observed by comparison of the reaction rates at different reaction times.\(^4^3,\,^5^1\) However, the active phase for the reaction, as well as the reason for the deactivation, is still a matter of debate. Baiker and co-workers\(^4^3\) demonstrated that metallic Pd was much more active than oxidic Pd in cinnamyl alcohol oxidation, which is consistent with the well-known dehydrogenation mechanism of selective alcohol oxidation to aldehydes on platinum-group metals catalysts.\(^2^2\) Deactivation caused either by the products from decarbonylation reactions poisoning the catalyst or by over-oxidation of the metal was suggested. The major role of oxygen was thought to rapidly and continuously remove the strongly adsorbed CO and thus ensure free Pd sites for the dehydrogenation reaction.\(^4^3\) On the other hand, Lee and co-workers\(^5^8,\,^5^9\) found Pd-based catalyst activity directly correlated with the proportion of exposed surface palladium oxide. The rapid reduction of PdO to Pd led to catalyst deactivation under static oxygen.\(^5^1\) Flowing oxygen at ambient pressure contributed to suppressing the reduction of palladium oxide to metal, which ensured the maintenance of a selective oxidation activity.

Compared with monometallic catalysts, bimetallic catalysts provide an attractive approach for catalyst development. Preparation of bimetallic catalysts and their use for the transformation of bio-renewable substrates, including selective oxidations, has been reviewed recently by Sankar et al.\(^6^0\). Dimitratos et al.\(^4^5\) found that Au-promoted Pd/C catalysts showed a significantly enhanced activity in the selective
oxidation of cinnamyl and benzyl alcohols, when compared to monometallic catalysts. However, no information about catalyst reusability was provided.

Generally, heterogeneous oxidation of cinnamyl alcohol has been carried out in batch reactors, due to their simplicity and flexibility. However, catalyst deactivation cannot be easily identified and monitored through a single run in batch reactors. To check the stability, catalysts need to be filtered and re-used in another run. Even with such experiments though, transient leaching of supported metal may not be detected easily, since the leached metal could redeposit onto the support.

2.1.4 Challenges for Aerobic Oxidation of Alcohols

Although significant progress has been attained on heterogeneous oxidation of alcohols, there is still a need to expand the scope of the reactions with high selectivities. During this process, catalyst deactivation is another issue that should not be ignored. Several mechanisms of catalyst deactivation have been identified for oxidation of alcohols, such as sintering, leaching, over-oxidation and/or poisoning of the active metal. Unfortunately, catalyst deactivation is barely reported with a specific cause in the literature. Overall, the development of aerobic oxidation of alcohols still necessitates more novel materials and methods for catalyst preparation and characterization, as well as efficient reactors for catalyst evaluations.

Additionally, how to establish safe operational conditions is of paramount importance for performing aerobic oxidation of alcohols at scale, since it involves a combination of oxygen with flammable materials under high pressure and temperature. Extensive risk assessment is still required to address the associated hazards (such as exothermic reaction and flammability of solvents) and develop
aerobic oxidation methodologies for organic solvents. A good understanding of the underlying mass transfer and reaction processes in the reactor can potentially prevent the explosive hazards currently hindering the wider application of aerobic oxidation of alcohols. Herein, the development and adaptation of continuous flow techniques for aerobic oxidation of alcohols would contribute to its future application in the fine chemicals and pharmaceutical industries.

2.2 Microreactors

2.2.1 Introduction

Microreactors are one type of chemical reactors with characteristic dimensions generally smaller than one millimeter. Therefore, microreactors have high surface-area-to-volume ratio, which is between 10,000 m²/m³ and 50,000 m²/m³, as compared to conventional reactors (normally less than 1,000 m²/m³). This contributes to several significant advantages:

1. Improved heat-and mass-transfer

The heat-transfer process in microreactors is significantly enhanced, since the heat-transfer coefficient is inversely proportional to the channel dimension. For highly exothermic or endothermic reactions, microreactors enable rapid removal or supply of heat. Hence, isothermal reaction conditions can be realized and local hot spots potentially as an ignition source could also be avoided. Similarly, the mass transfer in microreactors is also considerably improved, since the mixing times in microreactors are generally smaller than those in conventional systems and the diffusion times are shorter due to the smaller dimensions.

2. High-resolution reaction time control
The small dimensions also decrease the radial diffusion in the microreactors, which ensures a narrow residence time distribution. Simultaneously, the residence time for the reactants in the microreactors can be precisely controlled by changing the flow speed or the length of the channels.

(3) Small volumes

The small volumes of microreactors make it possible to use less amount of reagents and catalysts for testing or processing. This directly decreases the economical cost. Additionally, reaction parameters (pressure, temperature, etc.) are easier to be controlled in small volumes, which enhances the reactor safety. The power of an explosion, if happened, could also be minimized in microreactors.

(4) Easier for modelling and scale-up

The hydrodynamic flows in microreactors are commonly laminar, which allows the accurate simulation and prediction by numeral methods. Compared with the complexity of scale-up for conventional reactors, microreactors are easier to scale up through numbering up without changing the channel geometry.

Although there are several advantages in microreactors, the disadvantages in such type of reactor should not be ignored. Due to the small diameter channels, high pressure drops can be developed and particles cannot be simply handled in microreactors. Practically, it is difficult to utilize microreactors to produce substantial amounts of materials, even when massive parallelisation is achieved.

2.2.2 Microreactors for Three-Phase Reactions

Up to now, several types of microreactors have been developed for three-phase reactions. Based on how solid catalysts are distributed within the reactor,
conventional microreactors can be classified to packed bed microreactors, monolithic microreactors, wall-coated microreactors, and slurry-flow microreactors.

### 2.2.2.1 Packed Bed Microreactors

Packed bed reactors are commonly realized through direct packing and retaining catalyst particles in a cartridge, tube, or reactor channel. Thus, this approach has a wide range of catalyst options, together with the simplicity to characterize and quantify the catalyst. Traditional or already optimized catalysts can also be easily applied into such reactors. Gas and liquid are pre-mixed before reaching the packed bed, where stable flow patterns (e.g. bubble flow, slug flow) are developed. However, when the flow reaches the bed, it is difficult to characterize and control the fluid dynamics within the packed bed. Mass-/heat- transfer limitations may exist in the bed for exothermic or fast reactions. Pressure drop could also develop along the bed, particularly for small particles. At high reactor-to-particle diameter ratios, pressure drop is determined by capillary force. These should draw enough attention for reactor design, especially for pressure-sensitive reactions.

Several packed bed microreactors have been reported for aerobic oxidation of alcohols. Plucinski and co-workers designed and investigated a multifunctional compact reactor for continuous aerobic oxidation of benzyl alcohol. The reactor consisted of heat-transfer channels, static mixers and mm-scale packed bed reaction channels (shown in Figure 2-3). Due to the enhanced mass transfer in the reactor, the reaction was observed to be in the kinetic regime over a wide range of operating conditions.
A silicon-glass packed bed microreactor (shown in Figure 2-4) was developed in our group for solvent-free aerobic oxidation of benzyl alcohol. The channel dimensions within the reactor were 190 mm (L) × 0.6 mm (W) × 0.3 mm (H), and solid catalysts were packed at the end of the channel. The performance of the microreactor in terms of benzyl alcohol conversion and selectivity to benzaldehyde was very close to that in a conventional glass stirred reactor. External mass transfer resistance was indicated at low flow rate and internal mass transfer resistance was
observed for large catalyst particles. Furthermore, the disproportionation of benzyl alcohol to equimolar amounts of benzaldehyde and toluene was experimentally confirmed, which was difficult to realize in batch reactors. In a following study, four parallel micropacked bed reactors were fabricated on a single chip and demonstrated for efficient evaluation of catalysts for three-phase reactions.\textsuperscript{70}

\begin{figure}[h]
\centering
\includegraphics[width=\textwidth]{fig2-5.png}
\caption{General schematic of the H-Cube\textsuperscript{TM} flow reactor (Thales Nanotechnology Inc.).\textsuperscript{71} Reproduced from ref. 71 with permission from The Royal Society of Chemistry.}
\end{figure}

A commercial H-Cube\textsuperscript{TM} flow reactor (shown in Figure 2-5) has been investigated for continuous-flow oxidation of alcohols.\textsuperscript{11,71,72} The safe and practical process for aerobic oxidation of alcohols was demonstrated by Zotova \textit{et al.}\textsuperscript{11} through pre-mixing and saturating the liquid with gaseous reactant (oxygen or air) before it reached the catalyst bed. Obermayer \textit{et al.}\textsuperscript{71} carried out oxidation of benzyl alcohol with Fe/Al-SBA15 catalyst within such reactor, and the effect of different co-catalysts and solvents was investigated. Continuous recirculation was also performed and full conversion of benzyl alcohol could be achieved. Notably, the solubility of oxygen in the solvent could potentially limit the reaction, and this approach is only suitable for dilute solutions under high pressures.\textsuperscript{7}
Aellig et al.\textsuperscript{73} also reported a commercial TEMPO/SiO\textsubscript{2} catalyst packed Teflon tube for continuous oxidation of alcohols. The catalyst was initially demonstrated to stabilize for several hours in oxidation of benzyl alcohol and then applied to a wide range of primary and secondary alcohols. Mannel et al.\textsuperscript{38} developed a continuous-flow process using Ru/Al\textsubscript{2}O\textsubscript{3} catalyst for oxidation of alcohols. Diluted oxygen source (8-9\% oxygen in nitrogen) was used to avoid the oxygen/organic mixture entering the explosive regime. A good relative correlation of reaction rate constant was shown for the catalyst within the batch reactor and the continuous flow. After illustrating the catalyst activity profile, high steady-state conversion and single-pass yields could be achieved for oxidation of benzyl alcohol.

\textbf{2.2.2.2 Monolithic Microreactors}

To overcome the drawbacks in packed bed microreactor, monolithics are proposed to be applied in continuous-flow processes.\textsuperscript{65, 66, 74} Monolithic materials are usually a uniform block of structured metallic, inorganic or polymeric materials filled with interconnected pores.\textsuperscript{74, 75} The large flow-through macropores (typically 2 to 10 mm) allow a low pressure drop or high flow rate, whereas the meso- or micropores provide a large surface area (>300 m\textsuperscript{2}/g) for reaction. Through post-functionalization, catalytic sites can be anchored or loaded to the pores of monoliths. Although high productivity could be potentially realized in monolithic microreactors, several disadvantages are also observed, such as pore clogging, non-uniform radial permeability, and unutilized catalytic sites deeply buried inside the pores.\textsuperscript{65}

For applications in aerobic oxidation of alcohols, Pollington et al.\textsuperscript{76} investigated Au/C coated monolith catalyst housed inside a stainless steel reactor for oxidation of glycerol. A significant enhancement in the reaction rate was observed for the
monolith catalyst as compared to Au/C powder catalyst conducted in a autoclave batch reactor. Badran et al.\textsuperscript{77} incorporated catalytic monoliths to a continuous flow reactor for oxidation of benzyl alcohol. The catalytic monoliths were prepared through impregnating carbon monoliths with Pt. Both single and multi-tube monolith reactors were investigated to demonstrate the viability of this kind of reactor at pilot-scale.

\textbf{Figure 2-6} Schematic of the reactor system with paper-structured Ru catalyst.\textsuperscript{78} Reproduced from ref. 78 with permission from Elsevier.

Kitaoka and co-workers\textsuperscript{78-80} developed a paper-structured catalyst with micrometer-sized pores for solvent-free alcohol oxidation (shown in Figure 2-6). The paper-structured catalysts were fabricated using ceramic fibers by a wet papermaking process, followed by the impregnation of active Ru species. As compared to conventional pellet- and bead-type solid catalysts, higher conversion and selectivity were observed with the paper catalysts, which were attributed to the uniform distribution of the liquid phase and the efficient gas-liquid-solid mass transfer by the microfluidic effect.
2.2.2.3 Wall-Coated Microreactors

Wall-coated micoreactors are developed through immobilizing catalytic species on the wall with special coating procedures, such as sputtering, impregnation, or cross-linking. Because of the well-defined flow geometries, this approach has several advantages over packed bed and monolithic microreactors, such as substantially minimized pressure drop and potential prediction of the fluid dynamics. It should also be noted that the catalyst loading in wall-coated microreactors is generally lower than those in the above two types of microreactors. So, the microchannel surface area is necessary to be increased through chemical treatment of the walls or deposition of inorganic porous materials.

![Diagram of microstructured falling film reactor](image)

**Figure 2-7 Components and schematic of the microstructured falling film reactor.** reproduced from ref. 81 with permission from Elsevier.

Yeong et al. described a falling film microreactor for hydrogenation of nitrobenzene to aniline (shown in Figure 2-7). The reactor was fabricated using a stainless steel plate with 64 straight, parallel channels (300 μm wide, 100 μm deep, 78 mm long separated by 100 μm wide walls), and the catalyst could be coated on the plate with four different methods (sputtering, UV-decomposition, wet
impregnation, and incipient wetness). The overall mass transfer coefficient was estimated to be 3-8 s\(^{-1}\) with surface-area-to-volume ratio 9,000-15,000 m\(^2\)/m\(^3\).

Protasova et al.\(^8\) studied the kinetics of citral hydrogenation on Au/TiO\(_2\) and Pt-Sn/TiO\(_2\) thin films deposited within silica capillaries. The reaction orders were observed to be 1 for hydrogen and near zero for citral concentration.

Kobayashi and co-workers have developed a new approach for fabrication of metal-immobilized capillary column reactor and applied it to various oxidative processes.\(^8\) Metallic clusters were first microencapsulated and then cross-linked to the wall with polystyrene-based copolymers. The resultant cross-linked polymer backbones were proved to efficiently stabilize the nanoclusters without aggregation or leaching. Thus, high catalytic activity could be maintained for a series of aerobic oxidation reactions. Liu et al.\(^8\) reported a nano-gold catalyst coated stainless steel microcapillary for oxidation of cyclohexane. 2.1% conversion of cyclohexane was obtained at reaction temperature 180 °C, a residence time of 4 min and reaction pressure of 3 MPa, with a total selectivity of 90.6% to value products. Feng et al.\(^8\) also investigated a PTFE capillary tube coated with Pd nanocatalysts for three-phase nitrobenzene hydrogenation. The coated microreactor showed a more than 97% averaged conversion of nitrobenzene under different inlet concentrations from 30-90 mM in a 40 h continuous operation.

### 2.2.2.4 Slurry-Flow Microreactors

Slurry-flow microreactors are a type of microreactors where solid catalysts are continuously mixed and delivered with the reactants. The formation of slug flow is preferred, since the internal circulations in the liquid slugs can keep catalyst particles in motion and free from clogging.\(^8\) The large interfacial areas also provide good
mass and heat transfer performance. However, the feeding of liquid/solid slurries and the handling of particles in microreactors are still challenging, since the interaction among particles, fluid, and reactor surface is complex.

Enache et al. first proposed the three-phase slurry-flow microreactor for hydrogenation reaction. The resorcinol hydrogenation was shown to be operating under kinetic control within such reactor, and the reaction rate was enhanced by a factor of 7, as compared to that within an autoclave. However, the details on the flow feeding and flow patterns were not provided.

Figure 2-8 Gas-liquid-solid “slurry Taylor” flow: a) experimental set-up and b) typical “slurry Taylor” flow obtained for a N2-water with 6.0 g/L of 100 μm Al2O3 particles in suspension (flow velocity of 3.6 cm/s). Reproduced from ref. 92 with permission from Elsevier.
Liedtke et al.\textsuperscript{92} evaluated a slurry-flow microreactor for hydrogenation of 3-methyl-1-pentyne-3-o with Pd/SiO\textsubscript{2} catalyst. The slurry flow was delivered through injecting water into a polypropylene bag to expand it and push the suspension out of a stirred glass vessel (shown in Figure 2-8). The operation parameters (catalyst charge, particle size, liquid flow rate, \textit{etc.}) were studied to allow a stable and controlled gas-liquid-solid “slurry Taylor” flow in PFA capillary tube. The reaction results obtained from slurry reactor presented a good repeatability and were comparable to those from a stirred tank vessel, which demonstrated its good mass transfer performance.

Alex et al.\textsuperscript{93} carried out oxidation of benzyl alcohol in a PTFE tube with PVP-stabilised Pd and Au/Pd nanoparticles. The nanoparticles were dispersed in a \textit{i}PrOH/H\textsubscript{2}O mixture and delivered via syringe pumps. After reaction, the dispersed nanoparticles could be recovered through phase separation and precipitation. A conversion of 91\% was obtained with a selectivity of 97\% to benzaldehyde at 50 °C, 1 bar air pressure, and mean residence time of 15 min.

2.3 Membrane Reactors

2.3.1 Introduction

Membrane reactors are one type of reactors coupled with membranes, which integrate a chemical reaction with a membrane process in the same unit. The membranes can be made from different inorganic or organic materials such as metals, carbon, ceramics and polymers.\textsuperscript{94} As part of the reactor, the membranes generally possess specific structures, allowing at least one component to permeate but the others rejected. Thus, they can function as an extractor (selectively separate the
products from the reaction mixture), a distributor (controllably add reactants to the reaction mixture) or a contactor (intensify the contact of reactants with catalyst). Accordingly, the process productivity and/or the process safety can be enhanced.

**2.3.2 Membrane Reactors for Three-Phase Reactions**

Membrane reactors have been applied to various three-phase reactions, and an extensive review has been given by Comite *et al.* For example, Singh *et al.* have demonstrated a Pt-decorated polymeric membrane reactor for partial hydrogenation of soybean oil. The membrane was prepared through sputtering platinum onto an integral-asymmetric polyetherimide membrane. During the reaction, the oil was circulated across the sputtered side of the membrane and hydrogen was supplied from the other side. Obvious lower trans-fatty acid values were observed under similar conditions as compared to the conventional slurry reactor, which were possibly caused by the rapid hydrogen permeation/supply through the high performance membrane.

Pashkova *et al.* studied Pd-coated tubular ceramic membranes for safe synthesis of $\text{H}_2\text{O}_2$ directly from $\text{H}_2$ and $\text{O}_2$. Productivities up to $1.7 \text{ mol}_{\text{H}_2\text{O}_2}/(\text{g Pd} \cdot \text{h})$ could be achieved with a selectivity of 83%. Additionally, insufficient transport of the reactants to the catalytic zone was found to be a serious problem, when pure laminar flow dominated in an empty membrane channel. Glass beads were filled into the channel to guarantee more turbulent flow, which contributed to a significant increase in $\text{H}_2\text{O}_2$ concentration. Catalyst deactivation was also observed, presumably due to partial change of the Pd oxidation state.

Dotzauer *et al.* prepared a catalytic ceramic membrane through layer-by-layer
(LBL) adsorption of polyelectrolyte/metal nanoparticles films and applied it for wet air oxidation of formic acid, acetic acid and phenol. This special preparation method guaranteed catalyst deposition only near the interior of the tube, where gas-liquid interface generally located. Thus, normalized reaction rate in the LBL membrane was 2-3 times higher than those in membranes prepared with traditional methods (evaporation or anionic impregnation). Rezac and co-workers\textsuperscript{100} also fabricated catalytically active ceramic/polymer membranes through depositing a polymer-palladium complex onto the inner skin of the membrane. Nearly zero order in hydrogen was shown for partial hydrogenation of soybean oil, proving the sufficient supply of hydrogen through the membrane. Continuous oxidation of benzyl alcohol was also investigated in the ceramic membrane reactor in our group.\textsuperscript{101} A commercial ceramic membrane was used as gas and liquid contactor, and Au-Pd/TiO\textsubscript{2} catalysts were packed inside the tubular membrane. Improved conversion and selectivity were shown at higher oxygen pressure, temperature, and catalyst contact time. Though the performance of the membrane reactor was comparable to simulated trickle bed operation, the rate of oxygen supply through the membrane was indicated to be potentially insufficient for fast reactions.

Rezac and co-workers\textsuperscript{102} also investigated a porous PTFE membrane deposited with Ru catalyst particles for aqueous phase hydrogenation of levulinic acid. The highly hydrophobic PTFE material acted well as a barrier to the aqueous phase while allowing hydrogen to permeate and contact with catalysts at the liquid/membrane interface. Due to the fundamentally different method of hydrogen delivery, the membrane reactor was shown to be more catalytically efficient than a packed bed reactor.
2.3.3 Membrane Microreactors for Three-Phase Reactions

More recently, membrane microreactors have been an emerging type of membrane reactors for process intensification, since they combine the benefits of both the membrane reactors and the microreactors.\textsuperscript{103, 104} High heat and mass transfers, due to high surface/membrane-area-to-volume ratio and short transport distance, are guaranteed in the miniaturized membrane reactors, which contribute to enhanced reactor performances as compared to conventional membrane reactors.

An overview of membrane microreactors has been given by Seelam \textit{et al.}\textsuperscript{105} Particularly for three-phase reactions, Aran \textit{et al.}\textsuperscript{106} have developed a porous ceramic mesoreactor through catalyst deposition and surface modification on commercial $\alpha$-Al$_2$O$_3$ hollow fibers (Table 2-1, entry 1). The Pd catalyst deposition was realized on the fibers with an impregnation method, and the wetting behaviour of the reactor surface was tuned with hydrophobization techniques, which allowed a stabilized gas-liquid-solid interface. Catalytic hydrogenation of nitrite ions was tested as a model reaction in the fabricated reactor, and the results showed that the configuration with hydrophobized membrane support and hydrophilic catalyst support was the most effective. It should also be noted that the pressure difference between gas and liquid phases should be punctually controlled to ensure the stabilized interface without one phase dispersing into the other.

The internal and external mass transfers in this type of reactor were further experimentally investigated in their following study (Table 2-1, entry 1).\textsuperscript{107} The conversion rates per active catalyst surface area increased significantly with the decrease of the catalyst support layer, indicating the existing of internal mass transfer limitations caused by small pore sizes and high tortuosity of the support.
Furthermore, the external mass transfer was improved through reducing the inner diameter or the integrating slug flow in the liquid side. A 2D CFD model was also proposed to study the effects of a wide range of reaction and geometrical parameters on the reactor performance.\textsuperscript{108} Nitrite hydrogenation in the membrane microreactor was well described by Langmuir-Hinshelwood reaction kinetics, and the reactant concentration could also be derived as a function of catalytic membrane layer thickness. Small reactor geometries were suggested to enhance mass transfer to reactor boundaries.

![Figure 2-9 Carbon nanofibers decorated porous stainless steel membrane encapsulated with PDMS layer.\textsuperscript{109} Reproduced from ref. 109 with permission from Elsevier.](image)

Aran \textit{et al.}\textsuperscript{109} also developed a hybrid porous stainless steel membrane for nitrite hydrogenation (Table 2-1, entry 2). The membrane consisted of porous stainless steel tube, Pd deposited carbon nanofibers, and dense PDMS layers (shown in Figure 2-9). The preparation steps included the fabrication of porous stainless steel hollow fiber, the growth of carbon nanofibers on the stainless steel surface, the immobilization of palladium catalyst, and the encapsulation with gas permeable polymer coating. Therefore, the fabricated membranes possessed high surface area, mechanical strength, and catalytic activity for nitrite reduction, which made them promising for wider application in multiphase technologies.
### Table 2-1 Membrane microreactors for three-phase reactions

<table>
<thead>
<tr>
<th>Entry</th>
<th>Reaction</th>
<th>Membrane material</th>
<th>Catalyst</th>
<th>Highlights</th>
<th>Reference</th>
</tr>
</thead>
</table>
| 1     | Hydrogenation of nitrite ions | Pd/γ-Al₂O₃ coated α-Al₂O₃ hollow fiber  
Average pore diameter: 800 nm; fiber length: 13.5 cm; ID: 0.8 mm or 2.8 mm; OD: 2.8 or 3.8 mm | Pd/γ-Al₂O₃ | • The position of G-L interface tuned with hydrophobization techniques  
• Experimental study on the mass transfer limitations | Aran et al.¹⁰⁶,¹⁰⁷, 2011,2012 |
| 2     | Hydrogenation of nitrite ions | Pd/C and PDMS membrane coated porous stainless steel tube  
PDMS thickness: ~20 μm; tube ID: 0.8 mm; OD: 2.0 mm; Pt nanoparticles modified polysulfone hollow fiber  
Maximum pore size: 650 nm; fiber length: 15 cm; ID: 750 μm; OD: 1200 μm. | Pd/C | • Hybrid membrane with a controlled G-L-S interface and high surface area, mechanical strength, and catalytic activity  
• High conversion due to the short diffusion distance in the fiber lumen  
• Facile scale-up through numbering up | Aran et al.¹⁰⁹, 2011 |
| 3     | Oxidation of formic acid | PTFE tube  
tube length: 45.7 cm; ID: 1.6 mm; OD: 3.2 mm. | Ru/Al₂O₃ | • A “tube-in-shell” reactor with low-cost membrane materials and facile scale-up for safe oxidation | Greene, et al.⁶⁰, 2015 |
| 4     | Oxidation of alcohols | Pd/γ-Al₂O₃ coated and dopamine modified PDMS film  
Film thickness: 50-100 μm; reactor channel length: 652 mm; height: 210 μm; width: 500 μm. | Pd/γ-Al₂O₃ | • Enhanced mass transport in the membrane reactor with a large surface-area-to-volume ratio to | Liu et al.¹¹¹, 2016 |
| 5     | Hydrogenation of nitrobenzene | PTFE film  
Pore size: 100 – 500 μm; reactor channel width: 1.4 mm; height: 0.4-1.0 mm; | Pd/C | • Suspension-flow membrane microreactor with enhanced mass transfer and optimized selectivity | Selinsek et al.¹¹², 2016 |
Hogg et al.\textsuperscript{110} experimentally and theoretically studied a Pt-modified polysulfone hollow fiber as gas-liquid contactor for wet air oxidation of formic acid (Table 2-1, entry 3). Significantly higher conversion was experimentally observed in the modified fiber than that in prior ceramic membranes, due to the decreased short radial diffusion distances. Through modelling the diffusion of the reactant, the oxidation reaction rate was proved to be obviously lower than the diffusion-limited rate, which was caused by the disperse loading of Pt catalyst along the wall.

Greene \textit{et al.}\textsuperscript{113} developed a “tube-in-shell” membrane microreactor for homogeneous and heterogeneous oxidation of alcohols with molecular oxygen (Table 2-1, entry 4). The inexpensive Teflon tubing was applied to provide oxygen for the reaction and avoided hazard mixtures of oxygen and organic vapors. The reactor performance was shown to be highly determined by the ratio of membrane surface area to volume rather than the polymer type or thickness of the membrane.

Liu \textit{et al.}\textsuperscript{111} demonstrated a catalytic membrane microreactor for hydrogenation of nitrobenzene with dopamine modified and Pd/\textgamma-Al\textsubscript{2}O\textsubscript{3} coated PDMS film (Table 2-1, entry 5). The nitrobenzene conversion was shown to be enhanced, when low liquid flow rate, high gas flow rate, and thinner membrane were applied. This was due to the increased residence time and hydrogen permeation rate through the membrane.

Selinsek \textit{et al.}\textsuperscript{112} demonstrated a novel intensified suspension-flow membrane microreactor for direct synthesis of hydrogen peroxide (Table 2-1, entry 6). PTFE film was introduced to separate oxygen and hydrogen with the concern of safety issue, and Pd/C catalyst particles were suspended in the liquid flow. Higher productivity was achieved through increasing mass transfer in microchannels and keeping constant reactant feeding along the reactor channel. Simultaneously, the
selectivity was optimized by adjusting the oxygen and hydrogen distribution in the liquid phase.

2.3.4 Teflon AF-2400 Membrane Microreactors

Among the available membranes for membrane reactor fabrication, Teflon amorphous fluoropolymer 2400 (Teflon AF-2400) has drawn growing interest, since it uniquely combines excellent chemical resistance, thermal stability, and mechanical properties with high fractional free volume (shown in Table 2-2).\textsuperscript{114} Its permeability characteristics to gas and liquid also ensure high flux of gas and low pervaporation of liquid through the membrane. These advantages apparently make it potential to be a reliable contact interface for gas and liquid. The Ley group has first developed a tube-in-tube membrane contactor/microreactor, which comprises an inner Teflon AF-2400 tube and an outer PTFE tube (shown in Figure 2-10).\textsuperscript{115-118} Generally, liquid is pumped through the inner tube and gas is pressured at the annulus between the inner and outer tubes. When the liquid is flowing within the inner tube, the pressured gas can permeate through the semi-permeable tube and contact/react with the liquid. Compared with traditional mechanical mixing, this controlled method of gas-liquid contacting is a safe, well-controlled and reproducible approach for gas-liquid mixture or reaction. The reaction can also be simply optimized by modulating the gas pressure, the liquid flow rate and the Teflon tube dimensions.\textsuperscript{114,116} Since the transfer process consists of gas permeation solely through the Teflon AF-2400 tube and the laminar liquid flow, the mass transfer of gas in the tube-in-tube microreactor can also be accurately simulated, providing insights into the reactor operation.\textsuperscript{119}
### Table 2-2 Properties of Teflon AF-2400

<table>
<thead>
<tr>
<th>Property</th>
<th>Teflon AF-2400</th>
</tr>
</thead>
<tbody>
<tr>
<td>Crystallinity</td>
<td>None</td>
</tr>
<tr>
<td>Fractional free volume [%]</td>
<td>33.4</td>
</tr>
<tr>
<td>Permeability of O₂ [barrer]</td>
<td>1140</td>
</tr>
<tr>
<td>Glass transition temperature [ºC]</td>
<td>240 ± 10</td>
</tr>
<tr>
<td>Solubility</td>
<td>None in common organic solvents</td>
</tr>
<tr>
<td>Tensile strength at 23 ºC [MPa]</td>
<td>26.4 ± 1.9</td>
</tr>
</tbody>
</table>

![Figure 2-10](image)  
*Figure 2-10 Tube-in-tube membrane microreactor set-up.* Reproduced from ref. 120 with permission from John Wiley and Sons.

Various gas-liquid reactions have been successfully performed in this type of reactor, such as ozonolysis\textsuperscript{116} and hydrogenation\textsuperscript{121} of alkenes, synthesis of styrenes using ethylene\textsuperscript{115}, and carboxylation of Grignard reagents with CO\textsubscript{2}\textsuperscript{117}. For three-phase reactions, Skowerski et al.\textsuperscript{122} reported heterogeneous olefin metathesis in a tube-in-tube microreactor through packing the solid catalyst in the annulus between the inner and outer tubes, which improved yield and selectivity due to the efficient removal of ethylene during the reaction. O’Brien et al.\textsuperscript{123} carried out three-phase
heterogeneously catalytic reactions, where the liquid was first saturated with gas in a tube-in-tube configuration and subsequently reacted in a catalyst packed column. Chaudhuri et al.\textsuperscript{124} also employed the Teflon AF-2400 tube-in-tube contactor to saturate the liquid stream with oxygen/air for catalytic oxidation of alcohols. However, with this approach the amount of gas available for the reaction is limited by its solubility in the reaction mixture as it is not supplied in the downstream reaction section.\textsuperscript{119} This may severely affect the productivity of the tube-in-tube microreactor. Nevertheless, Teflon AF-2400 membrane reactors have been shown of great potential to facilitate the application of aerobic oxidation of alcohols.
CHAPTER 3. Oxidation of Alcohols using Bimetallic Au-Pd/TiO$_2$ Catalysts: a Deactivation Study in a Continuous Flow Packed Bed Microreactor$^*$

The stability of a bimetallic Au-Pd/TiO$_2$ catalyst was examined in a packed bed microreactor for aerobic oxidation of alcohols. The catalyst was prepared by co-impregnation with an Au-Pd weight ratio of 1:19. Experiments were performed at 80-120 °C, oxygen concentration 0-100% and total pressure 4 bar. Although the same catalyst was shown to possess good stability in the oxidation of benzyl alcohol, it deactivated during the oxidation of cinnamyl alcohol, particularly at elevated reaction temperatures. Higher concentration of oxygen used for the reaction led to improved cinnamaldehyde selectivity but lower conversion and higher deactivation rates. Deactivation was attributed to Pd leaching and a complex effect of oxygen.

3.1 Introduction

In recent years, there has been a growing demand for the development of heterogeneous catalysts for oxidation of alcohols with molecular oxygen, since it is a green and efficient process. However, large-scale applications of aerobic alcohol oxidation are still limited to few catalysts with sufficient reactivity and stability. Generally, heterogeneous oxidation of alcohols has been carried out in batch reactors, due to their simplicity and flexibility. Nevertheless, catalyst deactivation cannot be easily identified and monitored through a single run in batch reactors. To check the stability, catalysts need to be filtered and re-used in another run. Even with such experiments though, transient leaching of supported metal may not be detected easily, since the leached metal could redeposit onto the support.

Over the past decade, microreactors have emerged as an excellent alternative for chemical synthesis, due to high mass and heat transfer coefficients and high-resolution reaction time control. The small volumes of microreactors also make it possible to use smaller amount of both reagents and catalyst for the rapid assessment of catalysts. Liquid phase catalytic oxidation chemistry in continuous-flow microreactors has recently been summarized from both technological and chemical perspective by Gemoets et al.. In this study, the stability of a 1% bimetallic Au-Pd (5:95)/TiO2 catalyst is examined in a packed bed capillary microreactor for oxidation of alcohols with molecular oxygen. Various reaction conditions are investigated, in an attempt to explore the possible contributions to catalyst deactivation.

3.2 Experimental Section

3.2.1 Catalyst Preparation*

1 wt % bimetallic Au-Pd/TiO2 catalysts were prepared by a co-impregnation method. An aqueous solution of HAuCl₄·3H₂O (Johnson Matthey) and PdCl₂ (Johnson Matthey) with Au-to-Pd weight ratio equivalent to 1:19 was added to TiO₂ (Degussa, P25) to form a slurry. The resultant slurry was spray-dried (nozzle temperature 220 ºC), and then calcined in static air at 400 ºC for 3 h. The powder was pelletized, crushed, and sieved to the desired particle size range.

3.2.2 Capillary Microreactor Set-up

![Figure 3-1 Schematic of the capillary microreactor set-up used in this work.](image)

The schematic of the capillary microreactor set-up is shown in Figure 3-1. The gas flow was controlled with a mass flow controller (MFC, Brooks GF40) and the liquid feed was pumped with an HPLC pump (Knauer P 2.1S). Benzyl alcohol (Sigma-Aldrich, ≥99%) and 0.5 M cinnamyl alcohol (Sigma-Aldrich, 98%) dissolved in toluene (Sigma-Aldrich, ≥99.5%) were used as received. The gas and liquid flows were combined in a T-mixer (Upchurch) to generate a gas-liquid slug flow and fed into the reactor. The capillary microreactor was a 30 cm long PTFE

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*Catalyst preparation was carried out by Dr Peter Ellis, Johnson Matthey.

tube (inner diameter (ID): 0.8 mm; outer diameter (OD): 1.6 mm; Upchurch) packed with the Au-Pd/TiO₂ catalyst particles (particle size: 53–63 μm). Unless otherwise specified, each experiment was performed with fresh catalyst. The microreactor was submerged in a stirred oil bath for isothermal operation and the temperature of the oil bath was controlled and measured by a hot plate fitted with a thermocouple (Stuart). The catalyst was kept under flowing oxygen until the oil bath reached the desired reaction temperature and then the liquid was pumped into the reactor to start the reaction. A sample of the product mixture was collected in a 2 mL vial every 30 min, and a 50 mL vessel was connected in parallel to minimize the pressure drop after taking a sample. The system pressure was sustained by a back pressure regulator (BPR, Swagelok) at the outlet and measured with two pressure sensors (Zaiput). All pressures reported in this study are absolute.

The collected sample was quantitatively analysed by a gas chromatograph (Agilent 7820A) fitted with a DB-624 capillary column and a flame ionization detector. Alcohol conversion (X) and selectivity to each product were calculated according to the following equations.

\[
X = \frac{c_{\text{alcohol, in}} - c_{\text{alcohol, out}}}{c_{\text{alcohol, in}}} \times 100\% \quad (3-1)
\]

\[
S_i = \frac{c_i v_i}{c_{\text{alcohol, in}} X} \times 100\% \quad (3-2)
\]

where \( c_{\text{alcohol, in}} \) and \( c_{\text{alcohol, out}} \) are the alcohol concentrations (mol/L) at the inlet and outlet, respectively. \( v_i \) is the number of moles of alcohol consumed for the production of 1 mole of product \( i \). The cinnamaldehyde yield was calculated by the alcohol conversion multiplied by the selectivity to cinnamaldehyde.
3.2.3 Catalyst Characterization

X-ray photoelectron spectra (XPS) was carried out using a Thermo Scientific K-alpha photoelectron spectrometer with monochromatic AlKα radiation. Spectra were collected in steps of 0.15 eV and plotted with the CasaXPS software. Inductively coupled plasma (ICP) measurements were carried out on an Agilent 7900 ICP-MS equipped with a micromist nebuliser in organic phase mode. Quantification was carried out by comparison with calibration curves of Au and Pd.

3.3 Results and Discussion

3.3.1 Oxidation of Benzyl Alcohol as Benchmarking Experiment

The co-impregnation method employed in this study is one of the easiest ways to synthesize bimetallic catalysts. It is also considered to be ideal for the discovery of active catalyst phase, since a wide variation in nanostructures is available in the prepared catalyst. Oxidation of benzyl alcohol is often used as a model reaction for oxidation of alcohols. In order to verify the reactivity of the prepared Au-Pd/TiO₂ catalyst and the performance of the capillary microreactor system, oxidation of benzyl alcohol was initially carried out for benchmarking.

Conversion and selectivity of the oxidation of benzyl alcohol as a function of time are shown in Figure 3-2. It can be observed that the conversion of benzyl alcohol (88%) stabilized after 1 h and then stayed constant over the whole 30 h. The selectivities to benaldehyde and toluene, the two main products, also showed stable trends. Hence, the Au-Pd/TiO₂ catalyst exhibited high stability in this particular oxidation reaction. This is consistent with previous work. Furthermore, it has

* ICP analysis was carried out by Dr Gemma L. Brett, Cardiff University.
been indicated that benzaldehyde can be formed by both the oxidation and disproportionation of benzyl alcohol, while toluene is produced only from the disproportionation. Thus, the higher selectivity to benzaldehyde (67%) than that to toluene (25%) shows the good activity of Au-Pd/TiO₂ catalyst in the oxidation of benzyl alcohol. These results obtained in the packed bed capillary microreactor were similar to those in a silicon-glass packed bed microreactor.

![Figure 3-2](image)

**Figure 3-2** Solvent-free oxidation of benzyl alcohol as a function of reaction time. Reaction conditions: Au-Pd/TiO₂ catalyst, 20 mg; oxygen flow, 0.8 mL/min at STP; benzyl alcohol flow, 8 μL/min; reaction temperature, 120 °C; reaction pressure, 4 bar. X, conversion of benzyl alcohol; Sₖ, selectivity to benzaldehyde; Sₜ, selectivity to toluene.

### 3.3.2 Effect of Reaction Temperature on Oxidation of Cinnamyl Alcohol

For oxidation of cinnamyl alcohol, a blank experiment was first conducted at 100 °C with TiO₂ support and pure oxygen. The conversion was observed to be 9%, with a 22% selectivity to cinnamaldehyde. Then, the prepared Au-Pd/TiO₂ catalyst was studied for oxidation of cinnamyl alcohol at different reaction temperatures. Cinnamaldehyde, 3-phenyl-1-propanol, and trans-β-methylstyrene were observed to
be the main products. There were still 3 unidentified by-products in the GC chromatograms, even though all the potential products shown in Figure 2-2 had been tested. The carbon loss could be up to 20% at high conversion. Since, this study is focused on the stability of the Au-Pd/TiO₂ catalyst, the conversion of cinnamyl alcohol and the selectivity to cinnamaldehyde were the main criteria to evaluate catalyst performance.

From Figure 3-3, the conversion of cinnamyl alcohol was observed to be rather low (9%) at 80°C, although it was stable within the investigated duration. The selectivity to cinnamaldehyde fluctuated around 54%. 3-Phenyl-1-propanol was also detected with a selectivity of 6% while no trans-β-methylstyrene was produced. When the reaction was performed at 100 ºC, the conversion initially started from a high value (41%) and gradually decreased to 19% after 7 h. The selectivity to 3-phenyl-1-propanol also presented a similar trend, dropping from 20% to 7% within 3 h and then remaining stable. Since 3-phenyl-1-propanol is formed through the hydrogenation of cinnamyl alcohol, this might indicate that the hydrogenation reactivity of the catalyst was suppressed as reaction progressed, which caused the decrease in the total conversion. The selectivity to cinnamaldehyde increased from 54% at 80 ºC to 64% at 100 ºC. trans-β-Methylstyrene was detected only at low concentrations (both < 3%). When the reaction was conducted at 120 ºC, the initial conversion was further enhanced (58%), but it still decreased to 33% at 7 h. The selectivity to cinnamaldehyde was about 3% lower than that at 100 ºC and similar behaviour was also observed for the selectivities to 3-phenyl-1-propanol and trans-β-methylstyrene. The initial cinnamaldehyde yield was 35%, which corresponded to 2.8 gCD/(gcat·h) cinnamaldehyde produced.
Figure 3-3 Oxidation of cinnamyl alcohol as a function of reaction time at different reaction temperatures. a) 80 °C; b) 100 °C; c) 120 °C. Reaction conditions: Au-Pd/TiO₂ catalyst, 10 mg; oxygen flow, 2.0 mL/min at STP; 0.5 M cinnamyl alcohol in toluene, 20 μL/min; reaction pressure, 4 bar. X, conversion of cinnamyl alcohol; S_CD, selectivity to cinnamaldehyde; S_PP, selectivity to 3-phenyl-1-propanol; S_TR, selectivity to trans-β-methylstyrene; dotted line, average of the selectivity to cinnamaldehyde.
3.3.3 Effect of Catalyst Reduction on Oxidation of Cinnamyl Alcohol

The active phase for cinnamyl alcohol oxidation is still under debate, but the pre-reduction of Pd-based catalysts with alcohol or hydrogen before reaction has been reported to enhance the reaction rate.\textsuperscript{43} In an attempt to address the issue of catalyst deactivation, Au-Pd/TiO\textsubscript{2} was first pre-reduced \textit{in-situ} at the reaction temperature (100 °C) with 0.5 M cinnamyl alcohol in toluene for 0.5 h, followed by introducing oxygen to start the reaction. The reaction results are shown in Figure 3-4a. The initial conversion of cinnamyl alcohol was higher as compared to the non-reduced catalysts (Figure 3-3b), but the activity drop to a similar level after 6 h. To ensure that the catalyst had been satisfactorily reduced, the fresh catalyst was also pre-reduced with hydrogen at 100 °C in another experiment and then purged with helium before reaction. From Figure 3-4b, the pre-reduced catalyst with hydrogen showed similar results to the pre-reduced catalyst with cinnamyl alcohol. The hydrogen reduction process was carried out again and subsequently the reaction was continued (shown in Figure 3-4b). Only ~ 5% higher conversion was observed, which could be caused by the experimental error or a small part of the reversible deactivation.
Figure 3-4 Oxidation of cinnamyl alcohol as a function of reaction time under different pre-reduction conditions. Catalyst pre-reduced with a) 0.5M cinnamyl alcohol; b) 100% hydrogen (re-reduced with hydrogen during the reaction). Reaction conditions: Au-Pd/TiO₂ catalyst, 10 mg; oxygen flow, 2.0 mL/min at STP; 0.5 M cinnamyl alcohol in toluene, 20 μL/min; reaction temperature, 100 °C; reaction pressure, 4 bar. X, conversion of cinnamyl alcohol; S_CD, selectivity to cinnamaldehyde; S_PP, selectivity to 3-phenyl-1-propanol; S_TR, selectivity to trans-β-methylstyrene; dotted line, average of the selectivity to cinnamaldehyde.
Figure 3-5 XPS of Pd 3d (left) and C 1s (right). Catalysts without any pre-reduction a) fresh and b) used catalysts after 7 h reaction at 100 °C with 100% oxygen. Catalysts after pre-reduction with hydrogen c) fresh and d) used catalysts after 7 h reaction at 100 °C and with 100% oxygen.

The palladium oxidation state of fresh and spent catalysts was investigated using XPS. From Figure 3-5a & b, it can be observed that PdO was the main phase in the fresh catalyst without any pre-reduction while more Pd existed in the used catalyst. This shows that PdO could be reduced by cinnamyl alcohol during the reaction (corresponding reaction data are in Figure 3-3b), due to the relatively high reaction temperature. After pre-reduction with hydrogen, most of the PdO was converted into Pd (Figure 3-5c). Notably, Pd was still the main phase after the reaction. At first glance, it seems that these results agree with Lee and co-authors’ contention that Pd/C and Pd/SBA-15 catalyst deactivation was caused by the reduction of PdO to Pd. If so, the pre-reduced catalyst should present lower reactivity from the beginning, given that most of the PdO had been reduced before reaction. Actually, the opposite is observed, i.e. the pre-reduced catalyst shows higher initial conversion than the non-reduced (corresponding reaction data are in Figures 3-3b and 3-4). So, the reduction of PdO to Pd does not seem to be the reason for the deactivation in this case. As for adsorbed hydrocarbons, they were difficult to be identified on the catalyst surface through C1s scanning (shown in Figure 3-5), due to adventitious carbon contamination.

To further explore the cause of catalyst deactivation, the effluent was collected from the experiment in Figure 3-3b over 7 h and then analysed by ICP to check for metal leaching. The analysis revealed no Au but 2.4 ppm of Pd from catalyst leaching, which corresponded to 24% of total Pd loaded on the fresh catalyst. Thus, catalyst deactivation could be correlated to metal leaching. Indeed, metal leaching is often reported to be responsible for catalyst deactivation in oxidation of alcohols, though the cause of metal leaching is rarely specified. In palladium-catalysed
coupling reactions, metal leaching has also been detected during the oxidative addition and accordingly a (quasi)homogeneous mechanism has been suggested. Alloying of Pd with Au has been observed to greatly limit the Pd leaching. A “release and catch” catalytic system, which combines the benefits of homogeneous and heterogeneous catalysis, has been proposed as a potential solution.

3.3.4 Effect of Oxygen Concentration on Oxidation of Cinnamyl Alcohol

For selective oxidation reactions such as that of cinnamyl alcohol, the reaction conditions have been suggested to maintain the Pd constituent in a metallic state with an optimal availability of oxygen to avoid catalyst poisoning by alcohol degradation or over-oxidation of the metal. In batch reactors, Baiker and co-workers reduced the stirring speed and the gas flow rate to make oxygen transport from the gas phase to the Pd surface the rate-limiting step. In the packed bed capillary microreactor, different oxygen concentrations were used to investigate the role of oxygen in the cinnamyl alcohol oxidation. This was achieved using air or nitrogen, while keeping the same gas-to-liquid ratio.

From Figure 3-6a, it can be observed that the Au-Pd/TiO₂ catalyst still showed deactivation in terms of conversion, when using air as the oxidant gas (21 vol% oxygen). The average selectivity to cinnamaldehyde (59%) was lower than that with pure oxygen. Furthermore, relatively high and stable selectivity to 3-phenyl-1-propanol (12%) was maintained during the reaction. This indicates the hydrogenation of cinnamyl alcohol was favoured when less oxygen was present.
Figure 3-6 Oxidation of cinnamyl alcohol as a function of reaction time under different oxygen concentrations. a) 21 vol%; b) 0 vol%. Reaction conditions: Au-Pd/TiO₂ catalyst, 10 mg; gas flow, 2.0 mL/min at STP; 0.5 M cinnamyl alcohol in toluene, 20 μL/min; reaction temperature, 100 °C; reaction pressure, 4 bar. X, conversion of cinnamyl alcohol; S_{CD}, selectivity to cinnamaldehyde; S_{PP}, selectivity to 3-phenyl-1-propanol; S_{TR}, selectivity to trans-β-methylstyrene; dotted line, average of the selectivity to cinnamaldehyde.

Since the catalyst still deactivated even at low oxygen concentration, the reaction was carried out under pure nitrogen conditions (oxygen-free conditions). Surprisingly, a relatively high steady-state conversion of cinnamyl alcohol was obtained with average selectivity to cinnamaldehyde of 44% (Figure 3-6b). An
upward trend was observed for the selectivity to trans-β-methylstyrene (from 16% to 25%), while the selectivity to 3-phenyl-1-propanol dropped from 22% to 15%.

These results are consistent with the dehydrogenation mechanism of alcohol oxidation. The reaction on the metal surface should involve a first dehydrogenation of alcohols to produce adsorbed metal alkoxide and metal hydride.\(^4\)\(^{-22}\) The metal alkoxide is subsequently dehydrogenated into the corresponding aldehyde through β-hydride elimination. Last, the metal hydride reacts with dissociatively adsorbed oxygen to regenerate the metal surface. In this case, hydrogen is also required during the formation of 3-phenyl-1-propanol and trans-β-methylstyrene (shown in Figure 2-2). Interestingly, the average selectivity to cinnamaldehyde (44%) is almost equal to the sum of 3-phenyl-1-propanol and trans-β-methylstyrene (40%). This indicates the reactant itself acts as hydrogen acceptor in the absence of oxygen. This phenomenon was also observed in the oxygen-free oxidation of benzyl alcohol, where equal amounts of toluene and benzaldehyde were produced.\(^{35, 126}\)

The effluent collected from the reaction under oxygen-free conditions over 7 h (Figure 3-6b) was analysed with ICP to check for catalyst leaching. 1.6 ppm of Pd was found, which corresponded to 16% Pd loss. This explains the slight drop in the cinnamyl alcohol conversion under nitrogen conditions from 70% to 60% (as shown in Figure 3-6b). Pd leaching might be caused by the formation of complexes through the interaction of the products with the catalyst. The Pd loss is lower than that under oxygen-rich conditions (24%); this may be due to accelerated leaching when the metal is oxidised.\(^{131}\)

To better illustrate the effect of oxygen concentration at the reactor inlet, these results are summarized in Figure 3-7. Clearly, there is a strong correlation between
the oxygen concentration and the reactivity in terms of conversion and selectivity. With decreasing oxygen concentration, the conversion was enhanced and the deactivation rate decreased. On the other hand, comparatively low selectivity to cinnamaldehyde was obtained under low oxygen concentration. These contribute to the comparable yield to cinnamaldehyde at some points (shown in Figure 3-7c). On the basis of initial dehydrogenation of cinnamyl alcohol followed by secondary hydrogenation to form 3-phenyl-1-propanol and trans-β-methylstyrene, the cinnamaldehyde yield should be independent of oxygen concentration in the feed. Using the data of the initial conversion (to minimise the effects of catalyst deactivation) and the average selectivity, the aldehyde yields are 26% (100% O₂), 33% (21% O₂) and 31% (0% O₂), i.e. roughly independent of oxygen and confirming the correctness of the assumptions. Hardacre et al.¹³⁰ reported a higher alcohol conversion and lower selectivity to cinnamaldehyde in the absence of oxygen compared with experiments in the presence of oxygen when Pd/Al₂O₃ catalyst was used in a rotating disc reactor at 80 °C. Baiker and co-workers observed low conversion on Pd/Al₂O₃ catalyst under argon atmosphere at 40 °C, while in air, higher conversion was obtained due to oxidative removal of CO and other degradation products. However, they argue that an optimal level of oxygen is required to keep the catalyst surface clean but not overoxidise the catalyst.⁴³. With our catalyst, we believe that the presence of Au can alter the electronic and geometrical properties of the synthesized particles.²⁶,⁶⁰

Since the rate of catalytic dehydrogenation appears to be independent of oxygen concentration but different percentage decrease in cinnamyl alcohol conversion was observed under different oxygen concentrations, yet with similar percentage loss of
Pd, poisoning of the surface by adsorbed oxygen might be also responsible for the catalyst deactivation.\textsuperscript{6,22,132} In the absence of oxygen, the metal hydride is consumed in rapidly reducing cinnamyl alcohol to either 3-phenyl-1-propanol or \textit{trans}-\(\beta\)-methylstyrene. When oxygen is present, adsorbed oxygen on the metal surface competes with adsorbed cinnamyl alcohol for metal hydride. Therefore, the selectivities to 3-phenyl-1-propanol and \textit{trans}-\(\beta\)-methylstyrene go down and so does the cinnamyl alcohol conversion. Activity was only partially recovered when re-reducing the catalyst with hydrogen (Figure 3-4b). Hence, the irreversible deactivation might be caused by Pd leaching (possibly accelerated in the presence of oxygen) and strong adsorption of poisoning species formed under oxygen conditions.\textsuperscript{43,54}

To further explore the effect of oxygen on catalyst reactivity and deactivation, oxygen/nitrogen switching experiment was performed with the same catalyst packing. From Figure 3-8, deactivation as expected was observed within 3 h under pure oxygen conditions. Then, a distinct rise of conversion was obtained when changing the gas from oxygen to nitrogen. The selectivity to cinnamaldehyde also decreased simultaneously, together with the increase of selectivities to 3-phenyl-1-propanol and \textit{trans}-\(\beta\)-methylstyrene. The trend of conversion and selectivities under pure nitrogen conditions seems to be slightly different from that in Figure 3-6b, which might be caused by changes of surface properties or the presence of poisoning species from the previous reaction stage with oxygen. Nevertheless, these results clearly indicate an inhibiting effect adsorbed oxygen on the cinnamyl alcohol conversion.
Figure 3-7 Comparison of catalyst performance and deactivation under different oxygen concentration during reaction. a) conversion of cinnamyl alcohol; b) selectivity to cinnamaldehyde; c) yield to cinnamaldehyde. Reaction conditions: Au-Pd/TiO$_2$ catalyst, 10 mg; gas flow, 2.0 mL/min at STP; 0.5 M cinnamyl alcohol in toluene, 20 μL/min; reaction temperature, 100 °C; reaction pressure, 4 bar; dotted line, average of the selectivity to cinnamaldehyde.

Figure 3-8 Oxidation of cinnamyl alcohol as a function of reaction time under different gas conditions. Reaction conditions: Au-Pd/TiO2 catalyst, 10 mg; gas flow, 2.0 mL/min at STP; 0.5 M cinnamyl alcohol in toluene, 20 μL/min; reaction temperature, 100 ºC; reaction pressure, 4 bar. X, conversion of cinnamyl alcohol; SCD, selectivity to cinnamaldehyde; Spp, selectivity to 3-phenyl-1-propanol; STR, selectivity to trans-β-methylstyrene.

Switching experiments between oxidation of cinnamyl alcohol and benzyl alcohol were also carried out with the same catalyst packing, since different catalyst stabilities were observed in these two reactions. As shown in Figure 3-9, catalyst deactivation was first observed in cinnamyl alcohol oxidation. When the reaction was changed to benzyl alcohol oxidation, the catalyst still presented good stability in benzyl alcohol oxidation over the following 20 h. When it was changed back to cinnamyl alcohol oxidation, the trend of catalyst deactivation continued. This suggests that the catalyst deactivation in cinnamyl alcohol oxidation due to metal leaching and presence of oxygen does not apply to benzyl alcohol oxidation, and this may be due to the different catalytic sites for the two reactants, different reaction mechanism as well as the different products.

Figure 3-9 Oxidation of cinnamyl alcohol and benzyl alcohol as a function of reaction time. Reaction conditions: Au-Pd/TiO2 catalyst, 10 mg; oxygen flow, 2.0 mL/min at STP; 0.5 M cinnamyl alcohol in toluene, 20 μL/min; neat benzyl alcohol, 10 μL/min; reaction temperature, 100 ºC; reaction pressure, 4 bar. CA, cinnamyl alcohol; BnOH, benzyl alcohol; X, conversion of cinnamyl alcohol or benzyl alcohol; S_{CD}, selectivity to cinnamaldehyde; S_{PP}, selectivity to 3-phenyl-1-propanol; S_{TR}, selectivity to trans-β-methylstyrene; S_{B}, selectivity to benzaldehyde; S_{T}, selectivity to toluene.

3.4 Conclusions

A bimetallic Au-Pd/TiO2 catalyst was studied for oxidation of alcohols in a packed bed microreactor. The catalyst showed stable conversion and selectivity for the oxidation of benzyl alcohol to benzaldehyde. For cinnamyl alcohol oxidation, catalyst deactivation was observed which was more severe at high reaction temperature and high oxygen concentration. Catalyst pre-reduction with cinnamyl alcohol or hydrogen before reaction improved initial activity but did not affect catalyst deactivation. Higher oxygen concentration did not affect the rate of cinnamaldehyde formation but led to lower initial alcohol conversion by suppressing the formation of the by-products, 3-phenyl-1-propanol and trans-β-methylstyrene,
leading to higher cinnamaldehyde selectivity. Pd was detected in the reactor effluent during cinnamyl alcohol oxidation. Thus, Pd leaching and the presence of oxygen seem to be causes for catalyst deactivation. The higher deactivation rates and amount of Pd leached at higher oxygen concentration, along with partial recovery of activity after hydrogen treatment, indicate that the role of oxygen is complex, affecting deactivation in several ways (possibly including competing with alcohol on the catalyst, removing by-products from surface, accelerating Pd leaching). Switching substrates between benzyl alcohol and cinnamyl alcohol showed deactivation during cinnamyl alcohol oxidation but steady performance during benzyl alcohol oxidation, demonstrating different deactivation characteristics for the two reactions.
CHAPTER 4. Continuous Heterogeneously Catalysed Oxidation of Benzyl Alcohol Using a Tube-in-Tube Membrane Microreactor* 

A Teflon AF-2400 tube-in-tube membrane microreactor is investigated for the continuous, solvent-free, catalytic oxidation of benzyl alcohol with oxygen. The semi-permeable tube was packed with 1 wt% Au-Pd/TiO$_2$ catalyst particles and placed inside a PTFE tube to provide an annular region which was pressurised with pure oxygen. This design allowed continuous penetration of oxygen through the inner tube during the reaction, resulting in higher oxygen concentration in the catalyst bed and significantly improved conversion compared to a reactor operating with an oxygen pre-saturated feed. The amount of oxygen available for reaction in the tube-in-tube microreactor was 2 orders of magnitude higher than that in the non-permeable reactor with oxygen pre-saturated feed. The semi-permeable tube reactor performance in terms of both conversion and selectivity was enhanced by increasing the gas pressure, the catalyst contact time and by dilution of the catalyst.

CHAPTER 4. Continuous Heterogeneously Catalysed Oxidation of Benzyl Alcohol Using
a Tube-in-Tube Membrane Microreactor

4.1 Introduction

Heterogeneous catalytic gas-liquid reactions are an important category in synthetic chemistry, with great significance for both laboratory research and industrial applications. Conventionally such reactions are performed in a batch reactor, and the phases are brought in contact by stirring mechanically. However, this results in a poorly defined interfacial contact area and renders scale-up more complex.\textsuperscript{16, 62} Additionally, the immediate contact of the phases also raises safety concerns for hazardous reactions like oxidation reactions. Over the past decade, microreactor technology has emerged as an efficient tool for chemical synthesis.\textsuperscript{133} The intrinsic advantages associated with such reactors are the high efficiency of heat and mass transfer, because of the high surface-area-to-volume ratio, as well as the high-resolution reaction time control.\textsuperscript{16, 17, 62} The small physical scale can also significantly enhance safety.\textsuperscript{11, 84} Several types of microreactors have been developed for gas-liquid-solid reactions,\textsuperscript{134} including the coated-wall microreactor,\textsuperscript{84} the packed-bed microreactor,\textsuperscript{35} and the “slurry Taylor” flow microreactor\textsuperscript{92}.

Recently, membrane microreactors have drawn growing interest, as they combine the benefits of both the membrane reactor and the microreactor.\textsuperscript{103, 104} Gas and liquid phases can flow separately with a well-defined contacting interface and high mass and heat transfer rates. A tube-in-tube configuration employing a semipermeable Teflon AF-2400 tube as the contact interface for gas and liquid has recently been proposed by Ley’s group.\textsuperscript{116, 135} This material is permeable to a wide range of gases but is impermeable to liquids. This controlled method of gas/liquid contact could enhance safety, especially for hazardous reactions. Generally, for three-phase reactions, the liquid was first saturated with gas in the tube-in-tube configuration and
subsequently reacted in a catalyst packed column.\textsuperscript{123, 124} However, highly pressure-dependent conversion was observed for reactions in such apparatus.\textsuperscript{123} Since gas is not supplied in the downstream reaction section, the amount of gas available for the reaction is limited by its solubility in the reaction mixture.\textsuperscript{119}

Selective oxidation of alcohols is an important reaction in chemical synthesis,\textsuperscript{1, 2} and the aerobic oxidation of benzyl alcohol using molecular oxygen is one of the most intensively studied.\textsuperscript{11, 28, 71, 84} A few studies investigated this reaction under flow conditions using a gold-based catalyst. Wang \textit{et al}.\textsuperscript{84} prepared a gold-immobilized capillary column reactor through cross-linking of copolymer for the oxidation of various alcohols. In our previous studies,\textsuperscript{35, 136} silicon-glass micropacked-bed reactors were used for aerobic oxidation of benzyl alcohol, with a Au-Pd/TiO\textsubscript{2} catalyst. Conversions of benzyl alcohol obtained from conventional stirred batch reactor and micropacked reactor were found to be comparable. In this chapter, oxidation of benzyl alcohol was investigated on Au-Pd/TiO\textsubscript{2} catalyst, which was directly packed in a Teflon AF-2400 tube-in-tube microreactor. Thus, the gas transport process and the reaction section were integrated, and gas was supplied as the reaction progressed. Operating parameters were studied, with the aim to understand reactor behaviour and improve its performance.

\textbf{4.2 Experimental}

\textbf{4.2.1 Catalyst Preparation*}

Au-Pd/TiO\textsubscript{2} catalyst (1 wt\%) with Au-to-Pd weight ratio equivalent to 1:19 was used and was prepared by a modified impregnation method.\textsuperscript{32, 33} The gold precursor (H\textsubscript{3}AuCl\textsubscript{4}\cdot3H\textsubscript{2}O; Sigma-Aldrich) was dissolved in deionized water to form a solution

\* \textit{Catalyst preparation was carried out by Dr Moataz Morad, Cardiff University.}
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with a gold concentration of 9.8 mg/mL, and the palladium precursor (PdCl₂, Sigma-Aldrich) was dissolved in a 0.58 M aqueous HCl solution (conc. HCl, diluted using the requisite amount of deionized water) with vigorous stirring to form a solution with a Pd concentration of 6 mg/mL. Then, the requisite amount of each metal precursor solution was charged into a 50 mL round-bottom flask fitted with a magnetic stirrer bar. The volume of the mixture was adjusted using deionized water to a total volume of 15 mL. The flask was immersed into an oil bath placed on a magnetic stirrer hot plate. The solution was stirred vigorously at 1000 rpm and the temperature of the oil bath was raised from 27 to 60 °C over a period of 10 min. At 60 °C, 1.98 g of the TiO₂ support (Degussa Evonik P25, Germany) was added slowly over a period of 8 to 10 min with constant stirring. After the completion of the addition of the support material, the slurry was stirred at 60 °C for an additional 15 min. Following this, the temperature of the oil bath was raised to 95 °C, and the slurry was stirred at that temperature for a further 16 h until all the water evaporated leaving a dry solid. Subsequently the resultant solid powder was transferred into a mortar and pestle and was ground thoroughly to form a uniform mixture. A 400 mg portion of the uncalcined sample was transferred and spread out over a glass calcination boat (13 cm in length). This boat was then placed inside a calcination furnace fitted with an inlet and outlet valve. The temperature of the furnace was raised from 30 to 400 °C at a heating rate of 10 °C/min and then kept at 400 °C for 4 h under a steady flow of 5 % H₂ in Ar. After the furnace was cooled, the resultant solid powder was pelletized, crushed and sieved to the desired particle size range.
4.2.2 Tube-in-Tube Microreactor Set-up

The schematic representation of the tube-in-tube microreactor used in this work is shown in Figure 4-1. The reactor comprises an inner Teflon AF-2400 tube (Biogeneral) and an outer PTFE tube (Agilent). The inner diameter (ID) of the Teflon AF-2400 tube is 0.8 mm, and the outer diameter (OD) is 1.0 mm. The ID and OD of the PTFE tube are 2.4 and 3.2 mm, respectively. For the measurement of oxygen permeation, a 100 cm-length tube was used without any packing. For the catalytic oxidation reaction, the total length of the reactor was 30 cm. Packing the inner tube was achieved by placing in the tube outlet a piece of silica wool which was supported by a 5 cm-length stainless steel tube. 20 mg Au-Pd/TiO₂ catalyst particles (63-75 μm) were sucked into the inner tube by a vacuum pump. The

![Figure 4-1 Schematic of the Teflon AF-2400 tube-in-tube microreactor set-up.](image)
amount of catalyst used in all experiments was 20 mg. The bed length was
maintained at ca. 3.0 cm. The catalyst was also mixed with silica beads (63-75 μm)
before packing, to investigate the effect of catalyst dilution. The inner semi-
permeable tube and the outer PTFE tube were assembled together by a union and a
T-piece to make up the tube-in-tube microreactor. The outlet of the inner tube was
connected to a 7.9 bar back pressure regulator (BPR) (Upchurch). All pressures
reported in this study are absolute. The pressure of the gas and liquid phases was
monitored by pressure sensors placed upstream of the reactor (Zaiput). The effluent
was collected in a small vial which was placed in an ice-water bath. The reactor was
immersed in a stirred oil bath, whose temperature was controlled by a hotplate fitted
with a thermocouple (Stuart US152). The reaction was carried out at 120 ºC. Oxygen
was fed in the annulus between the inner and outer tube, and the pressure was
maintained by a gas pressure regulator (Swagelok). When the temperature reached
the desired value, liquid benzyl alcohol (99.0%, Sigma-Aldrich) was supplied into
the tube-in-tube microreactor by a HPLC pump (Knauer P2.1S). The temperature of
the fluid reached the required reaction temperature before entering the catalyst bed,
as observed by temperature measurements inside the inner tube. The collected
sample was quantitatively analysed by a gas chromatograph (Agilent 6890) fitted
with a HP-INNOWax capillary column and a FID detector. Benzyl alcohol
conversion (X) and selectivity (S_i) of each product i were calculated according to the
following equations:

\[ X = \frac{c_{BnOH, in} - c_{BnOH, out}}{c_{BnOH, in}} \times 100\% \]  \hspace{1cm} (4-1)

\[ S_i = \frac{c_i \cdot \nu_i}{c_{BnOH, in} \cdot X} \times 100\% \]  \hspace{1cm} (4-2)
where $c_{BnOH,in}$ and $c_{BnOH,out}$ are the concentration of benzyl alcohol at the inlet and outlet of the reactor, respectively. $v_i$ is the number of moles of benzyl alcohol giving rise to 1 mol of product $i$.

Catalyst contact time (CCT) was used to characterise the reaction time of benzyl alcohol, and was defined as:

$$CCT = \frac{m_{cat}}{v \rho_{BnOH}} \times 100\%$$

(4-3)

where $m_{cat}$ is the mass of catalyst, $v$ is the volume flow rate of benzyl alcohol at the inlet, $\rho_{BnOH}$ is the density of benzyl alcohol.

For each experiment, at least three samples were collected and the results were averaged. The relative error was less than ±1.0%. Reproducibility of the experiments and the stability of the catalyst were checked every day by a standard run (120 ºC; catalyst contact time, 115 g$_{cat}$·s/g$_{alcohol}$; gas pressure, 6 bar; liquid flow rate, 10 μL/min; liquid pressure, 7.9 bar) and the differences were less than ±2 %.

Figure 4-2 Packed bed PTFE reactors with different gas-liquid feed. (a) benzyl alcohol exposed to air before use, (b) benzyl alcohol presaturated with oxygen by a tube-in-tube contactor, (c) mixture of oxygen and benzyl alcohol.
CHAPTER 4. Continuous Heterogeneously Catalysed Oxidation of Benzyl Alcohol Using a Tube-in-Tube Membrane Microreactor

For comparison, a PTFE tube with ID and length the same as that of the Teflon AF-2400 tube was also packed with the same catalyst and tested. Three different feeding modes (shown in Figure 4-2) were investigated for the PTFE reactor: benzyl alcohol (a) exposed to air before use, (b) presaturated with oxygen at given gas pressure by an unpacked tube-in-tube contactor, and (c) mixed with oxygen. The liquid pressure for the liquid-only-feed design (Figure 4-2a,b) was kept at 7.9 bar using a fixed BPR (Upchurch, USA). An adjustable BPR (Zaiput, USA) was used for the mixed gas/liquid feed design (Figure 4-2c) to maintain the gas/liquid mixture pressure at 6 bar.

4.3 Results and Discussion

4.3.1 Oxygen Permeation

The amount of oxygen dissolved in benzyl alcohol is an important factor for the evaluation of the tube-in-tube microreactor.\(^{119}\) When the benzyl alcohol leaves the BPR, the dissolved oxygen escapes from the liquid, because of the pressure difference between the reactor and the outlet. Assuming the residual gas in the fluid at the outlet is the same as that in the inlet (benzyl alcohol presaturated with oxygen at 1 bar and room temperature), the volume of oxygen released per unit volume of benzyl alcohol equates approximately to the oxygen permeated while the liquid is passing through the reactor. This was measured using the traditional burette method.\(^{123}\) During the measurement, the tube-in-tube microreactor was kept at 120 °C, and the liquid outlet was connected to a burette filled with benzyl alcohol (presaturated with oxygen at 1 bar and room temperature). The duration for collecting a fixed volume of oxygen in the burette was recorded for different liquid flow rates and absolute gas pressures in the annulus. Thus, the outgassing rate of
oxygen could be calculated. The average liquid residence time was calculated by the tube length divided by the superficial velocity of the liquid in the inner tube.

![Graph showing outgassing rate of oxygen as a function of residence time and gas pressure in the annulus. Liquid pressure, 7.9 bar; temperature of the tube-in-tube microreactor, 120 ºC.]

Figure 4-3 Outgassing rate of oxygen as a function of residence time and gas pressure in the annulus. Liquid pressure, 7.9 bar; temperature of the tube-in-tube microreactor, 120 ºC.

From Figure 4-3 thermodynamic data, it can be noted that initially the outgassing rate increased gradually with residence time at each set gas pressure, and stabilised after 50 s, indicating saturation of benzyl alcohol with oxygen. O’Brien et al.\textsuperscript{123} reported a similar behaviour for a hydrogen/dichloromethane system. The final outgassing rates at different gas pressure varied almost linearly with the gas pressure difference (between annulus and burette). The conditions of oxygen-benzyl alcohol equilibrium were also determined with different equations of state (Peng-Robinson, UNIQUAC, NRTL, \textit{etc.}) in ASPEN PLUS,\textsuperscript{*} and the simulated outgassing rates were indicated to be much higher than the measured rates at stable stage (shown in Appendix A). This was further discussed in the following chapter. Nevertheless, these measurements show that the permeation of oxygen through the semi-permeable

\textsuperscript{*}ASPEN simulation in this thesis was carried out by Baldassarre Venezia.
tube can be altered by changing the liquid residence time and gas pressure.

### 4.3.2 Performance of Different Reactors

Reaction experiments of benzyl alcohol were first performed in the packed PTFE tube at 120 °C. Using the substrate without presaturation or mixing with oxygen (Figure 4-2a), the conversion was very low, just 1.5% (Table 4-1) in the packed PTFE reactor. Equimolar amounts of benzaldehyde and toluene were formed, because of the disproportionation of benzyl alcohol.\textsuperscript{34, 35} Relatively high selectivity to dibenzyl ether was observed, which was formed by the intermolecular dehydration of benzyl alcohol.\textsuperscript{34} These results indicate that practically no oxidation reaction was taking place in the absence of oxygen. To test the packed PTFE reactor under oxidative condition, benzyl alcohol was first saturated with oxygen by a tube-in-tube contactor (100 cm long, gas pressure 6 bar, liquid flow rate 10 μL/min, Figure 4-2b). The residence time in the tube-in-tube contactor was about 3000 s, which was long enough for the alcohol to become saturated with oxygen. Under these conditions, no obvious improvement in conversion was obtained. The main difference between the two feed approaches was the higher selectivity to benzaldehyde compared to toluene. This was due to the additional oxidation of benzyl alcohol, besides the disproportionation reaction, because of the small amount of dissolved oxygen present in the liquid substrate.
### Table 4-1

Conversions and selectivities (%) of oxidation of benzyl alcohol in different reactors. Reaction conditions: 120 °C; catalyst contact time, 115 g<sub>cat</sub>/s/g<sub>alcohol</sub>; liquid flow rate, 10 μL/min; gas pressure, 6 bar; catalyst dilution factor, 1.

<table>
<thead>
<tr>
<th>Reaction conditions</th>
<th>X</th>
<th>X_O</th>
<th>S&lt;sub&gt;benzaldehyde&lt;/sub&gt;</th>
<th>S&lt;sub&gt;toluene&lt;/sub&gt;</th>
<th>S&lt;sub&gt;benzene&lt;/sub&gt;</th>
<th>S&lt;sub&gt;dibenzyl ether&lt;/sub&gt;</th>
<th>S&lt;sub&gt;benzoic acid&lt;/sub&gt;</th>
<th>S&lt;sub&gt;benzyl benzoate&lt;/sub&gt;</th>
</tr>
</thead>
<tbody>
<tr>
<td>a</td>
<td>1.5</td>
<td>0.0</td>
<td>41</td>
<td>40</td>
<td>0</td>
<td>18</td>
<td>0.3</td>
<td>1.5</td>
</tr>
<tr>
<td>b</td>
<td>1.9</td>
<td>0.1</td>
<td>45</td>
<td>41</td>
<td>0</td>
<td>14</td>
<td>0</td>
<td>0.5</td>
</tr>
<tr>
<td>c</td>
<td>37</td>
<td>11</td>
<td>63</td>
<td>34</td>
<td>0.7</td>
<td>1.0</td>
<td>0.2</td>
<td>0.7</td>
</tr>
<tr>
<td>d</td>
<td>78</td>
<td>46</td>
<td>75</td>
<td>16</td>
<td>0.1</td>
<td>0.4</td>
<td>1.0</td>
<td>7.3</td>
</tr>
</tbody>
</table>

a: Reaction in packed bed PTFE reactor with benzyl alcohol exposed to air before use, liquid pressure 7.9 bar;
b: Reaction in packed bed PTFE reactor with benzyl alcohol presaturated with oxygen at 6 bar gas pressure, liquid pressure 7.9 bar;
c: Reaction in packed bed tube-in-tube microreactor, liquid pressure 7.9 bar;
d: Reaction in packed bed PTFE reactor with mixture of oxygen and benzyl alcohol feed, liquid pressure 6.0 bar.
These results confirm the two main reactions of the overall oxidation of benzyl alcohol, according to previous studies. Benzaldehyde, the target product, is formed by both the oxidation and disproportionation of benzyl alcohol, while another main product, toluene, is obtained only from the disproportionation. We can quantify the amount of benzaldehyde produced by the oxidation and further calculate the amount of benzyl alcohol consumed by the oxidation reaction ($X_O$):

$$X_O = X \cdot (S_{benzaldehyde} - S_{toluene})$$  \hspace{1cm} (4-4)

This conversion in Table 4-1b was just 0.1%. Assuming that all the oxygen dissolved in the liquid flow (~0.5 mL/mL_alcohol estimated from Figure 4-3) was consumed through oxidation reaction, the theoretical maximum $X_O$ would be 0.4%.

Next, the reactor with the catalyst packed in the semi-permeable tube-in-tube microreactor was evaluated (Figure 4-1). From Table 4-1c, it can be seen that the conversion of benzyl alcohol increased dramatically compared to the other two approaches at the same reaction temperature and gas pressure. The selectivity to benzaldehyde was about 30% higher than that of toluene. This improvement in the performance of the tube-in-tube microreactor was due to the increased amount of oxygen participating in the reaction, based on the corresponding $X_O$. The amount of consumed oxygen calculated from $X_O$ is 12.4 mL/mL_alcohol, which is ca. 150 times higher than 0.08 mL/mL_alcohol calculated for oxygen-presaturated benzyl alcohol. Comparing this value (12.4 mL/mL_alcohol) with the outgassing rate in Figure 4-3 (~0.5 mL/mL_alcohol) shows that oxygen permeation during the reaction was enhanced, because an oxygen concentration gradient was sustained across the semi-permeable tube wall. This makes the packed tube-in-tube microreactor more efficient for three-phase reactions compared with non-permeable PTFE reactor configurations with
CHAPTER 4. Continuous Heterogeneously Catalysed Oxidation of Benzyl Alcohol Using a Tube-in-Tube Membrane Microreactor

The amount of benzyl alcohol consumed by the disproportionation reaction can be estimated by subtracting \(X_O\) from \(X\). It was 1.8% for the conditions of Table 4-1b and 27% for those of Table 4-1c. This indicates that adding oxygen can accelerate the disproportionation reaction, which agrees well with previous studies that suggest two distinct pathways to disproportionation.\(^3\) Selectivities to other products are also listed in Table 4-1, and the values were all lower than 2%.

To further explore the effect of oxygen concentration on the reaction performance, a mixture of oxygen and benzyl alcohol was fed into the packed PTFE reactor (Figure 4-2c). The gas-to-liquid volume ratio (at standard temperature and pressure, 0 °C, 1.0 bar) was 100, and the outlet pressure was kept at 6 bar. Two-phase slug flow was observed upstream of the packed bed. It can be seen from Table 4-1d that the conversion was 2-fold higher than that in the tube-in-tube microreactor, with 12% higher selectivity to benzaldehyde. The difference between the two approaches clearly indicates the deficiency of oxygen for the packed tube-in-tube microreactor for the conditions of Table 4-1c. However, the packed tube-in-tube microreactor is intrinsically safer as gaseous oxygen does not come in contact with the organic mixture.

4.3.3 Effect of Gas Pressure in Packed Tube-in-Tube Microreactor

To alter the amount of oxygen available for the reaction, the oxidation of benzyl alcohol was carried out with the gas pressure in the annulus changing from 3 to 7 bar. The liquid pressure was maintained at 7.9 bar by the BPR, and the catalyst contact time and reaction temperature were kept the same. The conversion of benzyl alcohol and selectivities to benzaldehyde (\(S_B\)) and toluene (\(S_T\)) are shown in Figure 4-4;
other minor products (not shown) were all lower than 2.0%. The total conversion of benzyl alcohol increased gradually with the gas pressure. The highest conversion was 42% at 7 bar gas pressure. The same trend was observed for the selectivity to benzaldehyde, which reached 65% at 7 bar. As for the selectivity to toluene, it showed an opposite dependence on the gas pressure, decreasing from 36% to 32%. Similar behaviour was also observed in a batch and packed bed microreactor. The conversion increased from 79% at 2.1 bar to ca. 95% at 5 bar, and the selectivity to benzaldehyde increases from 58% to 78% in a packed bed microreactor with oxygen-alcohol mixture feed.

![Figure 4-4](image)

**Figure 4-4** Effect of gas pressure in the annulus on the conversion and selectivities during benzyl alcohol oxidation. Reaction conditions: catalyst contact time, 115 g-cat]/g/alcohol; liquid flow rate, 10 μL/min; reaction temperature, 120 °C; liquid pressure, 7.9 bar; catalyst dilution factor, 1. \(X = \) total benzyl alcohol conversion, \(X_O = \) benzyl alcohol conversion through oxidation reaction; \(S_T = \) selectivity to toluene; \(S_B = \) selectivity to benzaldehyde.

Additionally, \(X_O\) varied approximately linearly with the gas pressure. This directly proves that more oxygen participated in the reaction at higher gas pressure. Furthermore, it is reasonable to deduce that the oxygen concentration in the liquid...
was maintained at low level, and the plot of $X_O$ also quantitatively reflects the amount of oxygen penetrating through the inner tube. Hence, increasing the gas pressure can enhance the oxygen solubility in the liquid and the oxygen concentration gradient between the annulus and the inner tube, and thus results to more benzyl alcohol reacting and more oxygen permeation.

### 4.3.4 Effect of Catalyst Contact Time in Packed Tube-in-Tube Microreactor

The effect of catalyst contact time was investigated by changing the flow rate of benzyl alcohol, keeping the same mass of catalyst. The results are shown in Figure 4-5. The conversion increased from 26% to 47%, with selectivity to benzaldehyde from 59% to 68%, when catalyst contact time increased from 46 to 230 g$_{\text{cat}}$/g$_{\text{alcohol}}$. Correspondingly, the selectivity to toluene decreased from 38% to 29% at the investigated catalyst contact time range. This conversion trend agrees with that in a batch and packed bed microreactor. However, the selectivities to benzaldehyde and toluene in both batch and microreactor, where oxygen was in excesses, were almost the same under different catalyst contact time.$^{35}$ The selectivity differences in the tube-in-tube microreactor were probably caused by the insufficient amount of oxygen in the liquid. It should be noted that the plot of $X_O$ also showed a linear trend. Considering the bed length was the same because of the same amount of catalyst packed, longer catalyst contact time was due to lower flow rate of the liquid, and the average residence time for the liquid through the bed was correspondingly longer. Thus, oxygen had more time to permeate and react because of longer catalyst contact time, and this resulted in an increase in the oxygen consumed by the oxidation reaction, $X_O$. 
CHAPTER 4. Continuous Heterogeneously Catalysed Oxidation of Benzyl Alcohol Using a Tube-in-Tube Membrane Microreactor

**Figure 4-5** Effect of catalyst contact time on the conversion and selectivities during benzyl alcohol oxidation. Reaction conditions: gas pressure in the annulus, 6 bar; reaction temperature, 120 °C; liquid pressure, 7.9 bar; catalyst dilution factor, 1. $X = \text{total benzyl alcohol conversion}$, $X_O = \text{benzyl alcohol conversion through oxidation reaction}$; $S_T = \text{selectivity to toluene}$; $S_B = \text{selectivity to benzaldehyde}$.

**Figure 4-6** Effect of catalyst dilution on the conversion and selectivities during benzyl alcohol oxidation. Reaction conditions: catalyst contact time, 115 g$_{\text{cat}}$/s/g$_{\text{alcohol}}$; liquid flow rate, 10 μL/min; gas pressure in the annulus, 6 bar; liquid pressure, 7.9 bar; reaction temperature, 120 °C. $X = \text{total benzyl alcohol conversion}$; $X_O = \text{benzyl alcohol conversion through oxidation reaction}$; $S_T = \text{selectivity to toluene}$; $S_B = \text{selectivity to benzaldehyde}$.
4.3.5 Effect of Catalyst Dilution in Packed Tube-in-Tube Microreactor

The effect of catalyst dilution was investigated by diluting the same amount of catalyst with different amounts of silica beads. The catalyst dilution factor is defined as the total mass of catalyst bed (catalyst + silica beads) divided by the mass of catalyst. In this case, catalyst contact time was kept constant (inlet benzyl alcohol flow rate was constant), while the bed length (and hence bed residence time) was changed. The actual bed length was about 3.0 cm for every 20 mg of packing. In Figure 4-6, reaction performance is shown, when the catalyst was mixed with the same mass of silica beads (catalyst dilution factor, 2). The conversion (42%) and selectivity to benzaldehyde (68%), as well as $X_O$ (18%) were close to those when catalyst contact time was $230 \text{ g}_{\text{cat}} \cdot \text{s}/\text{g}_{\text{alcohol}}$ (47%, 68%, and 18%). Assuming the porosity of the bed was the same for both conditions, the average residence time for the liquid passing through the bed with the catalyst contact time of $115 \text{ g}_{\text{cat}} \cdot \text{s}/\text{g}_{\text{alcohol}}$ and catalyst dilution factor 2 equalled to that when catalyst contact time was $230 \text{ g}_{\text{cat}} \cdot \text{s}/\text{g}_{\text{alcohol}}$ and catalyst dilution factor was 1. Thus, the time for oxygen permeation was also the same. However, the main difference was that half of the bed in the former case did not contain catalyst. Therefore, similar results under these two conditions further support that oxygen permeation is a key factor controlling the performance of the tube-in-tube microreactor. Further dilution did not change the total conversion of benzyl alcohol very much, while the selectivity to benzaldehyde increased gradually, as well as the amount of benzyl alcohol converted by oxygen. Notably, the selectivity to benzaldehyde at catalyst dilution factor 4 reached 73%, which was close to the premixed alcohol-oxygen conditions (75%, Table 4-1d). This also approached the selectivity to benzaldehyde in a micropacked bed reactor (78%,
6 bar gas pressure; gas-to-liquid volume ratio, 200). Thus, dilution of the catalyst bed allowed increasing the surface area available for oxygen transfer without increasing the oxygen demand by the catalytic bed, resulting in higher conversion and selectivity.

### 4.4 Conclusions

A semi-permeable Teflon AF-2400 tube-in-tube microreactor was developed for the continuous solvent-free aerobic heterogeneous oxidation of benzyl alcohol. By comparison with non-permeable PTFE configurations, the packed tube-in-tube microreactor allowed the continuous supply of oxygen during the reaction and therefore significantly improved both conversion and selectivity. The amount of oxygen that permeated played a vital role in reactor performance. Oxygen permeation was increased by increasing the gas pressure and catalyst contact time. When the catalyst was diluted, the average residence time of liquid in the bed increased, allowing larger oxygen permeation. All these approaches of enhancing oxygen permeation resulted in improved reactor performance. Even though the amount of oxygen available for the reaction was not enough to achieve the same total conversion of benzyl alcohol as in a packed non-permeable PTFE reactor with a mixture of oxygen/benzyl alcohol feed, the selectivity for both reactors was similar. As gas does not come directly into contact with the organic mixture in the packed tube-in-tube microreactor, its intrinsic safety makes this type of reactor promising for wider applications in heterogeneous reactions.
CHAPTER 5. A Novel Approach for Measuring Gas Solubility in Liquids Using a Tube-in-Tube Membrane Contactor

Data on oxygen solubility in organic solvents are essential for catalyst evaluation, reactor design, and process safety of catalytic aerobic oxidations which is an emerging area in green chemistry. A novel approach using a Teflon AF-2400 tube-in-tube membrane contactor was developed for the measurement of gas solubility in organic solvents. The semi-permeable Teflon AF-2400 membrane ensures gas saturation of liquids in continuous-flow at specific pressure and temperature. After liquid decompression, the amount of gas outgassed was measured with a bubble meter and used for solubility calculation. In this study, this approach was demonstrated for the measurement of oxygen solubility in toluene and benzyl alcohol. Validation experiments were initially performed by comparing the obtained oxygen solubility in toluene with literature data. Then, oxygen solubility in benzyl alcohol was measured at pressures up to 10 bar and temperatures 298 - 393 K. Oxygen solubility in benzyl alcohol at 298 K (Henry’s law constant, 3462 bar) was observed to be ~1/3 of that in toluene (1057 bar). With increasing temperature, the solubility of oxygen in benzyl alcohol was found to increase, indicating that the oxygen-dissolving process is endothermic. Finally, an empirical correlation of the Henry’s law constant as a function of temperature was determined.
5.1 Introduction

Gas solubility data in organic solvents have great importance in chemical industry for the design of various unit operations (e.g. separators, reactors), and the solubility of oxygen is probably one of the most important. Generally, the solubility of oxygen in liquid could be experimentally measured with analytical methods or synthetic methods. These two main approaches are classified based on whether the equilibrium compositions are determined (analytically) and whether the total mixture is prepared with predetermined amounts of the components (synthetically). For example, Anthony et al. presented the measurement of oxygen solubility in ionic liquid using an analytical gravimetric method. The compositions were calculated according to the weight change of liquids on a gravimetric microbalance before and after gas adsorption. Lopez-Castillo et al. measured the oxygen solubility in CO$_2$-expanded liquids through analyzing the samples of the vapour and liquid phases at equilibrium with a gas chromatograph. Fischer and Wilken determined oxygen solubility in organic solvents using a static synthetic method, where the amount of pure components was precisely known and the changes of temperature and pressure during experiments were monitored.

Oxidation of alcohols is one of the most fundamental organic transformations and has great importance in chemical synthesis. Traditionally, this type of reactions involves stoichiometric inorganic oxidants (chromium (VI) or permanganate), which raises seriously environmental problems. From the view of green chemistry, there has been a growing demand to use molecular oxygen as oxidant for catalytic oxidation of alcohols. The amount of available oxygen influences the extent of oxidation reactions. On one hand, local shortage of oxygen, caused by the low
solubility of oxygen in solvents and poor mass transfer, could potentially limit the reaction rate.\textsuperscript{4, 11, 12, 35, 101, 144-147} On the other hand, excess of oxygen might cause the over-oxidation of products or catalysts.\textsuperscript{4, 38, 43, 148} Mixtures of oxygen and organic solvents under elevated temperatures and pressures could also raise serious safety issues.\textsuperscript{10} Thus, more data on oxygen solubility under reaction pressures and temperatures are highly required.

Recent advances in membrane science provide the opportunity to develop new methods for the measurement of oxygen solubility. Particularly, Teflon AF-2400 uniquely combines excellent chemical resistance, thermal stability, and mechanical properties with high fractional free volume.\textsuperscript{114} Its permeability characteristics to gas and liquid assure simultaneously high flux of gas and low permeation of liquid through the membrane. The Ley group first developed a tube-in-tube membrane contactor, which comprised an inner Teflon AF-2400 tube and an outer PTFE tube, and applied to various reactions (ozonolysis, hydrogenation, etc.).\textsuperscript{116-118} Generally, liquid is pumped through the inner tube and gas is pressurized at the annulus between the inner and outer tubes. When the liquid is flowing within the inner tube, the pressurized gas can permeate through the semi-permeable tube and be dissolved in the liquid until equilibrium is reached. The saturation concentration of gas in the liquid within the tube-in-tube contactor is observed to approximately obey Henry’s law.\textsuperscript{123, 147} In this study, the Teflon AF-2400 tube-in-tube membrane contactor was employed as a novel apparatus for the measurement of gas solubility in liquids, as exemplified by the case of oxygen in toluene and benzyl alcohol.
CHAPTER 5. A Novel Approach for Measuring Gas Solubility in Liquids Using a Tube-in-Tube Membrane Contactor

5.2 Experimental

5.2.1 Materials

Oxygen (99.995%) used in this study was obtained from BOC. Toluene (≥99.5%) and benzyl alcohol (≥99%) were obtained from Sigma-Aldrich. All the materials were directly used without further purification.

5.2.2 Apparatus

![Figure 5-1 Schematic of the tube-in-tube membrane contactor set-up used for the measurement of Henry’s law constants. MFC: mass flow controller; PUMP: HPLC pump; P: pressure sensor; T: Thermocouple used for temperature control; BPR: back pressure regulator.](image)

The schematic of the tube-in-tube membrane contactor set-up is shown in Figure 5-1. The membrane contactor consists of an inner Teflon AF-2400 tube (inner diameter (ID): 0.8 mm; outer diameter (OD): 1.0 mm; Biogeneral) and an outer PTFE tube (ID: 2.4 mm; OD: 3.2 mm; Agilent). The various parts of the set-up were connected with PFA tubing (ID: 1.0 mm; OD: 1.6 mm) unless otherwise specified. The total length of the tube-in-tube membrane contactor was 100 cm. Liquid (toluene or benzyl alcohol) was pumped through the inner tube, while oxygen flowed...
in the annulus between the inner and outer tubes. The liquid flow rate was controlled with a HPLC pump (Knauer, P 2.1S) and the liquid pressure was maintained by a back pressure regulator (BPR, Zaiput, BPR-01). The oxygen flow was controlled with a mass flow controller (Brooks, GF40 series) and the oxygen pressure ($P_{O_2}$) was sustained by another back pressure regulator (Swagelok, K series). The actual pressures of gas and liquid were measured by two pressure sensors (Zaiput, Hastelloy/PFA wetted parts).

The tube-in-tube membrane contactor was submerged in a stirred oil bath and the temperature of the oil bath was controlled by a hot plate fitted with a thermocouple (Stuart, US152). Liquid was pre-bubbled with oxygen using a gas sparger for 0.5 h at room temperature and atmospheric pressure to remove other dissolved gases. When the oil bath temperature and oxygen pressure stabilized, the liquid was pumped into the contactor. Oxygen passed through the semi-permeable membrane and dissolved into the liquid, reaching equilibrium in the contactor when the liquid residence time was long enough.\textsuperscript{119, 147} When the liquid flowed out of the contactor and passed through the BPR, the dissolved oxygen outgassed from the liquid due to decompression. The mixture of gas and liquid was directed to a 250 mL glass vessel separator, pre-fluxed with pure oxygen and located in a water bath maintained at 298 K, where gas and liquid were separated. The liquid collected in the separator was thus saturated with oxygen at 1 bar pressure and 298 K. The gas flow out from the separator included the contribution from the outgassed oxygen saturated with vapour and the gas volume change in the separator caused by the collection of liquid (see Figure 5-1). It was measured with a 1 mL bubble flow meter connected to the outlet of the separator by a silicon rubber tubing (~3 mm ID). The bubble meter contained
~1 mL soapy water which was assumed to be saturated with oxygen. Complete sealing of the separator was checked to avoid measurement errors caused by oxygen loss to the surroundings.

The experiment was carried out by varying the oxygen pressure from 1 to 10 bar, while keeping the liquid pressure at 10 bar (all pressures shown in this study are absolute). The temperature of the experiment was from 298 to 393 K. At a given oxygen pressure, three different liquid flow rates (0.25-0.45 mL/min) were applied and the gas flow rate from the separator was measured at least three times at each liquid flow rate.

5.2.3 Calculation Procedure

Liquid dissolution in the membrane and further permeation to the gas phase is neglected in the membrane contactor. In the separator, the gas phase is assumed to be saturated with organic vapour, which is needed to be subtracted from the measured gas flow rate. However, the loss of organic solvent into the gas phase can be neglected due to its low amount (< 0.1% of the liquid flow). Henry’s law constant ($H$) at given temperature $T$ was defined by

$$H = \frac{P_{O2}}{x(T)}$$  \hspace{1cm} (5-1)

where $P_{O2}$ is the oxygen pressure and $x$ is the mole fraction of oxygen in the liquid at given temperature $T$. The amount of oxygen outgassed per liquid volume ($\phi_{O2}$) can be expressed as (detailed derivation can be found in Appendix A):

$$\phi_{O2}(T) = \frac{\hat{v}_{O2} \rho_L}{M_L H} P_{O2} - \frac{\hat{v}_{O2} \rho_L}{M_L H_{298 K} P_{atm}}$$  \hspace{1cm} (5-2)

where $T$ is the experiment temperature, $\hat{v}_{O2}$ is molar volume of oxygen at STP.
(22400 mL/mol), $\rho_L$ is the density of liquid (g/mL), $M_L$ is molar mass of liquid (g/mol), $P_{O2}$ is the oxygen pressure (bar), $P_{atm}$ is the atmospheric pressure (1 bar), $H$ is Henry’s law constant at experiment temperature $T$, $H_{298K}$ is Henry’s law constant at 298K. By plotting $\phi_{O2} \text{ vs } P_{O2}$, the Henry’s constant at given temperature can be calculated.

5.3 Results and Discussion

Initially, the method was validated by measuring oxygen solubility in toluene. The measurements were carried out at 298 and 348 K with oxygen pressures up to 10 bar. The experimental data of $\phi_{O2} \text{ – } P_{O2}$ are plotted in Figure 5-2. The slopes of the fitting lines were 0.20001 mL/(mL Liquid ∙ bar) for 298 K and 0.21922 mL/(mL Liquid ∙ bar) for 348 K. Henry’s law constants were calculated to be 1057 bar and 965 bar, respectively. The Henry’s law constants are compared with literature data (obtained with static synthetic apparatuses)\textsuperscript{143, 149, 150} in Figure 5-3. The values in this work are very close to literature ones, and the differences at the same temperature ($\Delta H/H \times 100\%$) are within 5%, proving the reliability of this method. The Henry’s law constants of oxygen in toluene were also calculated based on the conditions of oxygen-toluene equilibrium obtained with equations of state (Peng-Robinson, UNIQUAC, NRTL, etc.) in ASPEN PLUS (shown in Appendix A). However, the simulated Henry’s law constants were observed to increase with the temperature increasing, and big differences were observed between different equations of state, especially at high temperature. These suggest that the equations of state used in ASPEN should be further modified.\textsuperscript{143, 149, 150}
CHAPTER 5. A Novel Approach for Measuring Gas Solubility in Liquids Using a Tube-in-Tube Membrane Contactor

Figure 5-2 Amount of oxygen outgassed from liquid ($\phi_{O_2}$) as a function of oxygen pressure ($P_{O_2}$) for oxygen-toluene at a) 298 K and b) 348 K.

Figure 5-3 Comparison of Henry’s law constants of oxygen in toluene obtained from this work with literature.\textsuperscript{143, 149, 150}
After validation of the method, oxygen solubility in benzyl alcohol was measured at temperatures from 298 K to 393 K with oxygen pressures up to 10 bar. The experimental $\phi_{O_2} - P_{O_2}$ data are shown in Figure 5-4. The slopes of the fitting lines, together with the calculated Henry’s law constants and the mole fraction of oxygen (x) in the benzyl alcohol under oxygen pressure 1 bar, are summarized in Table 5-1.

**Figure 5-4** Amount of oxygen outgassed from liquid ($\phi_{O_2}$) as a function of oxygen pressure ($P_{O_2}$) for oxygen-benzyl alcohol at a) 298 K, b) 313 K, c) 333 K, d) 353 K, e) 373 K and f) 393 K.
To provide oxygen solubility data for aerobic oxidation of alcohols, the saturation concentration of oxygen \((c_{O2})\) in benzyl alcohol under oxygen pressure 1 bar is also provided at each temperature. As seen in Table 5-1, the solubility of oxygen in benzyl alcohol increases with temperature. This trend agrees with that of oxygen in toluene, indicating that the solubilisation is endothermic.\(^{149}\) \(^{150}\) The Henry’s law constant for oxygen in benzyl alcohol at 298 K is 3462 bar, which is three times higher than that in toluene (1057 bar). This indicates a less favourable interaction between oxygen and benzyl alcohol than oxygen with toluene, possibly caused by a lower value for oxygen-benzyl alcohol enthalpy.\(^{150}\)

The oxygen-benzyl alcohol equilibrium compositions at different temperatures were also determined in ASPEN PLUS, and the calculated Henry’s law constants are shown in Appendix A. The calculated Henry’s law constants were much lower than the experimental values. An uptrend was also observed with the temperature increasing, which was similar to the oxygen-toluene equilibrium obtained in ASPEN.

| Table 5-1 Henry’s law constants and oxygen solubilities at 1 bar in benzyl alcohol |
|-----------------|------------------|-----------------|--------|
| \(T\) [K]       | Slope [mL/(mL\(_{\text{Liquid}}\) bar)] | Mole Fraction \(x \times 10^3\) [-] | Saturation Concentration \(c_{O2}\) [mM] | \(H\) [bar] |
| 298             | 0.06223          | 0.289           | 2.78   | 3462   |
| 313             | 0.06468          | 0.300           | 2.89   | 3332   |
| 333             | 0.06658          | 0.309           | 2.97   | 3237   |
| 353             | 0.06933          | 0.322           | 3.10   | 3108   |
| 373             | 0.07290          | 0.338           | 3.25   | 2956   |
| 393             | 0.07479          | 0.347           | 3.34   | 2881   |

At low pressure, Henry’s law constants can be considered to be independent of pressure, and thus they can be expressed as a function of temperature.\(^{149-151}\)

\[
\ln(H) = \frac{a}{T} + b \quad (5-3)
\]
where $a$ and $b$ are fitting parameters. The experimental data from Table 5-1 are plotted in Figure 5-5, as a $\ln(H)$ vs $1/T$ plot, and a straight line is obtained with correlation coefficient $R^2 = 0.987$. Thus, the correlation between Henry’s law constant and temperature for oxygen and benzyl alcohol system can be described by:

$$\ln(H) = \frac{228}{T} + 7.39 \quad (5-4)$$

![Figure 5-5 Henry’s law constants (bar) for oxygen in benzyl alcohol as a function of temperature.](image)

### 5.4 Conclusions

A novel apparatus for the measurement of oxygen solubility in toluene and benzyl alcohol was demonstrated using a semi-permeable Teflon AF-2400 tube-in-tube membrane contactor. The measured oxygen solubility in toluene is consistent with literature data, proving the reliability of this method. The oxygen solubilities in toluene and benzyl alcohol are both found to increase with temperature. However, the oxygen solubility in benzyl alcohol is much lower than that in toluene. This has implications for accurate reactor design in aerobic oxidation of alcohols. It is
important to measure solubilities in the solvents/substrates used in the actual reactions, as using solubilities of similar (but not identical) components can lead to errors in reactor design. As the Teflon AF-2400 membrane is highly permeable to other gases, such as carbon monoxide and hydrogen, this method is not limited to oxygen and organic solvents but is generally applicable for measurement of other gas solubilities in a wide range of liquids.
CHAPTER 6. Development of a Flat Membrane Microreactor for Scalable Aerobic Oxidation of Benzyl Alcohol in Flow

A Teflon AF-2400 flat membrane microreactor was demonstrated for safe and scalable oxidation of solvent-free benzyl alcohol with molecular oxygen on Au-Pd/TiO$_2$ catalysts. The microreactor employed a mesh supported Teflon AF-2400 flat membrane, with gas and liquid channels on each side. Catalyst particles were packed in the liquid flow channel. The effect of oxygen pressure was studied, and the oxygen was shown to have a positive effect on the oxidation of benzyl alcohol in the flat membrane microreactor. A conversion of benzyl alcohol of 70% with 71% selectivity to benzaldehyde was obtained at 1150 g$_{\text{cat}}$/s/g$_{\text{alcohol}}$, 8.4 bar oxygen pressure and 10 bar liquid pressure. Doubling the membrane thickness led to a 20% drop of oxygen consumption rate, indicating the main oxygen transfer resistance not existing in the membrane. When changing the catalyst particle size and the liquid flow rate, no significant effect was observed on the oxidation reaction rate. An effectiveness factor approach is proposed to predict the effect of oxygen permeation and transverse mass transfer on the catalyst packed in the membrane reactor, which suggests that the oxidation of benzyl alcohol on the highly active Au-Pd/TiO$_2$ catalyst is controlled by the oxygen transverse mass transfer in the bulk liquid within the catalyst bed. Scale-up of the flat membrane microreactor was demonstrated through widening the liquid channel width by ~10 times, which increased the reactor productivity by a factor of 8.
6.1 Introduction

Oxidation of alcohols is one of the most important processes in organic chemistry and often performed with stoichiometric inorganic reagents.\textsuperscript{2, 5, 6} To improve the atom efficiency and reduce the environmental costs, heterogeneous catalysts have been developed, and significant advances have been reported.\textsuperscript{3, 4, 12, 20, 60} However, large-scale applications of aerobic oxidation of alcohols are still limited by potential safety issues caused by the oxidant-organic reactant mixtures.\textsuperscript{7, 12, 38}

Recently, several approaches have been proposed to ensure intrinsic process safety. Stahl \textit{et al}.\textsuperscript{152, 153} reported a continuous-flow tube reactor for homogeneous Pd-/Cu-catalysed aerobic oxidation of primary alcohols to aldehydes, where a dilute oxygen source (8-9\% oxygen in nitrogen) was used to avoid the oxygen/organic mixture entering the explosive regime. Zotova \textit{et al}.\textsuperscript{11} developed a safe process for aerobic oxidation of alcohols with a commercially available XCube\textsuperscript{TM} reactor, by pre-mixing and saturating the liquid with the gaseous reactant (oxygen or air) before reaching the catalyst bed.

Among these approaches, membrane reactors have attracted great attention, since a membrane can act as a well-defined contacting interface for gas and liquid phases.\textsuperscript{96, 103, 104} It allows strict control of the reactants and avoids direct mixing of oxygen with organic reactants. For inorganic membrane reactors, a ceramic membrane reactor has been developed in our group for oxidation of benzyl alcohol with pure oxygen.\textsuperscript{101} The reactor consisted of a commercially available tubular ceramic membrane with catalyst packed in the inner tube. The liquid phase flowed through the inner tube, and pure oxygen flowed in the opposite side of the membrane. Deficiency of oxygen in the catalyst bed area was suggested to occur for relatively
fast catalytic system. Polymeric membranes were also applied for this process and Teflon AF-2400 membrane is one of the most intensively studied. As one type of amorphous fluoroplastics, it has high oxygen permeability (1140 barrer, second only to poly(1-(trimethylsilyl)-1-propyne) (PTMSP)), together with excellent chemical compatibility. A Teflon AF-2400 tube-in-tube reactor, which is first developed in the Ley group, has been used to saturate the liquid stream with oxygen/air for catalytic oxidation of alcohols. An in-situ continuous supply of oxygen through the membrane to the catalyst was also realized through packing the solid catalyst inside the inner tube of the tube-in-tube reactor. This contributed to a significant improvement in both conversion and selectivity. Though the price of Teflon AF-2400 membrane is much higher than that of other common fluoropolymers such as PTFE membrane, the high oxygen permeability makes it attractive for oxidation of solvent-free benzyl alcohol within several-centimetre catalyst bed.

Several potential drawbacks still exist for the current packed tube-in-tube membrane reactor. The gas phase present in the annulus between the inner and outer tubes could possibly act as a heat transfer resistance for exothermic reactions, due to its relatively low thermal conductivity. The scale-up of the tube-in-tube reactor is not facile, since increasing the membrane diameter will result to a proportional increase in radial mass transfer resistance under laminar flow. In an attempt to address these issues, a Teflon AF-2400 flat membrane microreactor is developed in this study. Oxidation of benzyl alcohol is performed on Au-Pd/TiO₂ catalyst, which has high activity and good reusability. Oxidation of benzyl alcohol on Au-Pd/TiO₂ is considered to comprise of two main steps, the oxidation reaction (equation 6-1) and
the disproportionation reaction (equation 6-2).\textsuperscript{34, 35} Both the oxidation and disproportionation of benzyl alcohol can form the target product, benzaldehyde, while toluene by-product, is generated from disproportionation. The amount of consumed oxygen during the reaction could be calculated based on the conversion of benzyl alcohol and selectivities to benzaldehyde and toluene.\textsuperscript{147}

\begin{align*}
2\text{PhCH}_2\text{OH} + \text{O}_2 & \rightarrow 2\text{PhCHO} + 2\text{H}_2\text{O} \\
2\text{PhCH}_2\text{OH} & \rightarrow \text{PhCHO} + \text{PhCH}_3 + \text{H}_2\text{O}
\end{align*}

\textbf{6.2 Experimental Section}

\textbf{6.2.1 Catalyst Preparation}\textsuperscript{*}

A nominal 1 wt\% bimetallic Au-Pd/TiO\textsubscript{2} catalyst (from the same batch of catalyst as used in Chapter 3) was prepared by co-impregnation, similar to previous work.\textsuperscript{26, 154} TiO\textsubscript{2} (Degussa, P25) was first suspended in demineralized water by stirring, and then \text{HAuCl}_4\cdot3\text{H}_2\text{O} (Johnson Matthey) and \text{PdCl}_2 (Johnson Matthey) with Au-to-Pd weight ratio equivalent to 1:19 was added. The resultant slurry was spray-dried (nozzle temperature 220 °C), and then calcined in static air at 400 °C for 1 h. The metal content was analysed by ICP-AES and found as 0.05 wt\% Au and 0.85 wt\% Pd, while the metal particle size as observed by TEM was 1–2 nm.\textsuperscript{101} The powder was pelletized, crushed, and sieved to the desired particle size range.

\textsuperscript{*} Catalyst preparation and characterization was carried out by Dr Peter Ellis, Johnson Matthey.
6.2.2 Design of Flat Membrane Microreactor

The flat membrane microreactor designed in this work consisted of several layers, which are shown in Figure 6-1 (more details are shown in Appendix B, Figure B-1). From the bottom to the top, these are a liquid flow plate machined in 316 stainless steel (channel size length: 75 mm; width: 3 mm; depth: 1 mm), a Teflon AF-2400 membrane (length: 85 mm; width: 30 mm; thickness: 0.07 mm; Biogeneral), a 304 stainless steel mesh (length: 85 mm; width: 30 mm; thickness: 0.05 mm; hole size: 76 μm; open area: 23%; Industrial Netting; shown in Figure B-2), viton gasket (length: 85 mm; width: 30 mm; thickness: 1 mm; open area length: 75 mm; width: 3 mm; Altec) and a gas channel plate (with the same channel size and material to the liquid flow plate). The sealing of the reactor was achieved by compressing the membrane and the gasket with screws; the membrane also acted as a gasket for the liquid flow plate. An O-ring groove was machined in the liquid flow plate, which allowed sintered metal also to be applied as the membrane support. Two holes were
drilled in both liquid and gas channel plates (perpendicular, 1 mm away from the channel, shown in Figure B-1) for thermocouple insertion. A small circular piece of another nickel mesh (diameter: ~2 mm; thickness: 0.05 mm; hole size: 25 μm; Tecan) was placed between the liquid outlet and the fitting for retaining catalyst (shown in Figure B-1). After assembling the reactor, 50 mg silica beads (particle size: 90-125 μm) were packed into the liquid channel followed by the Au-Pd/TiO₂ catalyst particles, with the help of a vacuum pump. The length of the 100 mg catalyst bed was ~5 cm.

The reactor was placed on a hotplate fitted with a thermocouple. The thermocouple was inserted in the hole beneath the catalyst bed for temperature measurement and control. To decrease heat loss, an insulation cap (insulation thickness: ~2 cm; WDS® Ultra, Morgan) was made to cover the reactor. The temperature differences between the gas and the liquid flow plates during the reaction at 120 °C was measured through the thermocouples in the other holes, which was less than 3 °C.

6.2.3 Scale-up of the Flat Membrane Microreactor

To scale up the flat-membrane microreactor, the gas/liquid channel widths were increased to 32 mm. A picture of the scaled-up microreactor and a mechanical drawing are shown in the Appendix B. The length of the liquid flow channel was 68 mm, with a depth of 0.5 mm. More details on the reactor design are shown in Appendix B. The scaled-up reactor was heated with two silicone rubber heaters (length: 12.7 cm; width: 5.1 cm; thickness: 1.3 mm Watlow, wire-wound elements 020050C2; shown in Figure B-5) The whole reactor was insulated by an insulation cap (insulation thickness: ~2 cm; WDS® Ultra, Morgan), and the temperature
difference within the reactor was observed to be ± 2 °C at set temperature of 120 °C.

6.2.4 Reaction Experiments

A schematic of the experimental set-up is shown in Figure 6-2. Neat benzyl alcohol (99.0%, Sigma-Aldrich) was delivered into the microreactor with a HPLC pump (Knauer P2.1S). An adjustable back pressure regulator (BPR, Zaiput, BPR-01) was used at the liquid outlet to maintain the liquid pressure. Pure oxygen was regulated by a mass flow controller (Brooks, GF40 series) and directed to the gas inlet of the reactor. A BPR (1-11 bar, Swagelok, K series) was connected at the gas outlet to maintain the gas pressure. The pressures of the gas and liquid phases were monitored by pressure sensors (Zaiput, Hastelloy/PFA wetted parts) placed upstream of the reactor. All pressures reported in this study are absolute.

The effluent from the liquid outlet was collected in a cold trap (ice-water bath) and quantitatively analysed by a gas chromatograph (Agilent 7820A) fitted with a DB-624 capillary column and a flame ionization detector. Benzyl alcohol conversion ($X$) and selectivity ($S_i$) of each product were calculated according to the following equations:
\[
X = \frac{c_{\text{BnOH, in}} - c_{\text{BnOH, out}}}{c_{\text{BnOH, in}}} \times 100\% \quad (6-3)
\]

\[
S_i = \frac{c_i v_i}{c_{\text{BnOH, in}}} \times 100\% \quad (6-4)
\]

where \(c_{\text{BnOH, in}}\) and \(c_{\text{BnOH, out}}\) are the concentration of benzyl alcohol at the inlet and outlet, respectively and \(v_i\) is the number of moles of benzyl alcohol consumed for the production of 1 mole of product \(i\). In this study, \(S_B\) and \(S_T\) stand for selectivities to benzaldehyde and toluene, respectively.

Catalyst contact time (CCT) was used to characterise the reaction time of benzyl alcohol, and defined as:

\[
\text{CCT} = \frac{m_{\text{cat}}}{v p} \quad (6-5)
\]

where \(m_{\text{cat}}\) is the mass of catalyst, \(v\) is the inlet volume flow rate of benzyl alcohol, \(p\) is the density of benzyl alcohol at 20 °C. For each experiment, at least three samples were collected and the results were averaged. The errors for the conversion and selectivities were less than ±2%. The quantitative oxygen consumption rates (OCR, at standard temperature and pressure, 0 °C and 1 bar, STP) were calculated by

\[
OCR = \frac{1}{2} \cdot F_{\text{BnOH, in}} \cdot X \cdot (S_B - S_T) \cdot \tilde{V}_{O2} \quad (6-6)
\]

where \(F_{\text{BnOH, in}}\) is the molar flow rate of benzyl alcohol, \(\tilde{V}_{O2}\) is the molar volume of oxygen at STP. Average turnover frequency (TOF) was calculated to better represent the reaction rate, where \(\text{TOF}_{\text{total}}, \text{TOF}_O\) and \(\text{TOF}_D\) correspond to the overall reaction, the oxidation reaction and the disproportionation reaction, respectively.

\[
\text{TOF}_{\text{total}} = \frac{F_{\text{BnOH, in}} \cdot X}{n_{\text{metal}}} \quad (6-7)
\]

\[
\text{TOF}_O = \frac{F_{\text{BnOH, in}} \cdot X \cdot (S_B-S_T)}{n_{\text{metal}}} \quad (6-8)
\]
\[ TOF_D = \frac{2F_{\text{BnOH.in}} X S_T}{n_{\text{metal}}} \] (6-9)

where \( n_{\text{metal}} \) is the moles of the metals contained in the packed catalyst.

6.3 Results and Discussion

6.3.1 Operation Pressure and Solvent Pervaporation

Initially, the operating pressure range of the flat membrane microreactor was studied without any catalyst packed. Reactor temperature was kept at 120 °C and atmospheric pressure was applied in the gas phase. The pressure in the liquid phase was gradually increased to 21 bar (20 bar pressure difference) and kept for 0.5 h. No leaking was detected. Noticably, only slight shape change of the membrane was observed after testing at such pressure difference and temperature (shown in Figure B-6a). In contrast, obvious membrane shape change was observed when no mesh was used to support the membrane (shown in Figure B-6b). No membrane breakage was observed, since the membrane touched the bottom of the gas channel due to the shape change. It should be noted that in our previous tube-in-tube membrane microreactor, the Teflon AF-2400 inner tube (1 mm thickness) was noticed to break when ~12 bar pressure difference was applied at 120 °C. This indicates the flat membrane microreactor could expand the operating pressure range of the Teflon AF-2400 membrane at elevated temperature.

Though direct mixing of oxygen with organic reactants is avoided by using the membrane reactor, organic vapour may appear in the gas phase through pervaporation. Polyakov et al.\textsuperscript{155} studied the permeation and sorption of various organic liquid (chlorinated hydrocarbons, lower alcohols, \textit{etc.}) in Teflon AF membrane within the temperature range 5-95 °C. To assure that the organic vapour
concentration remained outside the explosive limits, the maximum amount of organic vapour in the gas phase was experimentally measured using pure benzyl alcohol, benzaldehyde or toluene (the details are shown in Appendix B). Toluene presented the highest pervaporation rate through the membrane, which was 32.2 mg/h. It corresponds to 0.13 mL/min toluene vapour in the gas phase channel at STP. This highlights the necessity of using continuous flow in the gas phase to dilute the pervaporating organics. To keep the organic vapour concentration lower than 1 vol% (which is the lower explosive limit for toluene in air at 6 bar and 120 °C)\textsuperscript{156}, an oxygen flow rate of 15 mL/min at STP was applied in the following experiments.

6.3.2 Catalyst Stability

![Conversion or selectivity vs. time graph](image)

**Figure 6-3** Stability study of Au-Pd/TiO\textsubscript{2} catalyst in flat membrane microreactor. Reaction conditions: Au-Pd/TiO\textsubscript{2} catalyst (90-125 μm), 100 mg; neat benzyl alcohol, 10 μL/min; catalyst contact time, 577 g\textsubscript{cat}/s/g\textsubscript{alcohol}; oxygen pressure, 8.4 bar; liquid pressure, 10 bar; reaction temperature, 120 °C.

To evaluate the performance of the Au-Pd/TiO\textsubscript{2} catalyst and the membrane microreactor, a stability study was initially conducted and the results are shown in Figure 6-3. The catalyst contact time was 577 g\textsubscript{cat}/s/g\textsubscript{alcohol} and the conversion of
benzyl alcohol was stable around 57% over 24 h. The selectivities to benzaldehyde and toluene, the two main products, also presented stable trends (66% and 31%, respectively). The selectivities to other minor products (dibenzyl either, benzyl benzoate, etc.) were all less than 3% (not shown). These results demonstrate the high stability of the prepared Au-Pd/TiO$_2$ catalyst, as well as the excellent performance of the flat membrane microreactor. The productivity of benzaldehyde was calculated to be $2.3 \text{ g}_\text{B}/(\text{g}_\text{cat} \cdot \text{h})$.

6.3.3 Effect of Oxygen Pressure

To explore the effect of oxygen pressure on the oxidation of benzyl alcohol in the flat membrane microreactor, oxygen pressure was changed from 0 to 8.4 bar. Other reaction conditions were kept the same. From Figure 6-4, the conversion of benzyl alcohol was observed to be 4% at 0 bar oxygen pressure (nitrogen flowed through the gas channel at atmospheric pressure). Almost equimolar amounts of benzaldehyde (44%) and toluene (41%) were formed due to the disproportionation of benzyl alcohol.$^{34,35}$ With oxygen pressure increasing, the conversion gradually rose and reached 57% at 8.4 bar. The selectivity to benzaldehyde also increased from 44% to 66% within the investigated oxygen pressure range. Correspondingly, the selectivity to toluene drop from 41% to 31%. The same behaviour was also observed in previous tube-in-tube membrane microreactor.$^{147}$ When increasing the oxygen pressure from 3 to 7 bar, the conversion of benzyl alcohol increased from 22% to 42% with the selectivity to benzaldehyde from 61% to 65% at catalyst contact time $115 g_{\text{cat}}^{-1} S/g_{\text{alcohol}}^{-1}$.
CHAPTER 6. Development of a Flat Membrane Microreactor for Scalable Aerobic Oxidation of Benzyl Alcohol in Flow

Figure 6-4 Effect of oxygen pressure on the conversion and selectivities during oxidation of benzyl alcohol in the flat membrane microreactor. Reaction conditions: Au-Pd/TiO$_2$ catalyst (90-125 µm), 100 mg; neat benzyl alcohol, 10 µL/min; catalyst contact time, 577 g$_{cat}$/g$_{alcohol}$; liquid pressure, 10 bar; reaction temperature, 120 ºC.

Figure 6-5 Oxygen consumption rates at STP under different oxygen pressures. Reaction conditions: Au-Pd/TiO$_2$ catalyst (90-125 µm), 100 mg; neat benzyl alcohol, 10 µL/min; catalyst contact time, 577 g$_{cat}$/g$_{alcohol}$; liquid pressure, 10 bar; reaction temperature, 120 ºC.
The oxygen consumption rate at STP under different oxygen pressures is shown in Figure 6-5. Higher oxygen consumption rates were observed at higher gas pressure. The maximum rate was 0.21 mL/min (STP, 0.56 mmol/h) at 8.4 bar. The fact that the oxygen consumption rate was close to zero at 0 bar oxygen pressure (nitrogen was used in the gas channel), indicates that the amount of oxygen consumed in the reaction comes from oxygen permeating through the membrane during the reaction. On the basis that all other reaction conditions were kept the same and only the oxygen pressure was changed, increasing the oxygen pressure could therefore enhance the oxygen consumption and result in higher oxygen permeation.

To further investigate the effect of oxygen pressure on the reaction, TOFs under different oxygen pressures are calculated and shown in Figure 6-6. In the absence of oxygen (0 bar oxygen pressure), TOF_total was very close to TOF_D, and TOF_O was practically zero. This indicates that disproportionation was the only reaction...
occurring and no oxidation reaction was taking place under these conditions.\textsuperscript{34} With a rise in oxygen pressure, both TOF\textsubscript{D} and TOF\textsubscript{O} presented an upward trend without reaching a plateau. This contributed to an approximately linear increase in TOF\textsubscript{total}, as well as the increasing selectivity to benzaldehyde as shown in Figure 6-4. Since oxygen solubility in the liquid is increased at elevated oxygen pressure, this demonstrates that oxygen could promote both disproportionation and oxidation reactions,\textsuperscript{33} and thus have a positive effect on the oxidation of benzyl alcohol in the flat membrane microreactor with Au-Pd/TiO\textsubscript{2} catalyst. Higher TOF\textsubscript{total} and TOF\textsubscript{O} were observed in batch reactors and trickle bed reactors, when higher oxygen pressure was used.\textsuperscript{36} However, TOF\textsubscript{D} in both batch and trickle bed reactors was observed to decrease with oxygen pressure higher than 1 bar.\textsuperscript{36,137} This can be due to the twofold effect of oxygen on the disproportionation reaction possibly caused by the two different mechanisms for the disproportionation reaction, one anaerobic and the other aerobic (the detailed kinetics are under investigation).\textsuperscript{35,126} At low oxygen pressure, the increase of oxygen pressure could promote the disproportionation reaction, but the promotion effect was reduced with further increasing oxygen pressure.\textsuperscript{35,126} Moreover, TOFs obtained in the flat membrane microreactor (TOF\textsubscript{total}, 400 h\textsuperscript{-1} at 8.4 bar) were about two orders of magnitude lower than those reported in batch reactors using 2.5 wt% Au-2.5 wt% Pd/TiO\textsubscript{2} catalyst (TOF\textsubscript{total}, 26,400 h\textsuperscript{-1}).\textsuperscript{26} Since the TOF\textsubscript{D} was not observed to decrease in the flat membrane microreactor and much lower TOFs were obtained, the flat membrane microreactor might not supply oxygen efficiently as the batch and trickle bed reactors. So, the oxygen transfer resistance in the flat membrane microreactor was investigated in the following sections.
6.3.4 Effect of Membrane Thickness

Figure 6-7 Effect of membrane thickness on the conversion, selectivites and oxygen consumption rate during oxidation of benzyl alcohol in the flat membrane microreactor. Single layer, 0.07 mm; double layer, 0.14 mm. Reaction conditions: Au-Pd/TiO₂ catalyst (90-125 μm), 100 mg; neat benzyl alcohol, 5-30 μL/min; oxygen pressure: 8.4 bar; liquid pressure, 10 bar; reaction temperature, 120 ºC.
The effect of membrane thickness was investigated through placing a single-layer (0.07 mm thickness) or a double-layer (0.14 mm thickness) membrane in the microreactor. The results at different catalyst contact times are shown in Figure 6-7. The same trend was observed for the conversion and selectivities under different catalyst contact times, that is, increasing the catalyst contact time enhanced both the conversion of benzyl alcohol and the selectivity to benzaldehyde, while decreasing the selectivity to toluene. This trend agrees with our previous results in the membrane reactors, but differs from those obtained in batch and trickle bed reactors, in which the selectivities were similar under different catalyst contact times. This is probably caused by the low efficiency of oxygen supply in the membrane reactors as shown in Section 6.3.3. Longer catalyst contact time in the membrane reactor could contribute to longer time for both oxygen permeation and reaction, and thus more oxygen supply for per liquid volume. This could further promote the oxidation reaction, resulting in the increasing of the selectivity to benzaldehyde.

Notably, a conversion of benzyl alcohol of 70% with 71% selectivity to benzaldehyde was obtained at 1150 $g_{\text{cat}} \cdot s/g_{\text{alcohol}}$ and 8.4 bar oxygen pressure with single-layer membrane. At the same reaction conditions with double-layer membrane, the conversion and benzaldehyde selectivity were observed to be both 68%. When comparing the oxygen consumption rates under different membrane thicknesses (shown in Figure 6-7c), higher rate was observed for single-layer membrane at each catalyst contact time. The average oxygen consumption rate with single-layer membrane (0.19 mL/min) was 27% higher than that with double-layer (0.15 mL/min). This reveals that oxygen transfer resistance exists in the membrane.
However, the oxygen transfer resistance in the membrane seems not to be the main transfer resistance, since only 27% increase of average oxygen consumption rate was observed when halving the membrane thickness (i.e., halving the oxygen transfer resistance in the membrane).

6.3.5 Internal/External Mass Transfer Study

![Graph showing the effect of catalyst particle size on conversion and selectivities during oxidation of benzyl alcohol in the flat membrane microreactor. Reaction conditions: Au-Pd/TiO$_2$ catalyst, 100 mg; oxygen pressure: 8.4 bar; liquid pressure, 10 bar; reaction temperature, 120 °C.](image)

Figure 6-8 Effect of catalyst particle size on the conversion and selectivities during oxidation of benzyl alcohol in the flat membrane microreactor. Reaction conditions: Au-Pd/TiO$_2$ catalyst, 100 mg; oxygen pressure: 8.4 bar; liquid pressure, 10 bar; reaction temperature, 120 °C.
The internal mass transfer was studied using two ranges of catalyst particle sizes at different catalyst contact times. The conversion of benzyl alcohol and selectivities to benzaldehyde and toluene are shown in Figure 6-8a&b. Slightly higher conversion was observed when larger particles were used. This is possibly caused by the packed bed being slightly longer (caused by larger packed bed void volume) than that with small particles for the same catalyst mass, which contributes to longer time for oxygen permeation and reaction. The selectivities to benzaldehyde and toluene were very similar, indicating independence on the catalyst particle size. Due to the unknown (and likely transversely and axially varying) oxygen concentration within the catalyst bed, the Weisz-Prater criterion cannot be used to quantitatively determine if internal diffusion is limiting the reaction. The results with the two catalyst particle sizes indicate that internal mass transfer resistance is not a determining factor for the performance of the system investigated.

Table 6-1 Comparison of TOFs for different liquid flow rates. Reaction conditions: Au-Pd/TiO$_2$ catalyst (90-125 μm), 50 or 100 mg; catalyst contact time, 192-577 g$_{cat}$/g$_{alcohol}$; oxygen pressure: 8.4 bar; liquid pressure, 10 bar; reaction temperature, 120 ºC.

<table>
<thead>
<tr>
<th>CCT [g$<em>{cat}$/g$</em>{alcohol}$]</th>
<th>577</th>
<th>288</th>
<th>192</th>
</tr>
</thead>
<tbody>
<tr>
<td>Catalyst weight [mg]</td>
<td>50</td>
<td>100</td>
<td>50</td>
</tr>
<tr>
<td>Liquid flow rate [uL/min]</td>
<td>5</td>
<td>10</td>
<td>10</td>
</tr>
<tr>
<td>TOF$_{10}$/TOF$_5$</td>
<td>1.03</td>
<td>1.14</td>
<td>1.28</td>
</tr>
<tr>
<td>TOF$<em>{20}$/TOF$</em>{10}$</td>
<td>1.14</td>
<td>1.24</td>
<td>1.43</td>
</tr>
<tr>
<td>TOF$<em>{30}$/TOF$</em>{15}$</td>
<td>1.28</td>
<td>1.43</td>
<td>1.43</td>
</tr>
</tbody>
</table>

The external mass transfer in the flat membrane microreactor was studied through comparing the TOF under two different liquid flow rates at the same catalyst contact time. In Table 6-1, no clear difference was seen in all the TOFs when the liquid flow
rate was increased from 5 μL/min to 10 μL/min at 577 g\text{cat}\cdot s/g_{\text{alcohol}}. When liquid flow rate was increased from 10 μL/min to 20 μL/min at shorter catalyst contact time (288 g\text{cat}\cdot s/g_{\text{alcohol}}), TOF\text{total} increased by 14%. This is caused by the increase of TOF\text{D} (24%), despite the slight decrease in TOF\text{O}. Higher increase of TOF\text{total} (28%) was observed with liquid flow rate increasing from 15 μL/min to 30 μL/min at 192 g\text{cat}\cdot s/g_{\text{alcohol}}. A rise of 44% was also shown in TOF\text{D}, together with 12% drop in TOF\text{O}. The changes of TOF\text{O} could be possibly caused by the experimental errors for the low conversions at high flow rate. However, the changes of TOF\text{D} illustrate stronger influence of the liquid flow rate on the disproportionation reaction than the oxidation reaction.

The flat membrane microreactor has been indicated to not supply oxygen efficiently as the batch and trickle bed reactors in Section 6.3.3 & 6.3.4. Note that the direction of oxygen transfer in the membrane reactor is perpendicular to that of liquid flow (in laminar flow regime), and oxygen needs to permeate the membrane, dissolve in the liquid, transversely transfer through the bulk liquid within catalyst packed bed and reach the catalyst surface. Different liquid flow rates could potentially affect both the external mass transfer and the transverse mass transfer of oxygen in the bed, and thus affect the oxygen supply to the catalyst. The stronger influence of liquid flow rate on the disproportionation reaction can be possibly attributed to that the disproportionation reaction is more sensitive to oxygen at low oxygen supply due to the two different reaction mechanisms. A slight change of oxygen supply caused by the change of liquid flow rate may have a negligible effect on the oxidation reaction but a significant effect on the disproportionation reaction.

Based on the TOF\text{total} (26,400 h\text{\textsuperscript{-1}}) and the selectivity to benzaldehyde (~67%)
obtained with 2.5 wt% Au-2.5 wt% Pd/TiO₂ catalyst in the batch reactor, the TOF₀ of the Au-Pd/TiO₂ catalyst could go up to ~8800 h⁻¹ at 120 °C. This results to an oxygen demand of 13.6 mL/min at STP (0.0365 mol/h, calculation shown in SI) for the amount of Au-Pd/TiO₂ catalyst packed in the membrane reactor (100 mg). The actual oxygen consuming rate within the membrane reactor was only 0.21 mL/min at 8.4 bar oxygen pressure. This indicates a gap between oxygen demand by the highly active catalyst and oxygen supply in the membrane reactor. However, the maximum oxygen supply rate (the highest oxygen permeation rate) through the membrane area contacting the catalyst bed (~50 mm× 3 mm) could be 0.93 mL/min at 8.4 bar oxygen pressure, which is 4-fold of the oxygen consuming rate. The difference between the maximum oxygen supply rate and the actual oxygen consumption rate indicates the oxygen transfer resistance in the catalyst bed. Since the internal/external mass transfer resistance have been shown to have no significant effect on the oxidation reaction rate, the main oxygen transfer resistance is thus suggested to exist in the bulk liquid within the catalyst bed.

### 6.3.6 Interaction of Mass Transfer Processes with Reaction in the Packed Bed Membrane Reactor

In order to assess the relative importance of the various mass transfer processes that affect reactor performance (membrane oxygen permeation and transverse mass transfer), an effectiveness factor (η) is proposed based on the oxygen reaction rate (assumed as a first order in oxygen, the details are shown in Appendix B). By considering an analogy between our reactor and a single pellet where diffusion/reaction occurs, where the oxygen mass transfer through the membrane and transverse mass transfer in the reactor bed are analogous to external and internal
mass transfer in a pellet respectively, we can use the standard effectiveness factor-
Thiele modulus analysis.

To relate the reaction rate to the transverse mass transfer rate in the packed bed, 
Thiele modulus ($\phi$) is defined by

$$\phi^2 = \frac{d_1^2 k_{O_2}}{D_{O_2,T}}$$

(6-10)

where $d_1$ is the depth of the packed bed (liquid channel), $k_{O_2}$ is the oxygen reaction 
rate coefficient based on the bed volume, $D_{O_2,T}$ is the transverse mass transfer 
coefficient of oxygen in the packed bed.

To provide a measure of oxygen permeation rate to oxygen transverse mass 
transfer rate, a Biot number ($Bi$) is defined by

$$Bi = \frac{d_1 k_m}{D_{O_2,T}}$$

(6-11)

where $k_m$ is the oxygen mass transfer coefficient in the Teflon AF-2400 
membrane.\textsuperscript{119}

So, the permeation-transverse mass transfer effectiveness factor ($\eta$) is described 
by

$$\eta = \frac{\sinh \phi}{\sqrt{\cosh \phi + \frac{\phi}{Bi} \sinh \phi}}$$

(6-12)

For the Au-Pd/TiO$_2$ catalyst in the current flat membrane microreactor, the Biot 
number is 24.8 (see Appendix B), which indicates that the oxygen permeation rate in 
the membrane is much higher than the oxygen transverse mass transfer rate in the 
packed bed. This results to slight increase of oxygen consumption rate when halving 
the membrane thickness. Using the estimated reaction rate coefficient from the batch
reactor, the Thiele modulus is 39.4, which suggests that the oxygen reaction in the membrane reactor is highly transverse mass transfer-controlled. The permeation-transverse mass transfer effectiveness factor is found to be 1.0%, which agrees well with the ratio (1.5%) of the observed TOFo in the membrane reactor (~130 h⁻¹) to the batch reactor (~8800 h⁻¹).

Since the oxygen transverse mass transfer rate is inversely proportional to the liquid channel depth, reduction in the liquid channel depth is expected to significantly improve the performance of the membrane microreactor for oxidation of benzyl alcohol. To confirm this, another flat membrane microreactor was fabricated with half of liquid channel depth (0.5 mm) but same channel width (3 mm) as compared to the previous. The amount of packed catalysts was also halved to keep the length of catalyst bed the same. At the same catalyst contact time (577 g_cat·s/g_alcohol) and oxygen pressure (8.4 bar), a 75% conversion of benzyl alcohol with a 67% selectivity to benzaldehyde was obtained in the 0.5 mm channel depth reactor, as compared to 57% conversion and 66% benzaldehyde selectivity in the 1.0 mm channel depth reactor. This proves that the main oxygen transfer resistance exists in the bulk liquid within the catalyst bed.

To predict catalyst performance in the packed bed membrane reactor and guide reactor design, the effectiveness factor is shown as a function of an observable modulus ($\eta \Phi^2$) for various values of Biot number in Figure 6-9, together with the various regimes of possible controlling processes. When the reactor is in the transverse mass transfer controlling regime (as the symbol ■ and ● shown in the Figure 6-9, which indicated the values for the current flat membrane reactor with the liquid flow channel depth of 1.0 mm and 0.5 mm, respectively), decreasing the
channel depth can decrease the values of $\phi$ and $Bi$, which makes the effectiveness factor move towards the permeation or reaction controlling regime. When the reactor is in the permeation controlling regime, decreasing the membrane thickness can increase the value of $Bi$, which makes the effectiveness factor move towards the reaction or transverse mass transfer controlling regime. Similarly, the value of $\phi$ will be increased with the reaction rate coefficient increasing in the reaction controlling regime. This makes the effectiveness factor move towards the permeation or transverse mass transfer controlling regime.

![Diagram showing permeation-transverse mass transfer catalytic effectiveness factor ($\eta$) as a function of the observable modulus ($\eta\phi^2$) for various values of Biot number ($Bi$) with the various regimes of possible controlling processes. (Symbols indicate the values for the current flat membrane reactor with the liquid flow channel depth of 1.0 mm (■) and 0.5 mm (●)).](image)

6.3.7 Performance of the Scaled-up Membrane Reactor

Scale-up of the flat membrane reactor was carried out through widening the channel width by ~10 times, and the results are shown in Table 6-2. The reactor was kept vertical and the liquid flowed from top to bottom. As compared to the previous
microreactor (channel width, 3 mm), the scaled-up microreactor presented a lower conversion of benzyl alcohol with a similar selectivity to benzaldehyde for the same catalyst contact time. The relative drop in conversion of ~17% may be due to some fluid maldistribution in the wider liquid channel. Nevertheless, the production rate of the benzaldehyde was demonstrated to be increased by a factor of 8 in the scaled-up microreactor.

Table 6-2 Comparison of reaction results in reactors with different channel width. Reaction conditions: Au-Pd/TiO$_2$ catalyst (90-125 μm); catalyst contact time, 1150 g$_{\text{cat}}$/g$_{\text{alcohol}}$; liquid pressure, 10 bar; reaction temperature, 120 °C.

<table>
<thead>
<tr>
<th>Reactor</th>
<th>Channel width [mm]</th>
<th>Catalyst packed [g]</th>
<th>Liquid flow rate [μL/min]</th>
<th>Oxygen pressure [bar]</th>
<th>X [%]</th>
<th>S$_B$ [%]</th>
<th>S$_T$ [%]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Small</td>
<td>3</td>
<td>0.1</td>
<td>5</td>
<td>8.4</td>
<td>70</td>
<td>71</td>
<td>26</td>
</tr>
<tr>
<td>Large</td>
<td>32</td>
<td>1.0</td>
<td>50</td>
<td>8.2</td>
<td>58</td>
<td>69</td>
<td>26</td>
</tr>
</tbody>
</table>

Table 6-3 Effect of liquid flow directions on the performance of scaled-up microreactor. Reaction conditions: Au-Pd/TiO$_2$catalyst (90-125 μm), 1.0 g; liquid flow rate, 50 μL/min; catalyst contact time, 1150 g$_{\text{cat}}$/g$_{\text{alcohol}}$; oxygen pressure, 8.2 bar; liquid pressure, 10 bar; reaction temperature, 120 °C.

<table>
<thead>
<tr>
<th>Flow directions</th>
<th>a)</th>
<th>b)</th>
<th>c)</th>
</tr>
</thead>
<tbody>
<tr>
<td>X [%]</td>
<td>58</td>
<td>55</td>
<td>54</td>
</tr>
<tr>
<td>S$_B$ [%]</td>
<td>69</td>
<td>66</td>
<td>67</td>
</tr>
<tr>
<td>S$_T$ [%]</td>
<td>26</td>
<td>30</td>
<td>29</td>
</tr>
</tbody>
</table>

The effect of liquid flow direction on the performance of the scaled-up microreactor was further investigated with the same catalyst packing. The reactor was kept vertical with liquid flowing from top to bottom (Table 6-3a) or reverse (Table 6-3b), or horizontal with liquid flowing from left to right (Table 6-3c). As shown in Table 3a, the conversion of benzyl alcohol was slightly lower (~5%) under the same conditions to that in Table 6-2, which could be caused by a slight catalyst
deactivation after ~50 h reaction. However, for the three different flow directions, no obvious difference was observed in terms of conversion and selectivities. This indicates that the liquid flow direction has no effect on the liquid distribution within the liquid channel.

Finally, oxidation of benzyl alcohol was carried out with air in the scaled-up flat membrane microreactor, since air is cheaper and easier use as compared to pure oxygen. As shown in Figure B-8, the conversion of benzyl alcohol was 27%, with a selectivity of 63% to benzaldehyde. Under the same reaction conditions with pure oxygen, the conversion and selectivity to benzaldehyde were 65% and 68%, respectively. The different performance in terms of conversion and selectivity was caused by the different partial pressures of oxygen; as discussed in Figure 6-6, oxygen is beneficial for both conversion and benzaldehyde selectivity. These results highlight the efficiency of using pure oxygen, suggesting that to achieve similar performance with air, higher gas pressure would need to be utilised.

6.4 Conclusions

A Teflon AF-2400 flat membrane microreactor was developed and applied for continuous-flow aerobic oxidation of solvent-free benzyl alcohol on Au-Pd/TiO₂ catalyst. As compared to previous tube-in-tube membrane microreactor, the flat configuration provides wider operating pressure range and is easier to scale-up. The reactor can be combined with plate heat exchangers to realize better temperature control for highly exothermic reactions. Higher reaction rates and main product selectivities were observed at higher oxygen pressure, highlighting the importance of oxygen concentration for the reaction within the membrane reactor. Study of the various oxygen transfer processes (permeation through membrane, transverse mass
transfer in the catalyst bed, internal/external transfer in the catalyst particles) indicated that the main oxygen transfer resistance was transverse mass transfer in the bed. This agreed with an effectiveness factor analysis akin to internal/external mass transfer and reaction in a catalytic particle, which further provides guidance on catalyst choice and membrane reactor design. Scale-up of the microreactor by an order of magnitude was demonstrated by increasing the width of the catalyst bed channel. The simple assembly of the flat membrane microreactor suggests that other similar flat membranes could be used for scalable flow oxidation of alcohols with molecular oxygen.
CHAPTER 7. Aerobic Oxidation of Benzyl Alcohol on Ru/Al₂O₃ Catalyst in a Flat Membrane Microchannel Reactor: an Experimental and Modelling Study

A flat Teflon AF-2400 membrane microchannel reactor was experimentally and theoretically investigated for aerobic oxidation of benzyl alcohol on 5 wt% Ru/Al₂O₃ catalyst. The flat membrane microchannel reactor consisted of gas and liquid channels (75 mm (L) × 3 mm (W) × 1 mm (D)), separated by a 0.07 mm thick semipermeable Teflon AF-2400 flat membrane, which allowed continuous supply of oxygen during the reaction and simultaneously avoided direct mixing of gaseous oxygen with organic reactants. A stability test was first carried out, and the experimental data obtained from the stabilization period were used to estimate the kinetics of benzyl alcohol oxidation with a 2D reactor model. Using the estimated kinetics, the modelling results obtained from the 2D reactor model agreed well with the experimental data at different liquid flow rates and oxygen pressures. The mass transfer and catalytic reaction in the membrane microchannel reactor were then theoretically studied by changing the membrane thickness, the liquid channel depth, and the reaction rate coefficient. Oxygen transverse mass transfer in the catalyst bed was found to be the controlling process, and decreasing the liquid channel depth is suggested to improve the oxygen supply and enhance the benzyl alcohol conversion in the membrane reactor.
7.1 Introduction

Oxidation of alcohols is an important process in chemical industry, and the high demand for sustainable processes has driven researchers in both academia and industry to employ solid catalysts and molecular oxygen rather than traditional, toxic chemical oxidants.4, 6 However, aerobic oxidation of alcohols on solid catalysts is still associated with several issues, such as lack of catalysts with sufficient activity and stability, and complex mass transport processes and reaction mechanisms.12 Moreover, potential safety risks in conventional batch or semi-batch reactors with oxidant-organic reactant mixtures are another particular challenge, since such reactions are generally exothermic and performed at elevated temperatures and pressures.7, 8, 12, 157

With the development of microreaction technology, microchannel reactors provide an excellent opportunity to address the above issues, and a variety of gas-liquid and gas-liquid-solid microchannel reactors has been developed so far.14, 16, 62, 158, 159 Among these, membrane microchannel reactors are an emerging type of reactors for process intensification, which combine the benefits of both membrane reactors and microchannel reactors.103, 104, 106, 109-112 The integration of membrane into a microchannel reactor forces gas and liquid phases to flow separately with a well-defined contacting interface. This allows better control of gas dosing and avoids direct mixing of gas and liquid. Simultaneously, the high surface-area-to-volume ratio in microchannel reactors could also contribute to the enhanced heat and mass transfer. These advantages make membrane microchannel reactors highly promising for intrinsically safe operation of aerobic alcohol oxidation.

As for the choice of available membranes, Teflon AF-2400 membrane has gained
popularity, since it uniquely combines excellent chemical resistance, thermal 
stability, and mechanical properties with high fractional free volume. Its 
permeability characteristics to gas and liquid also assure high flux of gas and low 
pervaporation of liquid through the membrane. The Ley group first developed a 
Teflon AF-2400 tube-in-tube membrane microreactor and applied it to key C-C, C-N, 
and C-O bond forming and hydrogenation reactions. In our previous work, Au- 
Pd/TiO$_2$ catalyst particles were directly packed in the tube-in-tube membrane 
microreactor and used for aerobic oxidation of benzyl alcohol. The continuous 
supply of oxygen through the membrane during the reaction contributed to 
significantly improved conversion and selectivity as compared to a reactor operating 
with an oxygen pre-saturated feed. However, the performance of the membrane 
microreactor still lacks deep understanding, due to the complex reaction kinetics of 
benzyl alcohol oxidation on Au-Pd/TiO$_2$ catalysts.

Ru-based catalysts have been investigated for oxidation of alcohols, featuring low 
activity but high selectivity. The selectivity to benzaldehyde in oxidation of 
benzyl alcohol on Ru-based catalyst is commonly reported to be $>99\%$ under 
moderate reaction conditions. In this study, a flat AF-2400 membrane 
microchannel reactor was experimentally and theoretically investigated for aerobic 
oxidation of benzyl alcohol with Ru/Al$_2$O$_3$ catalyst. The stability of the Ru/Al$_2$O$_3$ 
catalyst was studied and the kinetics of the benzyl alcohol oxidation was evaluated 
with a developed 2D reactor model, in an attempt to better understand the mass 
transfer and catalytic reactions in the membrane microchannel reactor and provide 
guidance for its improvement.
7.2 Experimental

7.2.1 Catalyst Preparation*

Ruthenium chloride (Johnson Matthey, 40.34% Ru), sodium hydroxide (Alfa Aesar) and gamma-aluminium oxide (γ-Al₂O₃, Sasol HP14-150) were all used as received. The 5 wt% Ru/Al₂O₃ catalyst was prepared by deposition-precipitation. The γ-Al₂O₃ support was suspended in water at 338 K to form a slurry. To this slurry, two solutions were added: 0.03 M NaOH and 0.05 M RuCl₃. The pH of the slurry was maintained at pH 9 by addition of appropriate amounts of these two solutions. Once addition of the ruthenium salt was complete, the temperature of the slurry was raised to boiling, and was then allowed to cool. The catalyst was separated by filtration, and the filtrate was washed thoroughly with hot water. Finally, the catalyst was dried at 373 K overnight. The ruthenium content was analysed by Inductively Coupled Plasma (ICP) analysis and found to be 4.46 wt% Ru. The powder was pelletized, crushed, and sieved to the desired particle size range (90-125 μm). The physical properties of the catalyst particles were characterized with QUADRASORB evo™ 4 BET Stations and Pentapyc™ 5200e Gas Pycnometer. The surface area was found to be 142 m²/g, the pore radius 8.1 nm and pore volume 0.72 cm³/g. The skeletal density of the catalyst was 3.38 g/cm³, from which the catalyst porosity was calculated to be 0.71.

* Catalyst preparation and ICP analysis were carried out by Dr Peter Ellis, Johnson Matthey. Physical properties of the catalyst were characterized by Dr Han Wu, UCL.
7.2.2 Flat Membrane Microchannel Reactor Set-up

As shown in Figure 7-1, the reactor consists of liquid and gas flow plates machined in 316 stainless steel (channel size length: 75 mm; width: 3 mm; depth: 1 mm), between which Teflon AF-2400 membrane (length: 85 mm; width: 30 mm; thickness: 0.07 mm; Biogeneral), a 304 stainless steel mesh (length: 85 mm; width: 30 mm; thickness: 0.05 mm; hole size: 76 μm; open area: 23%; Industrial Netting), gasket (length: 85 mm; width: 30 mm; thickness: 1 mm; open area length: 75 mm; width: 3 mm; Altec) were assembled. An O-ring groove was machined in the liquid flow plate, which allowed other meshes (e.g. sintered metal) also to be applied as the membrane support. Catalyst particles (particle size: 90-125 μm; 100 mg) were retained at the end of the liquid channel by a small circular piece of another nickel mesh (diameter: ~2 mm; thickness: 0.05 mm; hole size: 25μm; Tecan). The length of the catalyst bed in the liquid channel was measured to be ~5 cm. Hence, the void fraction of the catalyst bed was estimated to be 0.32.

The reactor was heated with a home-made hotplate fitted with a thermocouple, and the actual temperature inside the reactor was measured through the inserted
thermocouples (1 mm away from the liquid channel). To decrease heat loss, an insulation cap (insulation thickness: ~2 cm; WDS® Ultra, Morgan) was made to cover the reactor. The temperature differences between the gas and the liquid flow plates during the reaction at 373 K were measured to be less than 3 K.

Neat benzyl alcohol (1-10 μL/min, 99.0%, Sigma-Aldrich) was delivered into the reactor with a HPLC pump (Knauer P2.1S). The pressure of the liquid was controlled with an adjustable back pressure regulator (BPR, Zaiput, BPR-01) and measured by a pressure sensor (Zaiput, Hastelloy/PFA wetted parts). Pure oxygen (BOC, grade N6.0), 5 mL/min (at standard temperature and pressure), regulated by a mass flow controller (Brooks, GF40 series)) was directed to the gas channel to ensure the organic vapour concentration in the gas phase lower than 1 vol% (the lower explosive limit for toluene in air at 6 bar and 393 K\textsuperscript{156}). The maximum pervaporation of benzyl alcohol and benzaldehyde through the Teflon AF-2400 membrane was experimentally measured and shown in the Appendix C. A BPR (Swagelok, K series) was connected at the gas outlet to maintain the gas pressure. The actual gas pressure was monitored by a pressure sensor (Zaiput, Hastelloy/PFA wetted parts). The effluent from the liquid outlet was collected in a 2 mL vial (placed in an ice-water cold trap) and quantitatively analysed by a gas chromatograph (Agilent 7820A) fitted with a DB-624 capillary column and a flame ionization detector. For each set of experiment, at least three samples were collected and the results were averaged. The errors for the conversion and benzaldehyde selectivity were less than ±2 %.

Benzyl alcohol conversion ($X$) and selectivity ($S_B$) to benzaldehyde were calculated according to the following equations:
\[ X = \frac{c_{\text{BnOH, in}} - c_{\text{BnOH, in}}}{c_{\text{BnOH, in}}} \times 100\% \]  

(7-1)

\[ S_B = \frac{c_B}{c_{\text{BnOH, in}}} \times 100\% \]  

(7-2)

where \( c_{\text{BnOH, in}} \) and \( c_{\text{BnOH, out}} \) are the concentration of benzyl alcohol at the reactor inlet and outlet, respectively. \( c_B \) is the concentration of benzaldehyde at the outlet.

Catalyst contact time (CCT) was used to characterise the reaction time of benzyl alcohol, and was defined as

\[ CCT = \frac{m_{\text{cat}}}{\nu \cdot \rho} \]  

(7-3)

where \( m_{\text{cat}} \) is the mass of catalyst, \( \nu \) is the volumetric flow rate of benzyl alcohol at the inlet, \( \rho \) is the density of benzyl alcohol.

Turnover frequency (TOF) was calculated to quantitatively represent the reaction rate

\[ TOF = \frac{F_{\text{BnOH}} \cdot X}{n_{\text{Ru}}} \]  

(7-4)

where \( F_{\text{BnOH}} \) is the molar flow rate of benzyl alcohol at the inlet and \( n_{\text{Ru}} \) is the moles of Ru contained in the catalyst bed.

### 7.3 Mathematical Model

![Figure 7-2 Membrane reactor model used for the simulations.](image)
A 2D reactor model was developed to simulate the mass transfer and catalytic reaction in the reactor based on the following assumptions: 1) steady-state and isothermal conditions (the temperature profiles in the reactor were simulated by including an energy balance in the model, and nearly isothermal conditions were present in the reactor (details can be found in the Appendix C)); 2) Henry’s law applies at the membrane-liquid interface; 3) ideal gas behavior is valid for the gas phase; 4) uniform fluid velocity across the cross-section of the liquid channel with constant physical properties and transport coefficients (catalyst bed packing was achieved with sieved catalyst pellets and a vacuum pump sucking at the outlet of the reactor, and thus uniform catalyst bed packing was assumed and an ideal plug-flow packed bed reactor model was used); 5) negligible dissolution of liquid in the membrane and pervaporation to the gas phase; 6) negligible pressure drop along the catalyst bed (the pressure drop was measured to be less than 0.1 bar during the experiment). Since the oxygen pressure in the gas phase was constant at given oxygen pressure, only the liquid channel and membrane were considered with boundary condition of constant oxygen concentration at the interface of gas and membrane.

7.3.1 Mass Balance in the Reactor

The simulation of the liquid phase was realized through multiscale coupling of the catalyst particles with the bulk liquid phase. The continuity equation for species in the unpacked zone can be expressed as:

\[ u_x \frac{\partial c_i}{\partial x} - D_i \left( \frac{\partial^2 c_i}{\partial x^2} + \frac{\partial^2 c_i}{\partial y^2} \right) = 0 \]  

(7-5)

where \( i \) stands for oxygen or benzyl alcohol, \( c_i \) is the concentration of \( i \) in the bulk
liquid phase, $D_i$ is the molecular diffusion coefficient in the bulk liquid, $u_x$ is the liquid velocity in the axial direction.

The continuity equation for species in the catalyst bed is
\[
u_x \frac{\partial c_i}{\partial x} - \left( D_{i,A} \frac{\partial^2 c_i}{\partial x^2} + D_{i,T} \frac{\partial^2 c_i}{\partial y^2} \right) + J_i A_b = 0 \quad (7-6)\]

where $D_{i,A}$ is axial dispersion coefficient, $D_{i,T}$ is transverse dispersion coefficient, $J_i$ is the molar flux of $i$ into the catalyst particles, $A_b$ is the specific surface area exposed to the liquid of the packed bed, assuming randomly packed spherical particles.$^{160}$

\[
A_b = \frac{3}{r_p} (1 - \varepsilon_b) \quad (7-7)
\]

where $r_p$ is the average radius of catalyst particles, $\varepsilon_b$ is catalyst bed void fraction.

The external mass transfer resistance is expressed in terms of mass transfer coefficient
\[
J_i = h_i (c_i - c_{i,ps}) \quad (7-8)
\]

\[
h_i = \frac{S_h D_i}{2r_p} \quad (7-9)
\]

\[
S_h = 2 + 0.552 Re^{1/2} Sc^{1/3} \quad (7-10)
\]

\[
Sc = \frac{\mu}{\rho \cdot D_i} \quad (7-11)
\]

\[
Re = \frac{2r_p \cdot \rho \cdot u_x}{\mu} \quad (7-12)
\]

where $c_{i,ps}$ is the concentration of $i$ at the catalyst particle surface, $h_i$ is the external mass transfer coefficient. $Sh$ is Sherwood number, which is calculated using the Frössling correlation.$^{161}$ $Sc$ is Schmidt number, $\mu$ is viscosity of benzyl alcohol, $\rho$ is the density of benzyl alcohol, $Re$ is particle Reynolds number.
The simulation of the reaction inside the catalyst particles was realized in COMSOL® using the Reaction Pellet Bed feature. This feature has a predefined extra dimension (1D) on the dimensionless catalyst particle radius of the catalyst particle. The mass balance inside the catalyst particles is derived from a shell balance across a spherical shell

\[ \frac{\partial}{\partial r} (r^2 D_{i,\text{eff}} \frac{\partial c_{i,p}}{\partial r}) = r^2 r_p^2 R_{i,p} \]  

(7-13)

where \( r \) is a dimensionless catalyst particle radius, \( D_{i,\text{eff}} \) is the effective diffusion coefficient of \( i \) in the catalyst pores, \( c_{i,p} \) is the concentration of \( i \) in the catalyst particle, \( R_{i,p} \) is the reaction rate per unit volume of catalyst particle.

For Ru/Al\(_2\)O\(_3\) catalyzed aerobic oxidation of benzyl alcohol, the amount of oxygen consumed was observed to be half of the benzyl alcohol consumed, with a benzaldehyde selectivity > 99\%.\(^{40} \) So, the overall reaction is

\( \text{PhCH}_2\text{OH} + \frac{1}{2}\text{O}_2 \rightarrow \text{PhCHO} + \text{H}_2\text{O} \)  

(7-14)

and the kinetics of the oxidation reaction was modeled as a \( \alpha \)-order reaction in benzyl alcohol and a \( \beta \)-order reaction in oxygen

\[ R_{\text{BnOH},p} = -k c_{\text{BnOH},p}^\alpha c_{\text{O}_2,p}^\beta \]  

(7-15)

\[ R_{\text{O}_2,p} = -\frac{1}{2} k c_{\text{BnOH},p}^\alpha c_{\text{O}_2,p}^\beta \]  

(7-16)

where \( R_{\text{BnOH},p} \) and \( R_{\text{O}_2,p} \) are the reaction rate of benzyl alcohol and oxygen in the catalyst particle, \( k \) is the reaction rate coefficient, which is estimated based on the experimental results.

Boundary conditions used were:
where \( c_{\text{BnOH,in}} \) is the inlet concentration of benzyl alcohol, \( c_{O2,m} \) is the oxygen concentration in the membrane, \( H \) is the dimensionless Henry solubility of oxygen in benzyl alcohol at 373 K. Reactant conversion was calculated based on the average concentration of benzyl alcohol at the outlet.

### 7.3.2 Mass Balance in the Membrane

The steady-state mass balance for oxygen transfer in the membrane, which is considered to be by diffusion alone, can be expressed as:

\[
D_{O2,m} \left( \frac{\partial^2 c_{O2,m}}{\partial x^2} + \frac{\partial^2 c_{O2,m}}{\partial y^2} \right) = 0
\]

where \( c_{O2,m} \) is the oxygen concentration in the membrane, \( D_{O2,m} \) is the oxygen diffusion coefficient in the membrane.

Boundary conditions:

\[
\begin{align*}
\text{at } y = d_1, & \quad c_{O2,m} = H c_{O2} \\
\text{at } y = d_2, & \quad c_{O2,m} = c_{O2,g}
\end{align*}
\]

where \( c_{O2,g} \) is the oxygen concentration in the gas phase.

The mass balance equations together with the boundary conditions were solved
using COMSOL Multiphysics software version 5.2a. A mesh consisting of 33,000 elements and 530,007 degree of freedom was used and the results were found to be mesh independent. The values of the variables used in the model are shown in Table 1 and the calculation details are shown in the Appendix C.

**Table 7-1 List of parameter values used in the model**

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Value</th>
<th>Units</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>$c_{O_2,g}$</td>
<td>$P_{O_2}$</td>
<td>mol/m$^3$</td>
<td>Oxygen concentration in the gas phase</td>
</tr>
<tr>
<td>$c_{BnOH,in}$</td>
<td>9621</td>
<td>mol/m$^3$</td>
<td>Benzyl alcohol inlet concentration</td>
</tr>
<tr>
<td>$d_1$</td>
<td>1.00</td>
<td>mm</td>
<td>Reactor liquid channel depth</td>
</tr>
<tr>
<td>$d_2$</td>
<td>1.07</td>
<td>mm</td>
<td>Reactor liquid channel depth plus membrane thickness</td>
</tr>
<tr>
<td>$D_{BnOH}$</td>
<td>$2.01 \times 10^{-9}$</td>
<td>m$^2$/s</td>
<td>Diffusion coefficient of benzyl alcohol in the liquid at 373 K$^{162,163}$</td>
</tr>
<tr>
<td>$D_{O_2}$</td>
<td>$5.19 \times 10^{-9}$</td>
<td>m$^2$/s</td>
<td>Diffusion coefficient of $O_2$ in benzyl alcohol at 373 K$^{162,164}$</td>
</tr>
<tr>
<td>$D_{BnOH,A}$</td>
<td>$0.46 \times 10^{-9}$</td>
<td>m$^2$/s</td>
<td>Axial dispersion coefficient of benzyl alcohol*</td>
</tr>
<tr>
<td>$D_{BnOH,T}$</td>
<td>$0.46 \times 10^{-9}$</td>
<td>m$^2$/s</td>
<td>Transverse dispersion coefficient of benzyl alcohol*</td>
</tr>
<tr>
<td>$D_{O_2,A}$</td>
<td>$1.19 \times 10^{-9}$</td>
<td>m$^2$/s</td>
<td>Axial dispersion coefficient of oxygen*</td>
</tr>
<tr>
<td>$D_{O_2,T}$</td>
<td>$1.19 \times 10^{-9}$</td>
<td>m$^2$/s</td>
<td>Transverse dispersion coefficient of oxygen*</td>
</tr>
<tr>
<td>$D_{BnOH,eff}$</td>
<td>$1.01 \times 10^{-9}$</td>
<td>m$^2$/s</td>
<td>Effective diffusion coefficient of benzyl alcohol in the catalyst particle$^{165}$</td>
</tr>
<tr>
<td>$D_{O_2,eff}$</td>
<td>$2.62 \times 10^{-9}$</td>
<td>m$^2$/s</td>
<td>Effective diffusion coefficient of oxygen in the catalyst particle$^{165}$</td>
</tr>
<tr>
<td>$D_{O_2,m}$</td>
<td>$1.19 \times 10^{-9}$</td>
<td>m$^2$/s</td>
<td>Diffusion coefficient of oxygen in the membrane at 373K$^{114,119}$</td>
</tr>
<tr>
<td>$H$</td>
<td>9.9</td>
<td>-</td>
<td>Dimensionless Henry solubility of oxygen in benzyl alcohol at 373 K</td>
</tr>
<tr>
<td>Symbol</td>
<td>Value</td>
<td>Unit</td>
<td>Description</td>
</tr>
<tr>
<td>--------</td>
<td>-------</td>
<td>------</td>
<td>-------------</td>
</tr>
<tr>
<td>$l_1$</td>
<td>25</td>
<td>mm</td>
<td>Length of the unpacked zone in the liquid channel</td>
</tr>
<tr>
<td>$l_2$</td>
<td>50</td>
<td>mm</td>
<td>Length of the catalyst bed</td>
</tr>
<tr>
<td>$M_{\text{BnOH}}$</td>
<td>108.1</td>
<td>g/mol</td>
<td>Molecular weight of benzyl alcohol</td>
</tr>
<tr>
<td>$P_{O_2}$</td>
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<td>Oxygen pressure in the gas phase</td>
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<tr>
<td>$r_p$</td>
<td>54</td>
<td>μm</td>
<td>Average radius of catalyst particles</td>
</tr>
<tr>
<td>$R$</td>
<td>$8.314 \times 10^{-9}$</td>
<td>m$^3$-bar/(K-mol)</td>
<td>Ideal gas constant</td>
</tr>
<tr>
<td>$u_x$</td>
<td>$\frac{v}{wd_1}$</td>
<td>cm/s</td>
<td>Axial liquid superficial velocity vector</td>
</tr>
<tr>
<td>$w$</td>
<td>3</td>
<td>mm</td>
<td>Liquid channel width</td>
</tr>
<tr>
<td>$\varepsilon_p$</td>
<td>0.71</td>
<td>-</td>
<td>Catalyst particle porosity</td>
</tr>
<tr>
<td>$\varepsilon_b$</td>
<td>0.32</td>
<td>-</td>
<td>Catalyst bed void fraction</td>
</tr>
<tr>
<td>$\mu$</td>
<td>$7.49 \times 10^{-4}$</td>
<td>Pa·s</td>
<td>Viscosity of benzyl alcohol at 373 K$^{166}$</td>
</tr>
<tr>
<td>$\rho$</td>
<td>1.0</td>
<td>g/mL</td>
<td>Density of benzyl alcohol at 373 K$^{167}$</td>
</tr>
<tr>
<td>$v$</td>
<td>1-10</td>
<td>μL/min</td>
<td>Liquid flow rate at the inlet</td>
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</table>

*The dispersion coefficient was calculated based on the molecular diffusion coefficient, due to the low Re (the details are shown in the Appendix C).$^{168}$
7.4 Results and Discussion

7.4.1 Stability of Ru/Al2O3 Catalyst

![Conversion or selectivity graph](image)

*Figure 7-3 Stability study of the Ru/Al2O3 catalyst in aerobic oxidation of benzyl alcohol.*

Reaction conditions: Ru/Al2O3 catalyst (90-125 µm), 100 mg; neat benzyl alcohol, 2 µL/min; catalyst contact time, 48 gcat.min/gBnOH; oxygen pressure, 8.0 bar; liquid pressure, 9.0 bar; reaction temperature, 373 K. X, conversion of benzyl alcohol; S_B, selectivity to benzaldehyde.

Initially, the performance of the Ru/Al2O3 catalyst was experimentally evaluated in the flat membrane microchannel reactor, and the results are shown in Figure 7-3. A fast deactivation process was observed in the first ~ 20 h. Only 5% conversion drop was observed in the following 25 h. Mannel et al.38 also observed a fast deactivation rate followed by a slower one in a catalyst packed stainless steel tube reactor, which was attributed to poisoning of active sites by benzoic acid. Brazier et al.11 further derived an equation for the effect of oxygen on Ru/Al2O3 catalyst deactivation, which accounted for a fast and reversible inhibition by the formed benzoic acid, combined with a slower and irreversible loss of active sites caused by Ru reduction. Since the selectivity to benzaldehyde stabilized at 99% within the 45 h,
100% selectivity to benzaldehyde was assumed in the model without considering any other minor by-products.

For the reaction kinetics, Yamaguchi et al.\textsuperscript{40} studied the kinetics and mechanism of alcohol oxidation on Ru/Al\textsubscript{2}O\textsubscript{3} catalyst in a batch reactor (Table 7-2, entry 1). The reaction rate was indicated to be of fractional order with the benzyl alcohol (first-order at low concentrations and zero-order at high concentrations; alcohol concentration range: 0-2 M) and a zero order on the oxygen pressure (0.2-3.0 bar). Using a multichannel packed bed reactor, Bavykin et al.\textsuperscript{41} observed that the yield of benzaldehyde in oxidation of benzyl alcohol (1 M) on Ru/Al\textsubscript{2}O\textsubscript{3} catalyst was affected by the oxygen pressure at pressure lower than 9 bar and independent of the oxygen pressure at pressures above 9 bar (Table 7-2, entry 2). Zotova et al.\textsuperscript{11} studied aerobic oxidation of alcohols to aldehydes and ketones on Ru/Al\textsubscript{2}O\textsubscript{3} catalyst in a commercially-available XCube\textsuperscript{TM} reactor by pre-mixing and saturating the liquid with the gaseous reactant before reaching the catalyst bed (Table 7-2, entry 3). The oxidation of benzyl alcohol was indicated to be a pseudo-first-order with respect to benzyl alcohol (up to 0.2 M), and oxygen was found to be integral to the kinetics of the reaction, since the reaction rate was significantly enhanced by increasing the oxygen pressure from 6 to 26 bar. In a following study, the reaction rate was re-examined in a plug flow differential reactor with an inline FTIR, and was found to be of zero order with benzyl alcohol (0.1 M) under reaction conditions not subject to mass transfer limitations, and a partial positive order in oxygen (oxygen saturation pressure up to 26 bar) (Table 7-2, entry 4).\textsuperscript{42} The different observed kinetics was probably caused by the different ratios of oxygen concentration to benzyl alcohol concentration in different types of reactors.
According to the above kinetics study, different reaction rates were assumed in the model and the average conversion at the stable period in Figure 7-3 (40%, 20-40 h) was used to estimate the corresponding reaction rate coefficient (k), which made the modelling conversion equal to the experimental value under the same reaction conditions (shown in Appendix C). The estimated k was then used for the reaction rate under other reaction conditions (e.g. different liquid flow rates and oxygen pressures). The modelling results obtained with different reaction rates are compared in Appendix C. Since the oxygen concentration likely varied axially and transversely in the catalyst bed, which resulted in varied ratios of oxygen concentration to benzyl alcohol concentration, a reaction rate assuming a first order in benzyl alcohol and a 0.5 order in oxygen was used in the following modelling with reaction rate constant k of $3.9 \times 10^{-4} \text{ m}^{3/2}/(\text{mol}^{1/2}\cdot\text{s})$ (Table 7-2, entry 5).
CHAPTER 7. Aerobic Oxidation of Benzyl Alcohol on Ru/Al2O3 Catalyst in a Flat Membrane Microchannel Reactor: an Experimental and Modelling Study

Table 7-2 Comparison of reaction kinetics under different reaction conditions

<table>
<thead>
<tr>
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<tbody>
<tr>
<td></td>
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<td></td>
<td></td>
<td></td>
<td></td>
<td>Benzyl alcohol</td>
<td></td>
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<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>Oxygen</td>
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</tr>
<tr>
<td>1</td>
<td>1.4</td>
<td>0.2-2</td>
<td>0.2-3.0</td>
<td>batch</td>
<td>310-356</td>
<td>fractional</td>
<td>zero</td>
</tr>
<tr>
<td>2</td>
<td>0.9</td>
<td>1</td>
<td>5-25 (trickle)</td>
<td>packed bed</td>
<td>353-408</td>
<td>-</td>
<td>fractional</td>
</tr>
<tr>
<td>3</td>
<td>5</td>
<td>0.1-0.2</td>
<td>6 or 26 (presaturated)</td>
<td>packed bed</td>
<td>363</td>
<td>pseudo-first</td>
<td>fractional</td>
</tr>
<tr>
<td>4</td>
<td>5</td>
<td>0.1</td>
<td>6 or 26 (presaturated)</td>
<td>packed bed</td>
<td>393-453</td>
<td>zero</td>
<td>fractional</td>
</tr>
<tr>
<td>5</td>
<td>4.5</td>
<td>9.6</td>
<td>1-8 (membrane)</td>
<td>packed bed</td>
<td>373</td>
<td>first</td>
<td>0.5</td>
</tr>
</tbody>
</table>
7.4.2 Effect of Liquid Flow Rate

![Graph showing the effect of liquid flow rate on the conversion of benzyl alcohol and TOF](image)

**Figure 7-4** Effect of liquid flow rate on the conversion of benzyl alcohol and TOF, obtained from experiment and modelling. Reaction conditions: Ru/Al$_2$O$_3$ catalyst (90-125 μm), 100 mg; neat benzyl alcohol; oxygen pressure, 8 bar; liquid pressure, 9.0 bar; reaction temperature, 373 K.

The effect of liquid flow rate on the conversion of benzyl alcohol was modelled under the same oxygen pressure (8 bar), and the modelling results are compared with the experimental values in Figure 7-4. Good agreement was observed between the modelling and the experimental results under different liquid flow rates. Since the reaction rate was estimated only based on results under liquid flow rate of 2 μL/min, the good agreement shows that the model based on the estimated kinetics provided satisfactory prediction of reactor performance.

In Figure 7-4, high conversion was obtained at low liquid flow rate. Since the length of catalyst bed was constant (due to the same amount of catalyst), the enhanced conversion was due to the longer time for oxygen permeation and reaction.\textsuperscript{101,147} Theoretically, full conversion of benzyl alcohol could be achieved at liquid flow rate of ~0.4 μL/min. This corresponds to a catalyst contact time of 240
Average TOF was calculated to better compare the reaction rates at different liquid flow rates. From Figure 7-4, the profile of TOF was observed to increase dramatically with liquid flow rate increasing and then reach a plateau at liquid flow rate higher than 2 μL/min. Notably, the TOF was ~11 h⁻¹ at liquid flow rate higher than 2 μL/min and 8 bar oxygen pressure, which corresponded to an oxygen consumption rate of $6.7 \times 10^{-7}$ mol/(gₜₐₜ·s). These values are comparable to previous work where Ru/Al₂O₃ was used in a packed bed reactor with diluted oxygen (~6 h⁻¹ at 353 K and 11 bar 8% oxygen in nitrogen)³⁸ or oxygen pre-saturated flow (13 h⁻¹ at 373 K and 26 bar oxygen pre-saturation pressure)⁴², and a PTFE membrane reactor (2 h⁻¹ at 353 K and 10.2 bar oxygen pressure)¹¹³. However, the TOF observed in the membrane reactor is lower than those obtained in a batch reactor with bubbled oxygen (63 h⁻¹ at 356 K and 1 bar)³⁷ and a multichannel packed bed reactor with high flow rate of pure oxygen (~110 h⁻¹ at 375 K and 25 bar pressure)⁴¹.

7.4.3 Effect of Oxygen Pressure

Figure 7-5 presents the effect of oxygen pressure on the conversion of benzyl alcohol in the flat membrane microchannel reactor. Model prediction agrees well with the experimental results. The conversion of benzyl alcohol varied approximately linearly with the oxygen pressure and reached ~70% at 8 bar. A linear trend was also observed in the Teflon tube-in-shell membrane reactors with homogeneous Cu/TEMPO catalyst¹¹³ and a Teflon AF-2400 tube-in-tube membrane reactor with heterogeneous Au-Pd/TiO₂ catalyst¹⁴⁷. Since the increase of oxygen pressure could increase the oxygen solubility in the liquid, the oxygen pressure-dependent conversion demonstrates that oxygen has a positive effect on the reaction
rate under the conditions investigated. So, the lower TOF of Ru/Al₂O₃ catalyst in the membrane reactor as compared to those in a batch reactor³⁷ and a packed bed reactor with pure oxygen⁴¹, was probably caused by the high loading of Ru in this study, as well as insufficient oxygen supply in the membrane reactor.

![Figure 7-5](image)

**Figure 7-5** Effect of oxygen pressure on the conversion of benzyl alcohol, obtained from experiment and modelling. Reaction conditions: Ru/Al₂O₃ catalyst (90-125 μm), 100 mg; neat benzyl alcohol, 1 μL/min; catalyst contact time, 96 g·cat·min/g_BnOH; liquid pressure, 1.0 bar higher than oxygen pressure; reaction temperature, 373 K.
7.4.4 Effect of Membrane Thickness

Figure 7-6 Effect of membrane thickness on the conversion of benzyl alcohol, obtained from modelling. The square indicates experimental point. Reaction conditions: Ru/Al₂O₃ catalyst (90-125 μm), 100 mg; neat benzyl alcohol, 2 μL/min; catalyst contact time, 48 gcat·min/gBnOH; oxygen pressure, 8.0 bar; reaction temperature, 373 K.

Having now a validated membrane reactor model, we proceed to investigate the effect of various parameters in order to better understand reactor performance and provide guidance for its improvement. The effect of membrane thickness was studied using the model and the results are shown in Figure 7-6. Less than 4% increase in the conversion of benzyl alcohol was observed when decreasing the membrane thickness from 0.10 mm to 0.01 mm. These results indicate that the oxygen transfer resistance in the membrane has no significant effect on the reactor performance, probably due to the high permeability of the Teflon AF-2400 membrane. Indeed, the oxygen diffusion coefficient in the membrane was in the same order of magnitude as the oxygen transverse dispersion coefficient or the oxygen effective diffusion coefficient in the catalyst particles, but the oxygen diffusion coefficient in the membrane was based on the oxygen concentration in the gas phase. To compare the oxygen
diffusion coefficients based on the oxygen concentration in the liquid phase, a hypothetical oxygen concentration in the liquid phase \(c_{O_2,g}'\) was assumed at the gas-liquid interface, which was in thermal equilibrium with oxygen concentration in the gas phase \(c_{O_2,g}\). 

\[
c'_{O_2,g} = \frac{c_{O_2,g}}{H} \tag{7-25}
\]

The oxygen flux at the gas-liquid interface

\[
D_{O_2,T} \frac{\partial c_{O_2}}{\partial y} = D_{O_2,m} \frac{\partial c_{O_2,g}}{\partial y} = D_{O_2,m} \cdot H \frac{\partial c_{O_2,g}'}{\partial y} = D_{O_2,m}' \frac{\partial c_{O_2,g}'}{\partial y} \tag{7-26}
\]

So, \(D_{O_2,m}'\) could be assumed as a hypothetical oxygen diffusion coefficient in the membrane based on the oxygen concentration in the liquid phase, which would be one order of magnitude higher than the oxygen transverse dispersion coefficient or the oxygen effective diffusion coefficient in the catalyst particles. This resulted to the negligible effect on the reactant conversion when further decreasing the membrane thickness (e.g. enhancing the oxygen mass transfer in the membrane).

### 7.4.5 Effect of Liquid Channel Depth

Next, the effect of liquid channel depth on the conversion of benzyl alcohol was studied with constant catalyst contact time, and the results are shown in Figure 7-7. An approximately linear increase in conversion was observed when decreasing the liquid channel depth. At a liquid channel depth of 0.1 mm, the conversion reached 100%, which corresponded to a TOF of 26 h\(^{-1}\). These results demonstrate a significant effect of the liquid channel depth on the conversion of benzyl alcohol and the TOF of the Ru/Al\(_2\)O\(_3\) catalyst.
CHAPTER 7. Aerobic Oxidation of Benzyl Alcohol on Ru/Al2O3 Catalyst in a Flat Membrane Microchannel Reactor: an Experimental and Modelling Study

Figure 7-7 Effect of liquid channel depth on the conversion of benzyl alcohol, obtained from modelling. The square indicates experimental point. Reaction conditions: Ru/Al2O3 catalyst (90-125 µm), 10-100 mg; neat benzyl alcohol, 0.2-2 µL/min; catalyst contact time, 48 \( \frac{g_{cat} \cdot min}{g_{BnOH}} \); oxygen pressure, 8 bar; reaction temperature, 373 K.

To better understand the effect of liquid channel depth, the concentration maps of benzyl alcohol and oxygen in the bulk liquid phase within the reactor are presented. For benzyl alcohol (shown in Figure 7-8), the bulk concentration decreased gradually along the axial direction, and at the same axial position, slightly lower concentration was observed in the catalyst bed close to the membrane. No obvious concentration difference was found between the bulk phase and the catalyst particle surface (see Figure 7-8c), as well as within the catalyst particle (see Figure C-5). However, oxygen was observed to exist only in a thin layer close to the membrane, and no oxygen was supplied in the main part of the catalyst bed (shown in Figure 7-9a). Specifically, the oxygen concentration at the same axial position dropped dramatically with \( y \) decreasing, and was equal to zero at \( y < 0.7 \) mm (see Figure 7-9b). Hence, the main part of the packed catalyst is unutilized resulting in low conversion and average TOF.
In Figure 7-9c, the oxygen concentrations in the bulk liquid are compared with those at the catalyst particle surface along the axial direction at different transverse positions. Only slight concentration difference (~0.8 mol/m³, < 10% of the oxygen concentration in the bulk liquid phase), which was caused by the external mass transfer resistance, was observed at the catalyst bed close to the membrane (y = 0.95 mm) where relatively high concentration of oxygen existed. Since the external mass transfer was affected by the liquid flow rate and the TOF nearly stabilized at liquid flow rate > 2 μL/min (shown in Figure 7-4), it seems that the external mass transfer resistance does not influence the oxygen transfer in the membrane microchannel reactor.
CHAPTER 7. Aerobic Oxidation of Benzyl Alcohol on Ru/Al2O3 Catalyst in a Flat Membrane Microchannel Reactor: an Experimental and Modelling Study

Figure 7-8 Benzyl alcohol concentration map and profiles in the bulk liquid phase within the reactor, obtained from modelling. a) Benzyl alcohol concentration map in the reactor, b) benzyl alcohol concentration profiles across the transverse direction at different axial positions, c) benzyl alcohol concentration profiles along the axial direction at different transverse positions. Reaction conditions: Ru/Al2O3 catalyst (90-125 μm), 100 mg; neat BnOH, 2 μL/min; catalyst contact time, 48 gcat·min/gBnOH; oxygen pressure, 8 bar; reaction temperature, 373 K. cBnOH: benzyl alcohol concentration in the bulk liquid, cBnOH,ps: benzyl alcohol concentration at the catalyst particle surface.
Figure 7-9 Oxygen concentration map and profiles in the bulk liquid phase within the reactor, obtained from modelling. a) Oxygen concentration map in the reactor, b) oxygen concentration profiles across the transverse direction at different axial positions, c) oxygen concentration profiles along the axial direction at different transverse positions. Reaction conditions: Ru/Al₂O₃ catalyst (90-125 μm), 100 mg; neat BnOH, 2 μL/min; catalyst contact time, 48 gₙₒₙ·min/g_BnOH; oxygen pressure, 8 bar; reaction temperature, 373 K. cₒ₂: oxygen concentration in the bulk liquid, cₒ₂ₛ: oxygen concentration at the catalyst particle surface.
The oxygen concentration profiles in the catalyst particles, which was caused by the internal mass transfer resistance, was less than 10% of the oxygen concentration at the particle surface even for the particles close to the membrane ($y = 0.95$ mm). This was comparable to that caused by the external mass transfer resistance. When halving the particle size and assuming the same bed properties in the model, the conversion was observed to increase by only ~1 %, indicating negligible internal mass transfer resistance.

Note that in the laminar flow regime, the oxygen transfer in the packed bed of the membrane reactor includes transverse mass transfer through the bulk liquid in the catalyst bed, diffusion through the “stagnant film” at the catalyst particle surface and then within the catalyst particle pores. The liquid channel depth has presented a
significant effect on the reactant conversion in Figure 7-7. Since the external/internal mass transfer resistance in the membrane reactor had no obvious effect on the conversion, it seems that the significant effect of liquid channel depth is caused by the oxygen transverse mass transfer resistance in the catalyst bed. In the laminar flow regime, the oxygen transverse mass transfer rate is inversely proportional to the liquid channel depth, so an approximately linear increase in conversion was observed when decreasing the liquid channel depth. Halving the liquid channel depth from 1.0 to 0.5 mm led to conversion of benzyl alcohol increasing from 40% to 70% (see Figure 7-7). Correspondingly, the limited oxygen supply in the catalyst bed was also improved with oxygen supply at \( y > 0.2 \) mm (shown in Figure C-6). These results suggest that the reactor performance could be improved through decreasing the liquid channel depth.

### 7.4.6 Effect of Reaction Rate Coefficient

The effect of the reaction rate coefficient (\( k \)) on the conversion of benzyl alcohol was also investigated. From Figure 7-10, the conversion of benzyl alcohol was found to gradually rise with \( k \) increasing, and reached 70% at \( k = 2 \times 10^{-3} \text{ m}^{3/2}/(\text{mol}^{1/2} \cdot \text{s}) \).

Since the oxygen consumed was supplied through the membrane, the gradual increase in the conversion indicates that the oxygen supply through the membrane could be enhanced at high reaction rate coefficient. Theoretically, the maximum oxygen flux through the membrane at 8 bar oxygen pressure could be \( 4.4 \times 10^{-7} \text{ mol}/(\text{cm}^2 \cdot \text{s}) \). However, the oxygen transverse mass transfer has been the controlling process with \( k = 3.9 \times 10^{-4} \text{ m}^{3/2}/(\text{mol}^{1/2} \cdot \text{s}) \), further increasing the reaction rate coefficient would increase the oxygen demand by the catalyst, and thus cause more serious oxygen starvation in the catalyst bed and result to lower catalyst utilization.
In order to fully utilize the catalyst packed in the membrane reactor, decreasing the reaction rate coefficient through decreasing the metal loading or choosing other less active catalyst can be considered. Alternatively, larger contribution of convective dispersion in the transverse direction in the catalyst bed could be induced by increasing the liquid velocity, in an attempt to enhance the transverse dispersion coefficient, at the expense of increasing pressure drop.

![Graph](image)

**Figure 7-11** Effect of reaction rate coefficient on the conversion of benzyl alcohol, obtained from experiment and modelling. Reaction conditions: Ru/Al$_2$O$_3$ catalyst (90-125 μm), 100 mg; neat benzyl alcohol, 2 μL/min; catalyst contact time, 48 g$_{cat}$·min/g$_{BnOH}$; oxygen pressure, 8 bar; liquid pressure, 9.0 bar; reaction temperature, 373 K.

### 7.5 Conclusions

Aerobic oxidation of benzyl alcohol on Ru/Al$_2$O$_3$ catalyst was experimentally and theoretically investigated in a Teflon AF-2400 flat membrane microchannel reactor. The proposed model with the estimated kinetics was validated with the experimental data and provided a better understanding of the mass transfer and catalytic reactions in the membrane reactor. Through modelling study, the benzyl alcohol conversion
was indicated to be not affected by the membrane thickness, but significantly determined by the liquid channel depth. Further investigation on the benzyl alcohol and oxygen concentration gradients caused by external and internal mass transfer resistance indicated that the transverse oxygen mass transfer, rather than the oxygen external/internal mass transfer, was the controlling step for the membrane reactor performance. The conversion could also be increased by increasing the catalyst activity, but utilization of the highly active catalyst would be decreased due to the limitation of the oxygen transverse mass transfer. To simultaneously improve the catalyst utilization and reactor conversion, the transverse dispersion coefficient should be increased, possibly by increasing the liquid velocity combined with recycling or using a forced convection mechanism (oscillation, ultrasonic vibration, etc.).
CHAPTER 8. Heterogeneously Catalysed Aerobic Oxidation of Benzyl Alcohol in a Continuous Stirred Membrane Reactor System

A continuous stirred membrane reactor system was demonstrated for aerobic oxidation of benzyl alcohol with 5 wt% Ru/Al₂O₃ catalyst powder. The solid catalyst powder was retained inside the reactor with a sintered metal filter. A Teflon AF-2400 tube-in-tube contactor was used externally to the stirred reactor at room temperature for oxygen supply. Thus, the direct mixing of gaseous oxygen with organic reactants could be avoided inside the reactor, making the reactor intrinsically safe for aerobic oxidation. The oxygen supply could also be adjusted by changing the oxygen pressure in the contactor or the recycle flow rate. At 5 bar oxygen pressure with 0.5 g catalyst, 10 μL/min reactor system inlet flow rate and 1 mL/min recycle flow rate, the conversion of benzyl alcohol reached 13% with 94% selectivity to benzaldehyde. Through mathematical simulation of the membrane reactor system, the effect of the reactor system inlet flow rate, the oxygen pressure, the recycle flow rate and the reaction rate coefficient was investigated on the reactant conversion and the catalyst utilization. These parameters were also correlated to estimate the effective use of the catalyst in the stirred membrane reactor and guide the reactor design and operation.
8.1 Introduction

Flow reactors are becoming more common in chemistry labs with the advent of flow chemistry.\textsuperscript{8, 169} However, multiphase reactors using solid catalysts are still relatively rare since catalyst particles cannot be easily manipulated, especially in flow.\textsuperscript{170, 171} Normally, packed bed reactors can be used for catalysts with particle size typically in the order of 100 μm.\textsuperscript{7} However, for catalysts with particle size smaller than 50 μm, pressure drop would develop along the catalyst bed. Though catalyst pelletizing could be considered, this would increase the complexity in the lab research.

A promising method to use small solid catalyst particles in continuous flow is to employ a filtration unit to couple heterogeneous catalytic reaction in a slurry reactor.\textsuperscript{172-174} Solid catalyst can be \textit{in-situ} separated from the reactant mixture by the filter, which makes the process facile to be continuous. Jin and co-workers\textsuperscript{172} developed a continuous flow system for phenol hydroxylation using a submerged ceramic membrane as a filter in a stirred reaction vessel. The submerged membrane reactor was demonstrated to maintain a steady production and enhance the productivity as compared to a batch reactor or a fixed-bed reactor. In their following study, a dual-membrane reactor was demonstrated for continuous heterogeneous oxidation catalysis, in which one tubular porous ceramic membrane worked as a separator and the other as a distributor to control the oxidant supply (H\textsubscript{2}O\textsubscript{2}).\textsuperscript{173} A uniform injection of the oxidant was achieved, which contributed to a significant increase of the product selectivity. Blacker and co-workers\textsuperscript{175} developed a multiphasic continuous process for catalytic hydrodeoxygenation of phenols using continuous stir tank reactors, in which solid catalysts were maintained with an in-line
filter (the details on the filter were not provided).

Oxidation of alcohols is one of the most fundamental organic transformations and often performed with stoichiometric inorganic reagents, which are often not chemoselective and could raise seriously environmental problems.\textsuperscript{2, 5, 6} To improve the atom efficiency and reduce the environmental costs, heterogeneous catalysts have been developed for this type of reaction with molecular oxygen.\textsuperscript{3, 4, 12, 20, 60}

However, the safety issues caused by the direct gaseous oxidant-organic reactant mixing potentially limit the large-scale applications of aerobic oxidation of alcohols.\textsuperscript{7, 12, 38} Membrane microreactors allow a bubble-free organic reactant stream with a control of gas dosing.\textsuperscript{103, 104} A tube-in-tube membrane contactor with a semi-permeable Teflon AF-2400 membrane as inner tube has been developed in Ley’s group and applied to various heterogeneous reactions.\textsuperscript{118} Chaudhuri et al.\textsuperscript{124} used the tube-in-tube membrane contactor to saturate the liquid stream with oxygen/air for catalytic oxidation of alcohols in a packed bed reactor with gold doped MAGSILICA\textsuperscript{®} catalyst.

In this chapter, a stirred membrane reactor system was developed and evaluated for direct use of catalyst powder in continuous flow aerobic oxidation of alcohols. A sintered metal disk was used to maintain the solid catalyst powder inside the reactor, and a Teflon AF-2400 tube-in-tube membrane contactor was employed externally to the stirred reactor within a recycle loop, to avoid the safety issue with the mixing of gaseous oxygen and organics and allow more flexibility in adjusting the oxygen supply through the membrane. A mathematical simulation was also performed to better understand the membrane reactor system and provide guidance on its design and operation.
8.2 Experimental Section

8.2.1 Catalyst Preparation *

The 5% Ru/Al₂O₃ catalyst was the same as used in Chapter 7. Ruthenium chloride (Johnson Matthey, 40.34% Ru), sodium hydroxide (Alfa Aesar) and gamma-alumina oxide (γ-Al₂O₃, Sasol HP14/150; particle size d₅₀: 40 μm) were all used as received. The 5 wt% Ru/Al₂O₃ catalyst was prepared by deposition-precipitation. The γ-Al₂O₃ support was suspended in water at 338 K to form a slurry. To this slurry, two solutions were added: 0.03M NaOH and 0.05M RuCl₃. The pH of the slurry was maintained at pH 9 by addition of appropriate amounts of these two solutions. Once addition of the ruthenium salt was complete, the temperature of the slurry was raised to boiling, and was then allowed to cool. The catalyst was separated by filtration, and the filtrate was washed thoroughly with hot water. Finally, the catalyst was dried at 373 K overnight. The ruthenium content was analysed by Inductively Coupled Plasma (ICP) analysis and found to be 4.46 wt% Ru.

8.2.2 Experimental Set-up

The picture and the schematic of the stirred reactor are shown in Figure 8-1. Initially, the reactor was designed as a stirred membrane reactor with Teflon AF-2400 membrane placed between the top and bottom plates (the schematic is shown in Figure D-1), in an attempt to in-situ provide oxygen through the membrane and avoid the transverse mass transfer resistance existing in packed bed membrane reactor. 1 mm high fins were designed in the top plate to support the membrane. To increase the ratio of membrane area to reactor volume, a shallow chamber was

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* Catalyst preparation was carried out by Dr Peter Ellis, Johnson Matthey.
designed in the bottom plate based on the size of the magnetic stir bar. One inlet and
one outlet were designed in each plate for gas/liquid flow. A frit ferrule with 0.5 μm
pore size (frit disk diameter: 1.2 mm; thickness: 0.8 mm; Upchurch P-275) was used
as filter at the inlet and outlet of the bottom plate. However, the frit was found to be
blocked quickly, when catalyst powder was used in the reactor. This was probably
casted by the small frit area. Dense catalyst layer was also seen at the connection
between the frit ferrule and the reactor chamber (connection diameter: 1 mm; length:
1 mm). In order to avoid the blocking, a piece of sintered metal with large filter area
was used between the top and the bottom plates of the reactor, and a membrane
contactor was used externally to the reactor for oxygen supply.

As shown in Figure 8-1, the stirred reactor consisted of a bottom plate (length: 11
cm; width: 11 cm; thickness: 1.5 cm; 316 stainless steel) and a top plate (the same
dimensions and material as the bottom plate). The reactor chamber (diameter: 6 cm;
height: 1 cm) was machined in the bottom plate, and the sintered metal disk
(diameter: 70 mm; thickness: 1.1 mm; 304 stainless steel; Mott, grade 0.1) was
placed between the top and bottom plates. A Kalrez® O-ring (inner diameter (ID): 70
mm; outer diameter (OD): 76 mm; DuPont™) was used in the bottom plate to seal
the metal disk and the reactor. Catalyst powder was added into the bottom plate,
together with a PTFE coated magnetic stir bar (diameter: 8 mm; length: 50 mm;
cylindrical), before the whole reactor was assembled with bolts and nuts. A
thermocouple opening (diameter: 4 mm; 2 mm away to the chamber) was drilled in
the bottom place for temperature measurement and control.
Figure 8-1 Picture a) and schematic b) of the continuous stirred reactor.
CHAPTER 8. Heterogeneously Catalysed Aerobic Oxidation of Benzyl Alcohol in a Continuous Stirred Membrane Reactor System

Figure 8-2 Schematic of the stirred membrane reactor set-up. MFC: mass flow controller; BPR: back pressure regulator; $P_1$ and $P_2$: pressure sensor; $T$: temperature measurement and control.

Figure 8-3 Picture of the Teflon AF-2400 tube-in-tube contactor.
A schematic of the experimental set-up is shown in Figure 8-2. The stirred reactor was placed on a hotplate (Stuart US152) fitted with a thermocouple, and covered with an insulation cap (insulation thickness: ~2 cm; WDS Ultra) to decrease heat loss. The reaction temperature was measured and controlled through the thermocouple inserted into the opening of the bottom plate (shown in Figure 8-1). The stirring speed was set at ~800 rpm. A Teflon AF-2400 tube-in-tube contactor (shown in Figure 8-3) was placed at room temperature (~ 298 K) and used to introduce oxygen into the liquid. The contactor comprised of an inner Teflon AF-2400 tube (ID: 0.8 mm; OD: 1.0 mm; Biogeneral) and an outer PTFE tube (ID: 2.4 mm; OD: 3.2 mm; Agilent). The length of the tube-in-tube contactor was 80 cm. PFA tubing (ID: 1.0 mm; OD: 1.6 mm; Upchurch) was used for the connections in the set-up. Pure oxygen (BOC, grade N6.0, 5 mL/min (at standard temperature and pressure)) was regulated by a mass flow controller (Brooks, GF40 series) and directed to the annulus between the inner and outer tubes in the contactor. A back pressure regulator (BPR, Swagelok K series) was connected at the gas outlet of the contactor to maintain the oxygen pressure, and the oxygen pressure was measured with a pressure sensor (Zaiput, Hastelloy/PFA wetted parts). Neat benzyl alcohol (99.0%, Sigma-Aldrich) was delivered with a HPLC pump (Knauer P2.1S) and combined with the recycle flow from a piston pump (milliGAT® LF) through a T-junction before introduced into the inner tube of the membrane contactor. The liquid outlet of the contactor was connected to the inlet of the reactor, and the outlet of the reactor was passed through a coil (~30 cm long PFA tubing (OD: 1.6 mm; ID: 1.0 mm)) submerged under water to cool down (milliGAT® pump operating temperature, 273-323 K; the pump components are resistant to toluene at room temperature), and
then divided into two streams. One was connected to the inlet of the recycle pump and the other was directed to an adjustable BPR (Zaiput, BPR-01). The liquid pressure was maintained by the BPR and the actual pressure was monitored by a pressure sensor (Zaiput, Hastelloy/PFA wetted parts) placed upstream of the contactor.

Before starting the experiment, the sealing of the set-up was checked with pure nitrogen at ~ 9 bar pressure. Then, the liquid was directed into the contactor and the reactor. When the contractor and the reactor were filled with liquid and the liquid pressure reached the set value, the gas phase of the membrane contactor was pressurized, and the recycling and the heating were started.

The liquid effluent was collected in a 2 mL vial (placed in an ice-water cold trap) and quantitatively analysed by a gas chromatograph (Agilent 7820A) fitted with a DB-624 capillary column and a flame ionization detector.

Benzyl alcohol conversion ($X$) and selectivity ($S_B$) to benzaldehyde were calculated according to the following equations:

$$X = \frac{c_{BnOH,IN} - c_{BnOH,OUT}}{c_{BnOH,IN}} \times 100\% \quad (8-1)$$

$$S_B = \frac{c_{Bald,OUT}}{c_{BnOH,IN} \times X} \times 100\% \quad (8-2)$$

where $c_{BnOH,IN}$ is the concentration of benzyl alcohol at the inlet of the reactor system (before mixing with recycle), $c_{BnOH,OUT}$ is the concentration of benzyl alcohol at the outlet of the reactor system, which is equal to the concentration of benzyl alcohol at the outlet of the reactor ($c_{BnOH,out}$). $c_{Bald,OUT}$ is the concentration of benzaldehyde at the outlet of the reactor system.
Catalyst contact time (CCT) was used to characterise the reaction time of benzyl alcohol, and was defined as:

$$CCT = \frac{m_{\text{cat}}}{\nu_{\text{IN}} \cdot \rho}$$  \hspace{1cm} (8-3)$$

where $m_{\text{cat}}$ is the mass of catalyst, $\nu_{\text{IN}}$ is the inlet flow rate of the reactor system, $\rho$ is the density of benzyl alcohol.

### 8.2.3 Reactor Simulation

Mathematical simulation of the membrane reactor system was performed using gPROMS ModelBuilder 4.2.0 based on the following assumptions: 1) steady-state and isothermal conditions; 2) the volume flow rate of liquid does not change with the reaction (the organics pervaporation in the membrane contactor was estimated to be less than $7 \times 10^{-4}$ mg/min, shown in Appendix D); 3) perfect mixing within the reactor; 4) oxygen solubility in the liquid is constant at room temperature ($\sim 298$ K).

The model equations are:

$$v_{\text{in}} = v_{\text{IN}} + v_R$$  \hspace{1cm} (8-4)$$

$$v_{\text{in}} \cdot c_{\text{BnOH,in}} = v_{\text{IN}} \cdot c_{\text{BnOH,IN}} + v_R \cdot c_{\text{BnOH,out}}$$  \hspace{1cm} (8-5)$$

$$-R = k \cdot c_{\text{BnOH,out}}^g \cdot c_{\text{O2,out}}^\beta$$  \hspace{1cm} (8-6)$$

$$-R \cdot V = v_{\text{in}} \cdot c_{\text{BnOH,in}} - v_{\text{out}} \cdot c_{\text{BnOH,out}}$$  \hspace{1cm} (8-7)$$

$$-\frac{1}{2} R \cdot V = v_{\text{in}} \cdot c_{\text{O2,in}} - v_{\text{out}} \cdot c_{\text{O2,out}}$$  \hspace{1cm} (8-8)$$

$$v_{\text{out}} = v_{\text{OUT}} + v_R$$  \hspace{1cm} (8-9)$$

$$v_{\text{in}} = v_{\text{out}}$$  \hspace{1cm} (8-10)$$

$$v_{\text{IN}} = v_{\text{OUT}}$$  \hspace{1cm} (8-11)$$
where \( v_{IN} \) is the flow rate of benzyl alcohol at the inlet of the reactor system (from the HPLC pump), \( v_R \) is the recycle flow rate from the piston pump, \( v_{in} \) is the inlet flow rate of the tube-in-tube contactor and the reactor, \( v_{out} \) is the outlet flow rate of the reactor, \( v_{OUT} \) is the flow rate at the outlet of the reactor system, \( c_{BnOH,IN} \) is the molar concentration of neat benzyl alcohol at the inlet of the reactor system, \( c_{BnOH,in} \) and \( c_{BnOH,out} \) are the molar concentrations of benzyl alcohol at the inlet and outlet of the reactor. \( c_{O2,in} \) and \( c_{O2,out} \) are the molar concentrations of oxygen at the inlet and outlet of the reactor. \( R \) is the reaction rate, \( \alpha \) and \( \beta \) are the reaction orders, \( k \) is the reaction rate coefficient. \( V \) is the volume of the reactor. The parameters and assigned variables are shown in Table 8-1.

When the tube-in-tube contactor is long enough for a given flow rate, the liquid flow at the outlet of the contactor would be saturated with oxygen.\(^{123, 147}\) To confirm this, a two-dimensional axisymmetric model (shown in Figure 8-4) was developed and validated in Jensen’s group.\(^ {119}\) The assumptions in the model are: 1) steady-state and isothermal conditions; 2) Henry’s law applied to the membrane-liquid interface; 3) ideal gas behavior is valid for the gas phase in the annulus; 4) laminar flow in the inner tube with constant physical properties and transport coefficients as pure benzyl alcohol; 5) negligible dissolution of liquid in the membrane and pervaporation to the gas phase. Given that the gas pressure in the annulus between the inner and outer tubes was constant, only the inner tube and membrane were considered with boundary condition of constant oxygen concentration at the interface of gas and membrane.
The mole balance for oxygen in the inner tube can be expressed as:

\[ u_z \frac{\partial c_{O_2}}{\partial z} - D_{O_2} \left[ \frac{\partial^2 c_{O_2}}{\partial z^2} + \frac{1}{r} \frac{\partial}{\partial r} \left( r \frac{\partial c_{O_2}}{\partial r} \right) \right] = 0 \]  
(8-12)

\[ u_z = 2u_{ave} \left[ 1 - \left( \frac{r}{r_1} \right)^2 \right] \]  
(8-13)

\[ u_{ave} = \frac{v_{in}}{\pi r_1^2} \]  
(8-14)

where \( r_1 \) is the radius of the inner tube, \( D_{O_2} \) is the oxygen molecular diffusion coefficient in benzyl alcohol, \( u_z \) is the axial liquid superficial velocity vector, \( u_{ave} \) is the average liquid flow velocity.

Boundary conditions:

\[ \text{at } r = 0, \frac{\partial c_{O_2}}{\partial r} = 0 \]  
(8-15)

\[ \text{at } r = r_1, c_{O_2} = \frac{c_{O_2, m}}{H} \]  
(8-16)

\[ \text{at } z = 0, c_{O_2} = c_{O_2, \text{out}} \]  
(8-17)
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\[ \text{at } z = L, \frac{\partial c_{O_2}}{\partial z} = 0 \]  

where \( H \) is the dimensionless Henry’s law constants for solubility of oxygen in benzyl alcohol at 298 K, \( c_{O_2,m} \) is the oxygen concentration in the membrane, \( c_{O_2,out} \) is the oxygen concentration at the reactor outlet.

The steady-state mole balance of oxygen in the membrane domain, considering only diffusion, can be expressed as:

\[ D_{O_2,m} \left[ \frac{\partial^2 c_{O_2,m}}{\partial z^2} + \frac{1}{r} \frac{\partial}{\partial r} \left( r \frac{\partial c_{O_2,m}}{\partial r} \right) \right] = 0 \]  

Boundary conditions:

\[ \text{at } r = r_1, \ c_{O_2,m} = H c_{O_2} \]  

\[ \text{at } r = r_2, \ c_{O_2,m} = c_{O_2,g} \]  

\[ \text{at } z = 0, \frac{\partial c_{O_2,m}}{\partial z} = 0 \]  

\[ \text{at } z = L, \frac{\partial c_{O_2,m}}{\partial z} = 0 \]  

where \( D_{O_2,m} \) is the diffusion coefficient of oxygen in the membrane, \( c_{O_2,g} \) is \( O_2 \) concentration in the gas phase.

The oxygen concentration at the outlet of the contactor was calculated in the model based on the velocity-weighted average of oxygen concentration at the inner tube outlet. The equations with the corresponding boundary conditions were solved using COMSOL Multiphysics software version 5.2. A mesh consisting of 7496 elements 3974 degree of freedom was used and the results were found to be mesh independent. The values of the variables used in the model are shown in Table 8-1.

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Figure 8-5 Iteration between COMSOL and gPROMS to get the oxygen concentration at the inlet of the reactor.

When the contactor was not long enough to saturate the liquid flow with oxygen, the oxygen concentration at the outlet of the contactor was affected by the oxygen concentration at the inlet of the contactor, which was determined by the reactor system inlet flow rate and the oxygen concentration at the outlet of the reactor. To calculate the oxygen concentration at the inlet of the reactor during the stable period, a manual iteration between COMSOL and gPROMS was performed as below (shown in Figure 8-5): $c_{O_2,\text{out}0}$ was initially assumed to be 0 M in COMSOL and the $c_{O_2,\text{in}1}$ was calculated based on the velocity-weighted average of oxygen concentration at the outlet of the contactor in the model. Then, $c_{O_2,\text{in}1}$ calculated from COMSOL was used in gPROMS to get $c_{O_2,\text{out}1}$, which was used again in COMSOL to get $c_{O_2,\text{in}2}$. The iteration was finished when the concentration difference, $c_{O_2,\text{in}n+1} - c_{O_2,\text{in}n} < 1 \times 10^{-4}$ M. The final $c_{O_2,\text{in}n}$ was used as the oxygen concentration at the inlet.
of the reactor during the stable period ($c_{O_2,m}$) in gPROMS.

### Table 8-1 Physicochemical and operation parameters used in the model

<table>
<thead>
<tr>
<th>Symbols</th>
<th>Values</th>
<th>Units</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>$c_{O_2,g}$</td>
<td>$\frac{P_{O_2}}{RT}$</td>
<td>M</td>
<td>Oxygen concentration in the gas phase</td>
</tr>
<tr>
<td>$c_{BnOH,IN}$</td>
<td>9.6</td>
<td>M</td>
<td>Benzyl alcohol concentration at reactor system inlet</td>
</tr>
<tr>
<td>$D_{O_2}$</td>
<td>$4.14 \times 10^{-9}$</td>
<td>m$^2$/s</td>
<td>Oxygen diffusion coefficient in benzyl alcohol at 298 K$^{162,164}$</td>
</tr>
<tr>
<td>$D_{O_2,m}$</td>
<td>$0.95 \times 10^{-9}$</td>
<td>m$^2$/s</td>
<td>Oxygen diffusion coefficient in the membrane at 298 K$^{114,119}$</td>
</tr>
<tr>
<td>$H$</td>
<td>14.7</td>
<td>-</td>
<td>Dimensionless Henry’s law constant of O$_2$ in benzyl alcohol at 298 K</td>
</tr>
<tr>
<td>$M_{BnOH}$</td>
<td>108.1</td>
<td>g/mol</td>
<td>Molecular weight of benzyl alcohol</td>
</tr>
<tr>
<td>$l$</td>
<td>80</td>
<td>cm</td>
<td>Length of tube-in-tube contactor</td>
</tr>
<tr>
<td>$P_{O_2}$</td>
<td>0-10</td>
<td>bar</td>
<td>Oxygen pressure in the annulus of the contactor</td>
</tr>
<tr>
<td>$r_1$</td>
<td>0.4</td>
<td>mm</td>
<td>Inner radius of Teflon AF-2400 tube</td>
</tr>
<tr>
<td>$r_2$</td>
<td>0.5</td>
<td>mm</td>
<td>Outer radius of Teflon AF-2400 tube</td>
</tr>
<tr>
<td>$\bar{R}$</td>
<td>$8.314 \times 10^{-5}$</td>
<td>m$^3$·bar/(K·mol)</td>
<td>Ideal gas constant</td>
</tr>
<tr>
<td>$T$</td>
<td>298</td>
<td>K</td>
<td>Temperature of the contactor</td>
</tr>
<tr>
<td>$\rho$</td>
<td>1.04</td>
<td>g/mL</td>
<td>Density of benzyl alcohol at 298 K</td>
</tr>
<tr>
<td>$\nu_{IN}$</td>
<td>2-30</td>
<td>µL/min</td>
<td>Inlet flow rate of the reactor system</td>
</tr>
<tr>
<td>$\nu_R$</td>
<td>0-10</td>
<td>mL/min</td>
<td>Recycle flow from the piston pump</td>
</tr>
</tbody>
</table>

### 8.3 Results and Discussion

#### 8.3.1 Experimental Demonstration of the Stirred Membrane Reactor

Initially, the pressure drop across the contactor (at 298 K) and the reactor (at 373
K) was measured without catalyst and recycling, which was less than 0.1 bar with 1 mL/min reactor system inlet flow rate (all pressures reported in this chapter are absolute). After adding the Ru/Al$_2$O$_3$ catalyst into the reactor (0.5 g catalyst; reactor volume, ~25 mL; catalyst concentration, ~20 g/L), the pressure drop was measured to be ~0.4 bar. Note that the slurry in the reactor was filtered through the sintered metal disk consisting of distributed micro pores to retain the catalyst in the reactor. The increased pressure drop was probably caused by catalyst build-up on the disk. The liquid effluent at the outlet of the reactor was observed with a digital microscope (Keyence VHX-600, observable particle size down to 1~2 μm), and no catalyst particle was observed (shown in Figure D-2), proving that the sintered metal disk works well as a catalyst filter.

Solvent-free oxidation of benzyl alcohol was then performed in the continuous stirred membrane reactor with 0.5 g Ru/Al$_2$O$_3$ catalyst at 373 K. The inlet flow rate of the reactor system was 10 μL/min, with a recycle flow rate of 1 mL/min. Oxygen pressure in the annulus of the tube-in-tube contactor was kept at 5.0 bar and the BPR pressure at the liquid outlet was 6.0 bar. No bubbles were seen in the liquid flow at the inlet and outlet of the reactor. The initial liquid pressure measured at the upstream of the contactor fluctuated from 5.5 to 6.8 bar, which could be possibly caused by the recycle flow (realized with a piston pump) coupled with the catalyst filtration through the disk. A blank test without catalyst was carried out with the other conditions the same, and the pressure fluctuation was less than 0.1 bar. In the following 90 h, the pressure measured at the upstream of the contactor increased to be 8.3~10.4 bar (shown in Figure D-3). The pressure at the downstream of the reactor was not measured, so the actual pressure drop across the contactor and the
reactor was unknown during the reaction. If averaging the measured pressure at the upstream of the contactor (to minimise the effect of pressure fluctuation), the pressure at the upstream of the contactor was estimated to increase by ~3 bar in the 90 h reaction. Since the BPR pressure at the liquid outlet was set at 6 bar, this suggests a ~3 bar increase in the pressure drop across the contactor and the reactor. To alleviate the pressure drop, an intermittent reversal of the flow direction was attempted by changing the flow direction of the recycle pump. However, the inlet of the current reactor was not filtered, and the catalyst powder was noticed to flow out from the inlet of the reactor when reversal flow was performed. Only short duration (5-10 s) of reversal flow was applied for several times (3-5) after the 90 h reaction without any reversal flow, and no clear pressure drop was observed in the liquid pressure. After the experiment, catalyst deposition was found on the side of the sintered metal disk contacting the slurry (shown in Figure D-4). This means that the catalyst powder could gradually accumulate on the disk during the reaction, which resulted to the pressure drop slowly increasing. To remove the catalyst deposition, the disk was ultrasonically washed in acetone for ~5 min. The pressure drop was found to be less than 0.1 bar when re-used in the reactor at 373 K with 1 mL/min reactor system inlet flow rate (without catalyst). This indicates that the catalyst deposition was removable from the disk. To avoid the catalyst deposition during the reaction, a new reactor design with filters at both inlet and outlet has been proposed and shown in Appendix D.
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Figure 8-6 Solvent-free oxidation of benzyl alcohol in the stirred membrane reactor system.

Reaction conditions: Ru/Al$_2$O$_3$ catalyst, 0.5 g; neat benzyl alcohol, 10 µL/min; catalyst contact time, 48 g$_{cat}$·min/g$_{BnOH}$; recycle flow rate, 1 mL/min; oxygen pressure, 5.0 bar; BPR pressure at the liquid outlet, 6.0 bar; reaction temperature, 373 K; contactor temperature, 298 K; reactor volume, 25 mL. $X$, conversion of benzyl alcohol; $S_B$, selectivity to benzaldehyde.

As shown in Figure 8-6, the conversion of benzyl alcohol gradually increased from 0 to 13% with selectivity to benzaldehyde from 78 to 94% at first ~60 h. The by-products were observed to be benzene and toluene with an initial selectivity of 16% and 3%, respectively. The reason for the relatively low selectivity to benzaldehyde (as compared to the 99% in the packed bed membrane reactor with the same Ru/Al$_2$O$_3$ catalyst) was not clear. The conversion of benzyl alcohol and the selectivity to benzaldehyde stabilized during the following ~30 h. The long start-up time was caused by the long average residence time of liquid in the reactor system, which was ~43 h. Due to the long start-up time, the effect of stirring speed on the reaction was not investigated. Since oxygen was introduced through the membrane contactor, which avoided the direct mixing of oxygen bubbles with organics in the
reactor under reaction conditions, and the contacorer was operated at room
temperature, which was lower than the flash point (the lowest temperature at which
the vapour will ignite) of benzyl alcohol in air (closed-cup flash point of benzyl
alcohol in air: 373 K\(^{176}\)), the reactor is intrinsically safe. The organic pervaporation
through the membrane could also be decreased due to the low vapour pressure at
room temperature. However, the benzyl alcohol conversion in the flat packed bed
membrane reactor (Chapter 7) at the same catalyst contact time and 8 bar oxygen
pressure was 40%. So, the relatively low conversion in the stirred membrane reactor
also suggests that the oxygen supplied through the recycle flow might potentially
limit the reaction.

### Table 8-2 Comparison of experimental and simulated conversions under different conditions.

Reaction conditions: Ru/Al\(_2\)O\(_3\) catalyst, 0.5 g; reactor volume, 25 mL; neat benzyl alcohol;
recycle flow rate, 1 mL/min; liquid pressure measured at the upstream of the contactor, 6-16
bar; BPR pressure at the liquid outlet, 6 bar; reaction temperature, 373 K, contacor
temperature, 298 K. \(P_{O_2}\), oxygen pressure in the contactor; \(X_{\text{exp}}\), the conversion from the
experiment; \(c_{O_2,\text{sat}}\), oxygen saturation concentration at given oxygen pressure; \(c_{O_2,\text{in}}\) and
\(c_{O_2,\text{out}}\), oxygen concentration at the inlet and outlet of the reactor; \(X_{\text{sim}}\), the conversion from
the simulation.

<table>
<thead>
<tr>
<th>Entry</th>
<th>(P_{O_2}) [bar]</th>
<th>(\nu_i) [(\mu L/\text{min})]</th>
<th>(X_{\text{exp}}) [%]</th>
<th>(X_{\text{sim}}) [%]</th>
<th>(c_{O_2,\text{sat}}) [mM]</th>
<th>(c_{O_2,\text{in}}) [mM]</th>
<th>(c_{O_2,\text{out}}) [mM]</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>5.0</td>
<td>10</td>
<td>12.5</td>
<td>12.7</td>
<td>13.8</td>
<td>13.0</td>
<td>7.0</td>
</tr>
<tr>
<td>2</td>
<td>8.0</td>
<td>20</td>
<td>7.5</td>
<td>10.2</td>
<td>22.0</td>
<td>20.8</td>
<td>11.2</td>
</tr>
<tr>
<td>3</td>
<td>9.1</td>
<td>30</td>
<td>5.0</td>
<td>7.7</td>
<td>25.0</td>
<td>23.6</td>
<td>12.7</td>
</tr>
</tbody>
</table>

The reaction was also performed at different oxygen pressures and inlet flow rates
within the same reactor (~40 h for each set), and the liquid pressure measured
upstream of the contactor was up to 16 bar (~10 bar pressure drop along the
contactor and the reactor). The results after the start-up time are averaged and shown
in Table 8-2. To better understand the reactor performance and provide guidance for
its design and operation, the reactor was simulated in the following study.

8.3.2 Reaction Rate Estimation

Considering the high selectivity obtained on Ru/Al₂O₃ catalyst, the selectivity in the simulation was assumed to be 100%. As shown in Appendix D, several reaction rates with different reaction orders in benzyl alcohol and oxygen were attempted in the model and the reaction rate coefficient, k, was estimated based on the average conversion (13%) in Figure 8-6. Then, these different reaction rate expressions were used in the model for other reaction conditions, and no obvious difference was observed in terms of the simulated conversions. To allow for the dimensionless analysis of these various parameters in the reactor, the reaction rate was simplified to be a first order on oxygen. The estimated k for the 0.5 g Ru/Al₂O₃ catalyst in 25 mL reactor was $1.2 \times 10^{-3}$ s⁻¹, which resulted to a reaction rate (R) of $8.3 \times 10^{-10}$ mol/(g₉cat·s) at 13.8 mM oxygen concentration. This could be seen as an observed reaction rate coefficient, since internal/external mass transfer resistance were not investigated in the reactor. Assuming the same oxygen concentration and the same reaction orders for benzyl alcohol oxidation on the Ru/Al₂O₃ catalyst in different reactors, the estimated catalyst activity in the stirred membrane reactor corresponded to ~1/10 of that in the packed bed membrane reactor (R = $1.0 \times 10^{-8}$ mol/(g₉cat·s)). This could be caused by external mass transfer resistance in the stirred reactor or the un-used catalyst deposited on the filter. The turnover frequency (TOF) in the stirred membrane reactor was ~3.0 h⁻¹, which was also lower than that in the packed bed membrane reactor (~11 h⁻¹).

As seen in Table 8-2, the simulated conversions were very close to those obtained from the experiments, proving that the model based on the estimated kinetics
provided satisfactory prediction of reactor performance. The oxygen concentration at the inlet of the reactor was also very close to the saturation concentration at each given oxygen pressure. To better illustrate the effect of oxygen availability at given oxygen pressure, oxygen saturation ($SA_{O2}$) at given oxygen pressure is defined by\textsuperscript{119}

$$SA_{O2} = \frac{oxygen\ concentration}{oxygen\ saturation\ concentration} \times 100\% \quad (8-24)$$

Under the conditions shown in Table 8-2, the oxygen saturations at the inlet of the reactor inlet are all > 94%. Since oxygen saturation at the inlet of the reactor could be achieved with long enough contactor, $c_{O2,in}$ is assumed to be equal to the saturation concentration at different recycle flow rates in the following simulation.

To characterize the catalyst utilization in the stirred membrane reactor, the catalyst utilization factor ($U$) is defined by

$$U = \frac{Actual\ reaction\ rate}{Reaction\ rate\ that\ would\ result\ if\ entire\ catalyst\ were\ exposed\ to\ the\ oxygen\ saturation\ conditions,\ c_{O2,sat}} \quad (8-25)$$

$$U = \frac{k_{cO2,out}}{k_{cO2, sat}} = \frac{c_{O2, out}}{c_{O2, sat}} \quad (8-26)$$

So, the catalyst utilization factor equals to the oxygen saturation at the outlet of the reactor.
8.3.3 Effect of Reactor System Inlet Flow Rate

Figure 8-7 Simulated conversion and utilization factor as a function of a) reactor system inlet flow rates of benzyl alcohol and b) residence time in the reactor. Reaction conditions: Ru/Al₂O₃ catalyst, 0.5 g; neat benzyl alcohol; recycle flow rate, 1 mL/min; oxygen pressure in the contactor, 7.0 bar; reaction temperature, 373 K, contactor temperature, 298 K. X, conversion of benzyl alcohol; U, utilization factor.

After the model validation, the effect of various parameters was numerically investigated with the model, in order to better understand reactor performance and provide guidance for its design and operation.

The residence time of the reactant in the reactor (τ) is defined by
The recycle ratio ($\hat{R}$) is defined by

$$\hat{R} = \frac{v_R}{v_{IN}}$$

(8-28)

The effect of reactor system inlet flow rate ($v_{IN}$), as well as the residence time in the reactor, was initially studied and the results are shown in Figure 8-7. When decreasing the reactor system inlet flow rate, the conversion of benzyl alcohol was observed to increase dramatically. It could reach 100% when the flow rate was lower than 2 $\mu$L/min. A linear relationship was shown between the conversion and the residence time in the reactor. Since the other reaction conditions were kept the same, the increase in conversion could be attributed to the longer residence time at lower inlet flow rate. Decreasing the inlet flow rate also resulted to reducing the amount of oxygen required to fully convert the benzyl alcohol. However, the utilization factor as well as the oxygen saturation profile was almost constant with inlet flow rate decreasing.

Using equations 8-4 ~ 8-11,

$$k \cdot c_{O_2,\text{out}} \cdot V = v_{IN} \cdot (c_{BnOH,IN} - c_{BnOH,\text{out}})$$

(8-29)

$$\frac{1}{2} k \cdot c_{O_2,\text{out}} \cdot V = (v_{IN} + v_R)(c_{O_2,\text{sat}} - c_{O_2,\text{out}})$$

(8-30)

$$\frac{c_{O_2,\text{out}}}{c_{O_2,\text{sat}}} = \frac{1}{k_V \cdot \frac{1}{2\{(v_{IN} + v_R) + 1\}}} = \frac{1}{k_T \cdot \frac{1}{2\{(1+\hat{R}) + 1\} \}}$$

(8-31)

Combining equation 8-1 and equation 8-8

$$X = \frac{k \cdot c_{O_2,\text{out}} \cdot V}{v_{IN} \cdot c_{BnOH,IN}} = \frac{k \cdot c_{O_2,\text{out}} \cdot \tau}{c_{BnOH,IN}}$$

(8-32)
Since the recycle flow rate \( (\nu_R) \) was at least one order of magnitude larger than the inlet flow rate of the reactor system \( (\nu_{IN}) \), the oxygen concentration at the outlet of the reactor \( (c_{O2,out}) \) could be assumed as constant at different inlet flow rates based on equation 8-31. This could also be seen from the profile of the utilization factor in Figure 8-7. So, based on the equation 8-32, the conversion of benzyl alcohol is inversely proportional to the inlet flow rate of the reactor system but proportional to the residence time of the reactant in the reactor.

### 8.3.4 Effect of Oxygen Pressure in the Contactor

![Graph](image)

**Figure 8-8** Simulated conversion and utilization factor as a function of oxygen pressure in the contactor. Reaction conditions: Ru/Al\(_2\)O\(_3\) catalyst, 0.5 g; neat benzyl alcohol, 10 \( \mu \)L/min; catalyst contact time, 48 g\(_{cat}\)-min/g\(_{BnOH}\); recycle flow rate, 1 mL/min; reaction temperature, 373 K; contactor temperature, 298 K; reactor volume, 25 mL. X, conversion of benzyl alcohol; U, utilization factor.

The effect of oxygen pressure in the contactor was studied with the model and the results are shown in Figure 8-8. The catalyst utilization factor was indicated to be not affected by the oxygen pressure, which could also be explained by equation 8-31. A linear increase was observed in the conversion of benzyl alcohol.
Using equation 8-31,

\[
C_{O_2, out} = \frac{C_{O_2, sat}}{k\nu} + 1 = \frac{C_{O_2, sat}}{k\tau + 1} \quad (8-33)
\]

Inputting equation 8-32,

\[
X = \frac{k\tau}{k_B} \frac{C_{O_2, sat}}{C_{BnOH, in}} + 1 \quad (8-34)
\]

Since the oxygen solubility in the liquid is proportional to the oxygen pressure, the oxygen supply increased with the oxygen pressure increasing, which contributed to a linear increase of the conversion. This also indicates that the amount of oxygen required for the full conversion in the stirred membrane reactor could be potentially reached through increasing the oxygen pressure in the membrane contactor.

### 8.3.5 Effect of Recycle Flow Rate

Since oxygen supply in the stirred membrane reactor was realized by recycling liquid flow through the tube-in-tube contactor, the effect of the recycle flow rate \((\nu_R)\) and the recycle ratio \((\bar{R})\) was investigated in the model and the results are shown in Figure 8-9. When the recycle flow rate was zero, meaning there was no recycling flow, the conversion of benzyl alcohol was only 0.4%. The oxygen saturation at the reactor outlet was 1%, indicating that almost all the oxygen present in the inlet flow was consumed. This indicates an insufficient oxygen supply only through oxygen pre-saturated flow without recycling.\(^{147}\) Indeed, the oxygen saturation concentration at 7 bar oxygen pressure was just 19 mM, which was much lower than the initial concentration of benzyl alcohol (9.6 M). The conversion was observed to rise with the recycle flow rate increasing, reaching 24% at 2 mL/min \((\bar{R} = 200)\). However, further increasing the recycle flow rate did not obviously improve the conversion.
Only a 8% increase was seen when increasing the recycle flow rate from 2 to 10 mL/min (Ṙ from 200 to 1000). The utilization factor was also found to be higher than 0.7 at recycle flow rate >2 mL/min and reach 0.9 at 10 mL/min recycle flow rate. This means that the stirred membrane reactor with high recycle flow rate can potentially work as a traditional CSTR but in an intrinsically safe approach.

**Figure 8-9** Simulated conversion and utilization factor as a function of recycle flow rate/recycle ratio. Reaction conditions: Ru/Al₂O₃ catalyst, 0.5 g; neat benzyl alcohol, 10 μL/min; catalyst contact time, 48 g·min/g_BnOH; oxygen pressure in the contactor, 7.0 bar; reaction temperature, 373 K; contactor temperature, 298 K; reactor volume, 25 mL. X, conversion of benzyl alcohol; U, utilization factor.

Using equation (8-34)

\[
X = \frac{c_{O_2,sat}}{c_{BnOH,N}} \frac{1}{2(1+Ṙ)} \frac{1}{k\tau} \tag{8-35}
\]

At recycle flow rate > 2 mL/min (Ṙ > 200), the value of \(\frac{1}{2(1+Ṙ)}\) was less than half of that of \(\frac{1}{k\tau}\). So, with the recycle flow rate increasing, the recycle ratio increased and the conversion in the reactor was indicated to be not mainly controlled by the
recycle ratio, but by the reaction rate coefficient (k). This suggests that the oxygen supply in the stirred membrane reactor with high recycle flow rate could potentially match the oxygen demand by the catalyst.

8.3.6 Effect of Reaction Rate Coefficient

![Figure 8-10](image)

**Figure 8-10** Simulated conversion and utilization factor as a function of reaction rate coefficient. Reaction conditions: Ru/Al₂O₃ catalyst; neat benzyl alcohol, 10 μL/min; recycle flow rate, 3 mL/min; oxygen pressure in the contactor, 7.0 bar; reaction temperature, 373 K; contactor temperature, 298 K; reactor volume, 25 mL. X, conversion of benzyl alcohol; U, utilization factor.

Since the benzyl alcohol conversion in the reactor would be affected by the reaction rate coefficient, the effect of the reaction rate coefficient was investigated in the model with relatively high recycle flow rate (3 mL/min). Note that the reaction rate coefficient was based on the reactor volume, so it was affected by both the activity of the catalyst and the amount of catalyst used in the reactor. As shown in Figure 8-10, the conversion of benzyl alcohol increased with the reaction rate coefficient increasing. Since the oxygen supply was constant in this case (due to the constant recycle flow rate and oxygen pressure), a reverse trend is observed for the
utilization factor, which decreased from 0.97 to 0.41 when the reaction rate coefficient increasing from $1.2 \times 10^{-4}$ to $5.9 \times 10^{-3}$ s$^{-1}$ at 3 mL/min recycle flow rate. This means that the conversion could be improved with increasing the reaction rate coefficient, but with a lower catalyst utilization.

8.3.7 Interaction of Various Parameters in the Stirred Membrane Reactor System

Through the reactor modelling, the benzyl alcohol conversion in the stirred membrane reactor was indicated to be affected by various parameters, such as the reactor system inlet flow rate, the oxygen pressure, the recycle flow rate and the reaction rate coefficient. In order to assess the relative importance of these parameters that affect the reactant conversion and the catalyst utilization, the Damköhler number ($Da$) for the first order reaction, which relates the reaction timescale to the flow timescale, is defined by

$$Da = k\tau$$  \hspace{1cm} (8-36)

The ratio of the oxygen saturation concentration to the alcohol inlet concentration ($CR$) is given by

$$CR = \frac{c_{O2, sat}}{c_{BnOH, IN}}$$  \hspace{1cm} (8-37)

Using equation 8-26 and 8-31,

$$U = \frac{c_{O2, out}}{c_{O2, sat}} = \frac{1}{\frac{Da}{2(1+\tilde{R})} + 1}$$  \hspace{1cm} (8-38)
CHAPTER 8. Heterogeneously Catalysed Aerobic Oxidation of Benzyl Alcohol in a Continuous Stirred Membrane Reactor System

Figure 8-11 Catalyst utilization factor (U) as a function of the Damköhler number (Da) for various values of recycle ratio ($\bar{R}$). ■: the value for the current reaction conditions in section 8.3.1.

Thus, the various parameters in the reactor are correlated in equation 8-38, which can be used to indicate if the catalyst in the stirred membrane reactor is utilized efficiently. In Figure 8-11, the catalyst utilization factor is shown as a function of the Da number for various values of recycle ratios. It can be seen that the utilization factor decreases with the Da number but increases with the recycle ratio. The value for the reaction conditions in section 8.3.1 is also shown in Figure 8-11 with the symbol ■. The Da number is 176, which suggests that the reaction timescale is much shorter than the flow time scale. However, the utilization factor is 0.54 at the recycle ratio of 100, which means that the catalyst is not fully used in the stirred membrane reactor. To increase the utilization factor, the Da number should be decreased (through decreasing the reaction rate coefficient or the residence time) or the recycle flow ratio should be increased.

To characterize the effect of the various parameters on the reactant conversion, the equation 8-35 is converted to
which can be used to estimate the conversions under different conditions and further guide the reactor design and operation. In Figure 8-12, the reactant conversion is shown as a function of the Da number for various values of the concentration ratio (CR) at different recycle ratios (\( \dot{R} \)). At \( \dot{R} = 0 \) (Figure 8-12a), the conversion increases with the Da number and then reaches a plateau for low \( \dot{M} \) values. When the \( \dot{R} \) value increasing to 100 (Figure 8-12b), no plateau was observed for the same range of Da number. This is caused by the relative values of \( \frac{1}{2(1+\dot{R})} \) and \( \frac{1}{D \dot{a}} \). When \( \dot{R} \rightarrow \infty \) (Figure 8-12c), the reactor behaves as a CSTR, and the conversion is proportional to both the concentration ratio and the Da number. The value for the reaction conditions in section 8.3.1 is also shown in Figure 8-12b) with the symbol ■. The Da number is 176 and the concentration ratio is \( 1.4 \times 10^{-3} \). It can be seen that the conversion can be increased by increasing the concentration ratio (either through increasing the oxygen pressure in the contactor or decreasing the alcohol concentration at the inlet of the reactor system). Notably, the experimental point is close to the plateau, which means that further increasing the Da number without increasing the recycle ratio will not significantly increase the conversion.
Figure 8-12 Conversion ($X$) as a function of the Damköhler number ($Da$) for various values of concentration ratio ($CR$) at different recycle ratios ($Ṙ$). a) $Ṙ = 0$; b) $Ṙ = 100$. ■: the value for the current reaction conditions in section 8.3.1; c) $Ṙ \to \infty$. 
8.4 Conclusion

This chapter demonstrated a continuous stirred membrane reactor system with oxygen pre-saturated effluent recycle for aerobic oxidation of benzyl alcohol with Ru/Al$_2$O$_3$ catalyst powder. A sintered metal disk worked well as catalyst filter to keep the solid catalyst powder within the reactor during the reaction, which allowed catalysts with small particle size to be directly applied in continuous processes. Through a modelling study, the stirred membrane reactor was indicated to be able to supply enough oxygen for the full conversion of the reactant or the catalyst by adjusting the oxygen pressure in the contactor or the recycle flow rate. The benzyl alcohol conversion in the reactor was determined by the reactor system inlet flow rate. At given inlet flow rate, the reactant conversion had in a linear relationship with the oxygen pressure in the contactor, but reached a plateau at high recycle flow rate. This was caused by the relatively low reaction rate coefficient as compared to the high oxygen supply. The catalytic utilization in the reactor was shown to be mainly determined by the recycle flow rate and the reaction rate coefficient, rather than the reactor system inlet flow and the oxygen pressure. Performance indices were defined such as the catalyst utilization factor and the reactant conversion, and related to the reactor operating parameters in order to guide the reactor design and operation.
CHAPTER 9. Conclusions and Future Outlook

Several types of continuous flow membrane reactors have been investigated for heterogeneously catalysed oxidation of alcohols with oxygen. In this chapter, the highlights of this research are presented, and a flow chart is provided to guide the application of membrane reactors. Future work on other alcohol oxidation reactions, membrane materials, and types of reactor is also recommended.
9.1 Conclusions

In this thesis, the activity and stability of Au-Pd/TiO$_2$ catalyst were studied for aerobic oxidation of alcohols, and oxygen solubility in the organic solvents was experimentally measured. Then, various types of membrane reactors were experimentally and theoretically investigated for heterogeneously catalysed aerobic oxidation of alcohols, and a fundamental process understanding was obtained.

In Chapter 3, a packed bed capillary microreactor was demonstrated for Au-Pd/TiO$_2$ catalyst deactivation study in aerobic oxidation of alcohols. As compared to batch reactors, the experimental conditions in the microreactor could be easily changed during the reaction, which contributed to a better understanding of the complex role of oxygen on the Au-Pd/TiO$_2$ catalyst in oxidation of cinnamyl alcohol, as well as the different deactivation characteristics for the oxidation of benzyl alcohol and cinnamyl alcohol.

After the catalyst activity and stability testing, a semi-permeable Teflon AF-2400 tube-in-tube microreactor was developed in Chapter 4 for heterogeneous oxidation of benzyl alcohol with molecular oxygen. Through packing the catalyst within the reactor, the packed tube-in-tube microreactor allowed a continuous supply of oxygen during the reaction, and therefore significantly improved both conversion and selectivity as compared to non-permeable PTFE configurations. The performance of the membrane microreactor in terms of conversion and selectivity was also indicated to be determined by the amount of oxygen that permeated through the membrane, which can be enhanced by increasing the oxygen pressure and the residence time of liquid in the bed.
Since gas saturation in the liquid flow can be achieved with Teflon AF-2400 membrane contactor and the data of oxygen solubility are important for the accurate reactor design in aerobic oxidation of alcohols, a novel approach using a Teflon AF-2400 tube-in-tube membrane contactor was demonstrated for the measurement of gas solubility in organic solvents in Chapter 5. Good agreement was observed between the obtained oxygen solubility in toluene and the literature data, proving the reliability of this method. The different solubilities of oxygen in benzyl alcohol and toluene also indicate the importance of measuring solubilities in the solvents/substrates used in the actual reactions, as using solubilities of similar (but not identical) components can lead to errors in reactor design.

Considering the potential drawbacks existing for the tube-in-tube membrane microreactor (notable heat transfer resistance in the annulus, difficult scale-up, etc.), a flat membrane microreactor was further developed with a Teflon AF-2400 membrane film in Chapter 6, which widened the operating pressure range and simplified the reactor scale-up. Scale-up of the microreactor by an order of magnitude was demonstrated by increasing the width of the catalyst bed channel. The importance of oxygen on the oxidation reaction in the membrane reactor was highlighted by the higher reaction rates and main product selectivities at higher oxygen pressure. Through study of the various mass transfer processes, oxygen transverse mass transfer in the catalyst bed is suggested to be the controlling process for the oxidation of benzyl alcohol on the highly active Au-Pd/TiO$_2$ catalyst in the flat membrane reactor. This agreed with an effectiveness factor analysis, which indicated that both oxygen permeation rate in the membrane and oxygen reaction rate in the catalyst were much higher than the oxygen transverse mass transfer rate.
The effectiveness factor analysis also provides a simple approach to assess the relative importance of the various mass transfer processes on the reactor performance, and guide catalyst choice and membrane reactor design.

In Chapter 7, an experimental and modelling study of the flat membrane reactor was performed with highly selective Ru/Al₂O₃ catalyst. A 2D reactor model was developed and validated with experimental data, which provided a better understanding of the mass transfer and catalytic reactions in the membrane reactor. The main part of the catalyst bed was indicated to have no oxygen supply, and the oxygen transverse mass transfer, rather than the oxygen transfer in the membrane or the oxygen external/internal mass transfer, was confirmed to be the controlling process in the membrane reactor. This agrees with the previous study with Au-Pd/TiO₂ catalyst. Decreasing the liquid channel depth and using highly active catalyst were suggested to increase the reactor conversion. To simultaneously improve the catalyst utilization and the reactor conversion, the transverse dispersion coefficient should be increased, possibly by increasing the liquid velocity combined with recycling or using a forced convection mechanism (oscillation, ultrasonic vibration, etc.)

In Chapter 8, a continuous stirred membrane reactor was demonstrated for the usage of small catalyst particles in aerobic oxidation of benzyl alcohol. A sintered metal disk worked well as catalyst filter, which maintained the solid catalyst within the reactor during the reaction and allowed catalysts with small particle size to be directly applied in continuous processes. Through a modelling study, the stirred membrane reactor was indicated to potentially provide enough oxygen for the full conversion of the reactant or the catalyst through adjusting the oxygen pressure in
the contactor or the recycle flow rate. The reactant conversion and the catalyst utilization were shown to be affected by the various parameters in the reactor, and performance indices were defined and related to these parameters in order to guide the reactor design and operation.

In summary, oxygen supply in the membrane reactor has been indicated to be a determining factor for the performance of the membrane reactor. In the packed bed membrane reactor, the performance of the reactor and the catalyst is determined by the oxygen transverse mass transfer in the catalyst bed. Since in the laminar flow regime, the oxygen transverse mass transfer coefficient is inversely proportional to the transfer distance, the tube-in-tube reactor with a shallow channel depth seems to be superior. However, the flat membrane reactor has wider operating pressure range and easier scale-up. The channel depth of the flat reactor could also be simply adjusted. Therefore, the flat membrane reactor is more promising for aerobic oxidation of alcohols. As compared to the packed bed membrane reactor, the stirred membrane reactor could potentially avoid the limitation of the oxygen transverse mass transfer and provide a comparable oxygen supply at high recycle flow rate. However, the issue of catalyst deposition on the filter and the energy efficiency of the reactor need to be considered.

As gaseous oxygen does not come directly into contact with the organic mixture in the membrane reactors, this type of reactors possesses intrinsic safety for aerobic oxidation of alcohols. The simple assembling and scale-up of the developed membrane reactors suggest that other similar membranes could also be used for continuous flow oxidation of alcohols with molecular oxygen. The controllable supply of gas also makes it promising for wider applications in heterogeneous
reactions.

To provide a guideline on the application of the membrane reactors, a flow chart is provided in Figure 9-1. The selection of membrane reactor consists of two steps: reactor selection and membrane selection. The reactor selection can be determined by the catalyst deactivation and the catalyst size. For catalyst with fast deactivation, slurry flow reactor can be considered to avoid the frequent replacement of catalyst in the reactor. For catalyst with small sizes (typically < 50 μm), stirred reactor is suggested to avoid high pressure drop in packed bed reactor. After the reactor selection, the membrane selection can be carried out, and the chemical compatibility of the membrane can be used as a criterion to determine the membrane material. Then, the gas/organics permeability through the membrane should be considered, in order to provide enough gas for the reaction and simultaneously minimize organics loss through the membrane. For highly permeable organics, porous inorganic membrane is suggested, which can potentially avoid the organics loss by using static gas in the gas phase. However, the organics partial pressure in the gas phase should be considered to keep the organic vapour concentration in the gas phase outside the explosive limits. For organics with high vapour pressure under reaction conditions, the membrane unit in the reactor system can be separated from the reactor and used at lower temperature to decrease the organic vapour pressure. Finally, the selection in each step could be combined to determine the final type of membrane reactor.
**Figure 9-1** Flow chart on the application of membrane reactors.
9.2 Future Outlook

Future work is suggested to expand the scope of alcohol oxidation reactions, membrane materials, and reactor types. The specific directions are provided in the following sections.

9.2.1 Scope of Alcohol Oxidation Reactions

Oxidation of alcohols is a broad subject and only oxidation of benzyl alcohol and cinnamyl alcohol is mainly investigated in this thesis. Future study could expand the scope to other green reactions with the potential to produce high-value products.

As an example of bio-derived alcohols oxidation, oxidation of glycerol is potentially a sustainable route to synthesize high-value fine chemicals and intermediates such as dihydroxyacetone, tartronic acid and mesoxalic acid. Oxidation of glycerol has been attempted with several types of catalysts in packed bed microreactors (shown in Appendix E). Ru/Al₂O₃ and Pd/Al₂O₃ catalysts showed obvious catalyst deactivation with low selectivity to specific products. Au-Pd/TiO₂ catalyst presented relatively good stability, but the highest selectivity to any product was below 50%. Various reaction parameters such as reaction temperature, base ratio, catalyst particle size, and flow rate were investigated and no improvement on the selectivity was observed.

Furfural, a C5 heterocyclic aldehyde and a promising platform compound, can be produced in large quantities from the lignocellulosic biomass. Oxidation of furfural was investigated on Au-Pd/TiO₂ catalyst, in an attempt to produce valuable products (such as maleic acid, succinic acid, and furoic acid). However, aerobic oxidation of furfural was not observed on Au-Pd/TiO₂ catalyst under base-free
conditions, whereas the addition of base could result in the uncatalysed Cannizzaro reaction even at room temperature (shown in Appendix F). Though in-situ mixing of the reactants with base could minimize the effect of the Cannizzaro reaction in the reactor, a fast analysis of the products is still required to minimize the effect of Cannizzaro reaction on the residual reactants at the reactor outlet. Moreover, the handling of solid products formed during the reaction is still challenging for continuous flow reactors.

Heterogeneously catalysed aerobic oxidation of “GSK diol”, an intermediate for a GlaxoSmithKline drug candidate, has been shown to be greener and more efficient as compared to other uncatalysed routes.\(^{179}\) Initial test of the reaction on Pt-Bi/Al\(_2\)O\(_3\) and Pt-Bi/C in packed bed microreactor has shown different product selectivities as well as catalyst deactivation, as compared to the literature (shown in Appendix G).\(^{179}\) This could be caused by the different catalyst to substrate ratios and the different efficiencies of oxygen supply in the packed bed reactor as compared to batch reactors. When investigating this reaction in the membrane reactor, large amount of acetonitrile pervaporation/permeation through the Teflon AF-2400 membrane was observed, which suggests that the solvent and membrane selection should be prior to the test of the reaction in the membrane reactor.

Currently, the products are off-line analysed with GC or HPLC. To meet the requirement for high-throughput screening of catalyst and reaction conditions in microfluidics, in-situ spectroscopy analysis could be integrated into the reaction system. For example, with the combination of FT-IR spectroscopy in microfluidics, reaction products could be probed in the millisecond range.\(^{180}\)

**9.2.2 Scope of Membrane Materials**
Although Teflon AF-2400 membrane possesses several excellent properties, its high cost could be one limitation for its wider application. From the view of economy, extending the lifetime of Teflon AF-2400 membrane seems to be reasonable. However, when the Kalrez® gasket was used in the scaled-up flat membrane reactor, the Teflon AF-2400 membrane was found to stick to the gasket. When replacing the catalyst, the gasket needed to be peeled off from the reactor, making it likely to break the fragile Teflon AF-2400 membrane. Thus, more work is still required to optimize the operation or re-use the broken Teflon AF-2400 membrane. Another possible solution is to decrease the usage of Teflon AF-2400 membrane in the membrane reactor. Since decreasing the membrane thickness can promote gas transfer within the membrane, it will be beneficial to coat a thin Teflon AF-2400 film onto a firm support. This has been tried with Teflon AF-2400 solution on metal and ceramic supports (shown in Appendix H). However, due to the nonpolar Teflon AF-2400 containing no reactive chemical functionality, the coating results were not successful. Further surface treatment is required to enhance the adhesion between membrane and support.

Except for Teflon AF-2400 membrane, other inexpensive polymer membrane, such as PTFE\textsuperscript{113} and PEEK\textsuperscript{181}, could also be investigated as membrane materials in the membrane reactor. Considering the corrosion of organic solvents, chemical compatibility would be an important criterion for membrane selection. The preliminary assessment of PEEK membrane is shown in Appendix I. The low gas permeability of the membrane could potentially be offset through decreasing the membrane thickness. Nevertheless, the organic pervaporation through the membrane is another factor that should not be ignored. The safe operation needs to ensure the
stream in the gas phase outside the flammability limits. The measurement of organics in the gas flow with in-line measurement (e.g. GC) could be considered during the membrane selection.

Due to the excellent chemical resistance and gas permeability properties, ceramic membranes have an enormous potential for gas-liquid-solid reactions. Facile scale-out to the desired production scale is promising for tubular ceramic membranes by numbering up. A packed bed ceramic membrane reactor has been demonstrated for aerobic oxidation of alcohols. However, experiments conducted so far with commercial Pall™ membranes were limited to low transmembrane pressure difference (< 0.4 bar), which was caused by membrane defects (larger pores than the nominal). In the future work, ceramic membranes screening should be carried out for better breakthrough stability. The potential pervaporation of liquids through the membrane should also be considered. Since oxygen transverse mass transfer in the catalyst bed is a controlling process in the packed bed membrane reactor, impregnation or coating of the top layer of the ceramic membrane with active metals could significantly decrease the distance for oxygen transverse mass transfer, which is a potential strategy to enhance the oxygen transverse mass transfer for aerobic oxidation of alcohols. In that case, the position of the gas-liquid interface needs to be adjusted to simultaneously keep a thin liquid layer for oxygen transfer and assure the full usage of the catalytic layer.

9.2.3 Scope of Reactor Types

This thesis mainly focuses on packed bed reactor and stirred membrane reactor, where catalysts are confined within the reactor. It would also be attractive to develop slurry flow reactors for aerobic oxidation of alcohols, where solid catalysts are
delivered together with the reactants and the issue of maintaining and replacing catalyst within the reactor can be avoided. As mentioned in the literature review, the complex interaction among catalyst particles, fluid, and reactor surface makes it challenging to handle particles in continuous flow reactors. Nevertheless, how to efficiently feed the solid-containing fluid would be the first tough issue.\textsuperscript{170} Several types of pumps have been attempted and the details are shown in Appendix J. More work is still required to avoid the catalyst fouling on the flow sensor and the reactor wall. As compared to microscale solid particles whose dispersion in fluid normally needs the help of stirring within vessel and proper fluid velocity within reactor, nanoscale solid particles could be easily dispersed in the fluid, which could significantly simplify the particles handling. Though the separation of nanoscale particles from reactants can not be realized through gravitational settling, phase separation or magnetic separation could be considered.\textsuperscript{93}

Due to the simplicity and excellent compatibility, fluoropolymer tubing is widely used in continuous flow microreactor. However, the maximum working temperature of such material is up to 250 °C with maximum pressure ~50 bar. Moreover, visualization cannot be simply realized. Transparent silicon-glass microreactors can be used for high-temperature and pressure reactions, but scale-up of such reactors is not simple due to the complex fabrication process. Kovar material, a nickel-cobalt ferrous alloy, can be mechanically milled to have millimetre-scale channels and directly bonded with Pyrex glasses to be used as reactors.\textsuperscript{184} The bonding of Kovar-glass has been tried but unsuccessful due to the relatively rough surface of Kovar plate (shown in Appendix K). Further lapping and polishing is required to make the variation in flatness less than 2 μm and the surface roughness less than 0.15 μm.
References


References

Appendix A. Supporting Information_Measuring Gas Solubility in Liquids Using a Tube-in-Tube Membrane Contactor
### Table A-1 Comparison of oxygen outgassing rates obtained from experimental measurement or ASPEN calculation

<table>
<thead>
<tr>
<th>Oxygen pressure [bar]</th>
<th>( \nu_{O_2}^{\exp.} ) [mL/mL_alcohol]</th>
<th>( \nu_{O_2}^{\sim} ) [mL/mL_alcohol]</th>
<th>UNIQUAC</th>
<th>Peng-Robinson</th>
<th>NRTL</th>
</tr>
</thead>
<tbody>
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<td>3</td>
<td>0.14</td>
<td>0.48</td>
<td>0.46</td>
<td>0.26</td>
<td></td>
</tr>
<tr>
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<td>0.93</td>
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</tr>
<tr>
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<td>0.51</td>
<td>1.43</td>
<td>1.40</td>
<td>0.79</td>
<td></td>
</tr>
</tbody>
</table>
Figure A-1 Picture of a) gas-liquid separator and b) tube-in-tube membrane contactor.
A1. Calculation procedure

The gas phase in the tube-in-tube contactor was assumed to be saturated with liquid vapour due to the low gas flowrate, and the oxygen pressure (P_{O2}, \text{bar}) was corrected by subtracting the liquid vapour pressure (P_V) from the measured gas pressure (P_M). After the oxygen-liquid equilibrium was reached in the tube-in-tube membrane contactor at given oxygen pressure (P_{O2}, \text{bar}) and temperature (T, K), the gas flow rate exiting the separator was measured by the bubble meter at atmospheric pressure (P_{atm} = 1 \text{ bar}) and room temperature (T_1, 298 \pm 2 K). Since the gas measured by the bubble meter (\nu_{bubble}) should be saturated with water vapour (P_W, 0.031 \text{ bar} at 298 K), it must be corrected to obtain the gas flow rate exiting the separator (\nu_{total}):

\[
\nu_{total} = \nu_{bubble} \frac{P_{atm} - P_W}{P_{atm}} \quad (A-1)
\]

The rate of gas flow from the separator (\nu_{total}, \text{mL/min}) includes the flow contribution from the outgassed oxygen saturated with vapour (\nu_{gas}) and the volume change in the separator caused by the collection of liquid (\nu_{gas'}):

\[
\nu_{total} = \nu_{gas} + \nu_{gas}' \quad (A-2)
\]

The outgassed oxygen flowrate (\nu_{gas}) consists of contributions from outgassed oxygen (\nu_{O2}) and solvent vapour in the gas stream (\nu_{vapour}):

\[
\nu_{gas} = \nu_{O2} + \nu_{vapour} \quad (A-3)
\]

The vapour concentration can be calculated from the ratio of the vapour pressure of the liquid (P_V, 0.028 \text{ bar} for toluene and 1.3\times10^{-4} \text{ bar} for benzyl alcohol at 298 K) to atmospheric pressure (P_{atm}). Thus, the flow rate of solvent vapour (\nu_{vapour}) is
calculated by

\[ v_{\text{vapour}} = v_{\text{gas}} \frac{P_V}{P_{\text{atm}}} \]  \hspace{1cm} (A-4)

Combining equations (A-3) and (A-4),

\[ v_{O2} = v_{\text{gas}} \frac{P_{\text{atm}} - P_V}{P_{\text{atm}}} \]  \hspace{1cm} (A-5)

The highest \( v_{\text{gas}} \) utilized in the experiments was < 1.5 mL/min. Hence, the maximum amount of the liquid stream (entering the contactor) that was evaporated after the back pressure regulator (\( v_{\text{loss}} \)) corresponded to < 2.1\times10^{-4} \text{ mL/min}, as calculated by

\[ v_{\text{loss}} = \frac{v_{\text{gas}} P_V T_0}{P_{\text{atm}} T_1 M_L} \frac{M_L}{\rho_L} \]  \hspace{1cm} (A-6)

where \( P_{\text{gas}} \) is molar volume of gas at STP (22400 mL/mol), \( \rho_L \) is the density of liquid (g/mL), \( M_L \) is molar mass of liquid (g/mol), \( T_I \) is room temperature (298 K), and \( T_0 \) is the standard temperature (273K). Considering that the lowest flow rate of the liquid stream (entering the contactor) was 0.25 mL/min, the maximum liquid loss caused by the evaporation was less than 0.1% of the liquid flow rate. Hence, the liquid loss by evaporation in the gas stream after depressurisation was neglected, and the rate of volume change in the separator (\( v_{\text{gas'}} \)) equalled to the flow rate of liquid (\( v_L \), mL/min) set in the HPLC pump:

\[ v_{\text{gas'}} = v_L \]  \hspace{1cm} (A-7)

Combining (A-2) and (A-7),

\[ v_{\text{gas}} = v_{\text{total}} - v_L \]  \hspace{1cm} (A-8)

Combining (A-5) and (A-8),

\[ v_{\text{gas}} = v_{\text{total}} - v_L \]
$v_{O2} = (v_{total} - v_L) \frac{P_{atm} - P_V}{P_{atm}} \quad (A-9)$

The flowrate $v_{O2}$ was measured at room temperature ($T_1$, 298 K), which needs to be converted to the flow rate at standard temperature ($T_0$, 273 K):

$v'_{O2} = (v_{total} - v_L) \frac{P_{atm} - P_V T_0}{P_{atm} T_1} \quad (A-10)$

The amount of oxygen outgassed per liquid volume ($\phi_{O2}$, mL/mL Liquid) at standard temperature and pressure (273 K and 1 bar, STP) could be calculated by

$$\phi_{O2}(T) = \frac{(v_{total} - v_L) \frac{P_{atm} - P_V T_0}{P_{atm} T_1}}{v_L} \quad (A-11)$$

Note that the liquid collected in the separator was saturated with oxygen at 1 bar oxygen pressure and 298 K. Thus, the amount of oxygen dissolved per liquid volume ($\phi_{total}$) in the tube-in-tube membrane contactor at experimental temperature ($T$) and oxygen pressure ($P_{O2}$) included the amount of oxygen outgassed ($\phi_{O2}(T)$) and the amount of oxygen remaining in the liquid at 1 bar oxygen pressure and 298 K ($\phi_1$):

$$\phi_{total} = \phi_{O2}(T) + \phi_1 \quad (A-12)$$

The mole fraction of oxygen in the liquid at given temperature $T$ ($x(T)$) is:

$$x(T) = \frac{n_{O2}}{n_{O2} + n_L} \quad (A-13)$$

where $n_{O2}$ is the moles of oxygen in the liquid, and $n_L$ is the moles of liquid. Due to the low solubility of oxygen in the liquid ($n_{O2}$ was less than 1/100 of $n_L$ at $P_{O2} < 10$ bar), $x$ is approximately equal to the ratio of $n_{O2}/n_L$:

$$x(T) \approx \frac{n_{O2}}{n_L} = \frac{\rho_{O2} v_L}{\rho L} = \frac{M_L}{\rho_{O2} \rho L} \left( \phi_{O2}(T) + \phi_1 \right) \quad (A-14)$$
where $V_{O_2}$ is molar volume of oxygen at STP (22400 mL/mol); $V_L$ is the volume of liquid (mL); $\rho_L$ is the density of liquid (g/mL); $M_L$ is molar mass of liquid (g/mol).

The Henry’s law constant ($H$) at given temperature $T$ is defined by

$$H = \frac{P_{O_2}}{x(T)} \quad (A-15)$$

where $P_{O_2}$ is oxygen pressure. Thus,

$$x(T) = \frac{P_{O_2}}{H} \quad (A-16)$$

Combining (A-14) and (A-16),

$$\frac{M_L}{\bar{V}_{O_2} \rho_L} (\phi_{O_2}(T) + \phi_1) = \frac{P_{O_2}}{H} \quad (A-17)$$

At 298K and atmospheric pressure ($P_{atm}$, 1 bar),

$$x(298 \text{ K}) \approx \frac{\pi_{O_2}}{n_L} = \frac{\bar{V}_{O_2}}{P_{atm} \rho_L} = \frac{M_L}{\bar{V}_{O_2} \rho_L} \phi_1 = \frac{P_{atm}}{H_{298 \text{ K}}} \quad (A-18)$$

Combining (A-17) and (A-18),

$$\frac{M_L}{\bar{V}_{O_2} \rho_L} \phi_{O_2}(T) = \frac{P_{O_2}}{H} - \frac{P_{atm}}{H_{298 \text{ K}}} \quad (A-19)$$

$$\phi_{O_2}(T) = \frac{\bar{V}_{O_2} \rho_L}{M_L H_{298 \text{ K}}} P_{O_2} - \frac{\bar{V}_{O_2} \rho_L}{M_L H_{298 \text{ K}}} P_{atm} \quad (A-20)$$

The term $\frac{\bar{V}_{O_2} \rho_L}{M_L H}$ is constant at given temperature T. Thus, a linear fit would be obtained when $\phi_{O_2}(T)$ is plotted against oxygen pressure ($P_{O_2}$). Henry’s law constant ($H$) is then calculated from the slope of the fitting line ($A$, mL/(mLLiquid·bar)):

$$A = \frac{\bar{V}_{O_2} \rho_L}{M_L H}$$

by
\[ H = \frac{\dot{V}_{O_2} P_L}{M_L A} \]  
(A-22)
Figure A-2 Henry’s law constants of oxygen in toluene calculated with different equations of state in ASPEN PLUS.

Table A-2 Comparison of Henry’s law constants of oxygen in benzyl alcohol obtained from experimental measurement or ASPEN calculation

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<th>$H_{\text{exp.}}$ [bar]</th>
<th>$H_{\text{sim.}}$ [bar]</th>
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Appendix B. Supporting Information_Development of a Flat Membrane Microreactor for Scalable Aerobic Oxidation of Benzyl Alcohol in Flow
Figure B-1 Mechanical drawings of the flat membrane microreactor. (a) gas flow plate and (b) liquid flow plate.
Figure B-2 Mesh used as membrane support in the reactor (a) and enlargement showing the pore size of 76 μm (b).
B1. Design of the Scaled-up Microreactor

During the design of the scaled-up microreactor, CFD modelling was performed with COMSOL Multiphysics® Modelling Software (using Navier-Stokes and the
Appendix B. Supporting Information_Development of a Flat Membrane Microreactor for Scalable Aerobic Oxidation of Benzyl Alcohol in Flow

continuity equations\textsuperscript{185} to ensure the gas/liquid flow uniformly distributed in the channels.

As shown in Figure B-3, the gas flow channels were 1 mm deep in the gas flow plate, and pillars and fins were designed to distribute the gas flow and assist in supporting the membrane. The pillars machined at the inlet and outlet regions were 1 mm high (diameter: 1 mm), with 1.3 mm high fins (length: 54 mm; width: 1 mm) forming 2 mm wide channels in the middle. The slight higher fins were designed to ensure the support of the membrane, since a wide liquid channel was machined in the scaled-up liquid flow plate and 0.5 mm thick gasket was placed between the gas flow plate and the mesh. The effect of these pillars and fins on the gas distribution was simulated in COMSOL, and as shown in Figure B-3a, uniform gas flow could be obtained in the gas flow channels, especially in the channels formed by the fins.

In the liquid flow plate, the liquid flow channel was 68 mm long, 32 mm wide and 0.5 mm deep. Distribution channels (width: 2 mm; depth: 2 mm) were machined at the inlet and outlet of the liquid flow plate. In Figure B-3b, the simulation results in COMSOL presented uniform liquid velocity in the liquid flow channel. The catalyst was packed in the liquid flow channel, and an opening (diameter: 2 mm) was drilled on one side of the liquid flow plate for catalyst packing. Sintered metal filters (length: 40 mm; width: 3.5 mm; thickness: 1 mm; 304 stainless steel; Mott, grade 0.1) were inserted between the liquid distribution channels and the liquid channel (inserted depth: 3 mm) to retain the catalyst particles. To better seal the top edge of the catalyst filters, a Kalrez® gasket (length: 115 mm; width: 50 mm; thickness: 0.5 mm; open area length: 68 mm; width: 32 mm; DuPont\textsuperscript{TM}) was added between the liquid flow plate and the membrane, which also made the total depth of liquid
channel (machined liquid channel plus open area of the Kalrez® gasket) close to 1 mm. After catalyst packing, the opening was blocked with a flat-bottom PFA plug (size: UNF 1/4”-28; Upchurch P-316). The other layers of the scale-up reactor were the same as in Figure 6-1.
Figure B-4 Velocity profiles of fluids in the scaled-up reactor simulated in COMSOL. a) Gas distribution in the middle of the gas channels with oxygen inlet flow rate 15 mL/min. b) liquid distribution in the middle of the liquid channel with benzyl alcohol inlet flow rate 0.1 mL/min.
Figure B-5 Picture a) and schematic b) of assembled scaled-up membrane microreactor with heaters. c) Measured temperatures at different positions (temperature set point 120 °C).
Appendix B. Supporting Information_Development of a Flat Membrane Microreactor for Scalable Aerobic Oxidation of Benzyl Alcohol in Flow

**Figure B-6** Membrane tested under 20 bar pressure difference and 120 °C. a) Membrane supported with mesh; b) Un-supported membrane.

Initially, pure toluene was used as the liquid phase to measure the organics pervaporation through the Teflon AF-2400 membrane. The reactor temperature was set at 120 °C and the pressure in the liquid phase was maintained at 10 bar, with atmospheric pressure in the gas phase. Nitrogen flow (2 mL/min at STP) was introduced into the gas inlet of the reactor, and the gas outlet of the reactor was directly connected to a vial located in a cold trap (0 °C and 1 bar). Toluene liquid was observed in the vial, which indicated that the gas exiting from the vial was saturated with toluene at 0 °C and 1 bar. The amount of the collected toluene liquid in the vial ($\dot{m}_c$) was 27.7 mg/h. The mass flow rate of toluene vapour in the outlet gas ($\dot{m}_{\text{vapour}}$) from the vial can be calculated by

$$\dot{m}_{\text{vapour}} = \frac{\nu_{N2}}{P_{\text{atm}} - P_V} \cdot \frac{P_V}{P_{\text{atm}}} \cdot \frac{M_L}{\tilde{V}_{O2}}$$  \hspace{1cm} (B-1)$$

where $\nu_{N2}$ is the nitrogen flow rate (2 mL/min at standard temperature and pressure (STP, 0 °C and 1 bar), $P_V$ is the vapour pressure of toluene at 0 °C (0.009 bar),$^{186} P_{\text{atm}}$ is the atmospheric pressure (1 bar), $M_L$ is molar mass of toluene (92.1 g/mol), $\tilde{V}_{O2}$ is molar volume of toluene vapour at STP (22400 mL/mol).

The mass flow rate of toluene vapour was calculated to be 4.5 mg/h. So, the total amount of toluene pervaporation through the membrane ($\dot{m}_{\text{total}}$) is calculated by

$$\dot{m}_{\text{total}} = \dot{m}_c + \dot{m}_{\text{vapour}}$$  \hspace{1cm} (B-2)$$

which was 32.2 mg/h.

The flow rate of toluene vapour in the gas flow channel ($\nu_{\text{toluene}}$) at STP is calculated by
$$v_{\text{toluene}} = \frac{\dot{m}_{\text{total}} \bar{v}_{O_2}}{M_L}$$

(B-3)

which was 0.13 mL/min. So, the toluene vapour partial pressure in the gas phase outlet was \(~0.06\) bar. Since the toluene vapour pressure is 1.3 bar at 393 K\(^{186}\), the toluene vapour pressure difference across the membrane was close to the toluene vapour pressure, indicating that the measured pervaporation rate approximated to the maximum.

The maximum pervaporation rate of benzyl alcohol and benzaldehyde was also experimentally measured with the above method, which was \(~1\) mg/h and 10 mg/h, respectively. So, the maximum vapour concentration in the liquid \((\phi_{\text{toluene}})\) is calculated with the maximum pervaporation rate of toluene.

$$\phi_{\text{toluene}} = \frac{v_{\text{toluene}}}{v_{O_2} + v_{\text{toluene}}} \times 100\%$$

(B-4)

When the oxygen flow rate \((v_{O_2})\) is 15 mL/min at STP, \(\phi_{\text{toluene}}\) is 0.86 vol\%, which is lower than the lower explosive limit (1 vol\%). Note that the vapour concentration is the maximum value, since it was estimated with pure toluene.
B3. Estimation of Weisz-Prater Criterion

Due to the high ratio of the initial benzyl alcohol concentration to oxygen saturation concentration at low oxygen pressure (less than 10 bar, the ratio is larger than 90 based on the oxygen solubility in toluene\textsuperscript{143}), benzyl alcohol concentration is in excess and thus neglected. For simplicity, oxygen reaction rate is assumed to be first order in oxygen concentration. The Weisz-Prater criterion ($C_{WP}$) is used to estimate if internal mass transfer is limiting the oxidation reaction:

$$C_{WP} = \frac{-R_{O2}(obs) \cdot \rho_c \cdot r^2}{D_{O2,e} \cdot c_{O2,s}}$$

(B-5)

where $R_{O2}(obs)$ is observed oxygen reaction rate in Section 3.2 (conversion, 57%; selectivity to benzaldehyde, 66%; selectivity to toluene, 31%; catalyst contact time, 577 g\text{cat} / g_{alcohol}$, oxygen pressure, 8.4 bar), $\rho_c$ is the catalyst density (assuming as 1.2 g/cm$^3$), $r$ is the radius of the catalyst (54 μm), $D_{O2,e}$ is the effective diffusion coefficient of oxygen in the catalyst, which is estimated to be 1/10 of the molecular diffusion coefficient of oxygen in benzyl alcohol. $c_{O2,s}$ is the oxygen concentration at the catalyst surface, (94 mol/m$^3$, assuming as the oxygen saturation concentration in toluene\textsuperscript{149}).

The molecular diffusion coefficient of oxygen in benzyl alcohol ($D_{O2,m}$) was calculated based on Wilke-Chang correlation.\textsuperscript{164}

$$D_{O2,m} = 1.1728 \times 10^{-16} \frac{T \chi M_{BnOH}^{0.6}}{\mu V_{O2}} = 5.47 \times 10^{-9} \text{m}^2/\text{s}$$

(B-6)

where $T$ is the temperature in K, $\chi$ is association factor of benzyl alcohol (1 for non-associated solvents), $M_{BnOH}$ is the molecular weight of the benzyl alcohol in g/mol, $\mu$ is the viscosity of benzyl alcohol in Pa·s,\textsuperscript{166} $V_{O2}$ is the molar volume of oxygen at
normal boiling point in m$^3$/kmol.$^{162}$

$C_{WP}$ is calculated to be 0.11, which is $< 0.15$. This indicates that no internal diffusional limitation for oxygen reaction, when oxygen is saturated at the catalyst surface. However, the oxygen concentration was likely axially and transversely varying in the catalyst bed, which makes the use of this $C_{WP}$ (based on oxygen saturation concentration) misleading. So, the Weisz-Prater criterion cannot be used to quantitatively determine if internal diffusion is limiting the reaction.
B4. Calculation of Oxygen Supply Rate and Oxygen Demand Rate

Oxygen supply rate through the membrane at STP ($v_{O_2,mem}$) is calculated by

$$v_{O_2,mem} = \frac{\kappa A_{mem} \Delta P}{d_m} \quad (B-7)$$

where $\kappa$ is the oxygen permeability in the Teflon AF-2400 membrane (1140 barrer, assuming constant permeability at different temperatures), $A_{mem}$ is the membrane area contacting the catalyst bed (1.5 cm$^2$), $\Delta P$ is the pressure difference across the membrane (maximum value, 8.4 bar), $d_m$ is the membrane thickness (0.07 mm).

Oxygen demand rate at STP ($v_{O_2,demand}$) is estimated from the TOF$_O$ in the batch reactor.\textsuperscript{26}

$$v_{O_2,demand} = \frac{1}{2} \cdot TOF_O \cdot n_{meta} \cdot \tilde{V}_{O_2} \quad (B-8)$$

where $TOF_O$ is the turnover frequency of the oxidation reaction observed in batch reactor (8800 h$^{-1}$), $n_{meta}$ is the moles of the metals contained in the packed catalyst ($8.3 \times 10^{-6}$ mol$_{metal}$ in 100 mg catalyst), $\tilde{V}_{O_2}$ is the molar volume of oxygen at STP (22400 mL/mol).
B5. Permeation-Transverse Mass Transfer Effectiveness Factor

The external/internal mass transfer in the catalyst particles has been demonstrated to have no significant effect on the oxidation reaction rate. Considering a slice $\Delta x$ in the reactor, the one-dimensional oxygen permeation, transverse mass transfer and reaction in the packed bed membrane reactor at steady state and isothermal conditions can be described as

$$-D_{O_2,T}\left(\frac{\partial^2 c_{O_2}}{\partial y^2}\right) = R_{O_2}$$  \hspace{1cm} (B-9)

where $D_{O_2,T}$ is the transverse dispersion coefficient of oxygen in the catalyst packed bed, $c_{O_2}$ is the oxygen concentration in the catalyst packed bed, $R_{O_2}$ is the oxygen reaction rate per unit volume of catalyst bed.

As mentioned above, oxygen reaction rate is assumed to be first order in oxygen concentration,

$$R_{O_2} = -k_{O_2}c_{O_2}$$  \hspace{1cm} (B-10)

$$D_{O_2,T}\left(\frac{\partial^2 c_{O_2}}{\partial y^2}\right) = k_{O_2}c_{O_2}$$  \hspace{1cm} (B-11)

where $k_{O_2}$ is the oxygen reaction rate coefficient based on the bed volume.
The boundary conditions are

\[ at \ y = 0, \ \frac{\partial c_{O_2}}{\partial y} = 0 \]  \hspace{1cm} (B-12)

\[ at \ y = d_1, \ D_{O_2,T} \ \frac{\partial c_{O_2}}{\partial y} = k_m (c'_{O_2,g} - c_{O_2}) \]  \hspace{1cm} (B-13)

where \( d_1 \) is the channel depth, \( k_m \) is a hypothetical oxygen mass transfer coefficient in the membrane based on the oxygen concentration in the liquid. \( c'_{O_2,g} \) is a hypothetical oxygen concentration in the liquid phase in thermal equilibrium with the oxygen concentration in the gas phase \( (c_{O_2,g}) \), which makes the oxygen concentration consistent at the membrane-liquid interface.

\[ H c'_{O_2,g} = c_{O_2,g} \]  \hspace{1cm} (B-14)

where \( H \) is the dimensionless Henry solubility of oxygen in the liquid.

Writing the diffusion-reaction equation in dimensionless form through introducing dimensionless variables \( \Psi \) and \( \lambda \)

\[ \frac{\partial^2 \Psi}{\partial \lambda^2} = \phi^2 \lambda \]  \hspace{1cm} (B-15)

with

\[ at \ \lambda = 0, \ \frac{\partial \Psi}{\partial \lambda} = 0 \]  \hspace{1cm} (B-16)

\[ at \ \lambda = 1, \ \frac{\partial \Psi}{\partial \lambda} = Bi \cdot (1 - \Psi) \]  \hspace{1cm} (B-17)

where

\[ \Psi = \frac{c_{O_2}}{c'_{O_2,g}} \]  \hspace{1cm} (B-18)

\[ \lambda = \frac{y}{d_1} \]  \hspace{1cm} (B-19)
Appendix B. Supporting Information _Development of a Flat Membrane Microreactor for Scalable Aerobic Oxidation of Benzyl Alcohol in Flow_

\phi is the Thiele modulus

\[ \phi^2 = \frac{d_1^2 k_{O_2}}{D_{O_2,T}} = \frac{\text{Reaction rate}}{\text{Transverse mass transfer rate}} \]  \hspace{1cm} (B-20)

\textit{Bi} is Biot number

\[ \text{Bi} = \frac{d_1 k_m}{D_{O_2,T}} = \frac{\text{Permeation rate}}{\text{Transverse mass transfer rate}} \]  \hspace{1cm} (B-21)

The solution is

\[ \psi = \frac{\cosh \phi \lambda}{\cosh \phi \frac{d_1}{Bi} \sinh \phi} \]  \hspace{1cm} (B-22)

Permeation-transverse mass transfer effectiveness factor (\( \eta \))

\[ \eta = \int_0^1 \psi d\lambda = \frac{\sinh \phi}{\phi (\cosh \phi + \frac{d_1}{Bi} \sinh \phi)} \]  \hspace{1cm} (B-23)

gives overall effectiveness in terms of oxygen permeation and transverse mass transfer in the packed bed membrane reactor.

The observed global reaction rate is

\[ -R'_{O_2} = \eta k_{O_2} c'_{O_2,g} \]  \hspace{1cm} (B-24)

Multiply by \( \frac{d_1^2}{D_{O_2,T} c'_{O_2,g}} \),

\[ -R'_{O_2} \frac{d_1^2}{D_{O_2,T} c'_{O_2,g}} = \eta k_{O_2} c'_{O_2,g} \frac{d_1^2}{D_{O_2,T} c'_{O_2,g}} = \eta \phi^2 \]  \hspace{1cm} (B-25)

Hence, \( \eta \) can be expressed by a value of the observed global reaction rate, the oxygen concentration in equilibrium with the gas phase, the depth of the liquid channel, and the oxygen transverse mass transfer coefficient.

The Reynolds number (\( Re \)) which is required to calculate the mass transfer
Appendix B. Supporting Information_Development of a Flat Membrane Microreactor for Scalable Aerobic Oxidation of Benzyl Alcohol in Flow

The Reynolds number is defined by

\[ Re = \frac{d_{pe} \rho U}{\mu} \]  

(B-26)

where \( d_{pe} \) is the diameter of the catalyst pellets, \( \rho \) is the density of benzyl alcohol, \( U \) is the axial liquid superficial velocity, \( \mu \) is viscosity of benzyl alcohol.

When \( Re < 1 \) (\( Re \approx 0.01 \) in the flat membrane microreactor), the transverse mass transfer coefficient is approximately equal to the molecular diffusion coefficient.\(^{168}\)

So, in the one-direction model, \( D_{O2,T} \) is calculated by

\[ D_{O2,T} = \frac{\varepsilon}{\tau} D_{O2,m} = \frac{0.5}{1.4} \times 5.47 \times 10^{-9} = 1.95 \times 10^{-9} \text{ m}^2/\text{s} \]  

(B-27)

where \( \tau \) is the tortuosity of the bed (1.4 for spheres),\(^{168}\) \( \varepsilon \) is the bed porosity, which is estimated by\(^{165}\)

\[ \varepsilon = 0.38 + 0.073 \left( 1 + \left( \frac{d_h}{d_{pe}} - 2 \right)^2 \right) \]  

(B-28)

where \( d_h \) is the hydraulic diameter of the liquid flow channel, \( d_{pe} \) is the catalyst pellet diameter.

Oxygen diffusion coefficient in the membrane (\( D_{O2,mem} \)) is calculated by\(^{119}\)

\[ D_{O2,mem} = \frac{T}{273K} \cdot (1.01 \text{ bar}) \cdot \kappa = 1.25 \times 10^{-9} \text{ m}^2/\text{s} \]  

(B-29)

where \( T \) is the reaction temperature, \( \kappa \) is the membrane permeability (1140 barrer)\(^{114}\).

Note that the obtained \( D_{O2,mem} \) is based on the oxygen concentration in the gas phase. To calculate the \( k_m \), the oxygen flux at the membrane-liquid interface

\[ D_{O2,T} \frac{\partial c_{O2}}{\partial y} = D_{O2,mem} \frac{\partial c_{O2,g}}{\partial y} = D_{O2,mem} \cdot H \frac{\partial c'_{O2,g}}{\partial y} = D'_{O2,mem} \frac{\partial c'_{O2,g}}{\partial y} \]  

(B-30)
So, $D_{O_2,\text{mem}}'$ can be assumed as a hypothetical oxygen diffusion coefficient based on the oxygen concentration in the liquid phase.

So,

$$k_m = \frac{D_{O_2,\text{mem}}'}{d_m} = \frac{D_{O_2,\text{mem}} H}{d_m} = 4.83 \times 10^{-5} \text{ m/s}$$  \hspace{1cm} (B-31)$$

where $d_m$ is the thickness of the membrane $H$ is the dimensionless Henry’s law constant for oxygen in toluene ($H = 2.7$ at $120$ °C).$^{149}$

The oxygen reaction rate coefficient ($k_{O_2}$) is estimated using the TOF$_O$ obtained in batch reactor (~8800 h$^{-1}$ for benzyl alcohol at 2 bar oxygen pressure).$^{26}$ The oxygen concentration in batch reactor is assumed to be equal to the saturation concentration ($c_{O_2,\text{sat}}$, estimated based on the oxygen solubility in toluene$^{149}$),

$$-R_{O_2} = k_{O_2} c_{O_2,\text{sat}}$$  \hspace{1cm} (B-32)$$

$$-R_{O_2} = \frac{1}{2} \cdot \text{TOF}_O \cdot \frac{n_{\text{metal}}}{m_{\text{cat}}} \cdot \frac{m_{\text{cat}}}{V_{\text{bed}}} = 67.6 \text{ mol/(m}^3 \cdot \text{s})$$  \hspace{1cm} (B-33)$$

where $n_{\text{metal}}$ is the moles of the metals contained in the packed catalyst, $m_{\text{cat}}$ is the mass of packed catalyst, $V_{\text{bed}}$ is the volume of the packed bed.

So,

$$k_{O_2} = \frac{-R_{O_2}}{c_{O_2,\text{sat}}} = 3.02 \text{ s}^{-1}$$  \hspace{1cm} (B-34)$$

The Biot number (Bi) in current flat membrane reactor is

$$Bi = \frac{d_1 k_m}{D_{O_2,T}} = \frac{1 \times 10^{-3} \text{ m} \times 4.83 \times 10^{-5} \text{ m/s}}{1.95 \times 10^{-9} \text{ m}^2/\text{s}} = 24.8$$  \hspace{1cm} (B-35)$$

Thiele modulus ($\phi$) is calculated with the estimated reaction rate coefficient
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\[ \frac{d^2 k_{O_2}}{D_{O_2} \tau} = \frac{(1 \times 10^{-3} \text{m})^2 \times 3.02 \text{s}^{-1}}{1.95 \times 10^{-9} \text{m}^2/\text{s}} = 1548.7 \quad \text{(B-36)} \]

\[ \phi = 39.4 \quad \text{(B-37)} \]

The permeation-transverse mass transfer catalytic effectiveness factor (\( \eta \)) is calculated by

\[ \eta = \frac{\sinh \phi}{\phi (\cosh \phi + \frac{\phi}{B_i} \sinh \phi)} = \frac{\sinh 39.4}{39.4 \times (\cosh 39.4 + \frac{39.4}{24.8} \sinh 39.4)} = 1.0\% \quad \text{(B-38)} \]
Figure B-8 Oxidation of benzyl alcohol on Au-Pd/TiO$_2$ catalyst with air in flat membrane microreactor. Reaction conditions: Au-Pd/TiO$_2$ catalyst (90-125 μm), 1.0 g; neat benzyl alcohol, 25 μL/min; catalyst contact time, 2300 g$_{cat}$·s/g$_{alcohol}$; air pressure, 8.2 bar; liquid pressure, 10 bar; reaction temperature, 120 ºC.
Appendix C. Supporting Information_Aerobic Oxidation of Benzyl Alcohol on Ru/Al₂O₃ Catalyst in a Flat Membrane Microchannel Reactor
C1. Measurement of Organics Pervaporation

Based on the solution-diffusion model, the driving force for the pervaporation was determined by the vapour pressure difference between the two sides of the membrane. To measure the benzyl alcohol pervaporation, the reactor temperature was set at 373 K and the pressure in the liquid phase was maintained at 8.4 bar, with atmospheric pressure in the gas phase. Nitrogen flow (2 mL/min at STP) was introduced into the gas inlet of the reactor, and the gas outlet of the reactor was directly connected to a vial located in a cold trap (298 K and 1 bar). Limited amount of benzyl alcohol condensation was observed in the vial within several hour collection. To minimize the experimental error, the condensation was collected for ~15 h, and the amount of the collected benzyl alcohol in the vial ($\dot{m}_c$) was averaged to be ~0.8 mg/h. Assuming the gas exiting from the vial was saturated with benzyl alcohol at 298 K and 1 bar, the mass flow rate of benzyl alcohol vapour in the outlet gas ($\dot{m}_{\text{vapour}}$) from the vial can be calculated by

$$\dot{m}_{\text{vapour}} = \frac{\nu_{N2}}{P_{\text{atm}} - P_V} \cdot \frac{P_V}{P_{\text{atm}}} \cdot \frac{M_L}{\nu_{O2}}$$  \hspace{1cm} (C-1)

where $\nu_{N2}$ is the nitrogen flow rate (2 mL/min at standard temperature and pressure (STP, 273 K and 1 bar), $P_V$ is the vapour pressure of benzyl alcohol at 298 K ($1.1 \times 10^{-4}$ bar), $P_{\text{atm}}$ is the atmospheric pressure (1 bar), $M_L$ is molar mass of benzyl alcohol (108.1 g/mol), $\nu_{O2}$ is molar volume of toluene vapour at STP (22400 mL/mol). So, The mass flow rate of benzyl alcohol vapour in the outlet gas from the vial (298 K and 1 bar) was calculated to be 0.06 mg/h. The total amount of benzyl alcohol pervaporation through the membrane ($\dot{m}_{\text{total}}$) is calculated by

$$\dot{m}_{\text{total}} = \dot{m}_c + \dot{m}_{\text{vapour}}$$  \hspace{1cm} (C-2)
which was \( \sim 0.9 \) mg/h.

The volume flow rate of benzyl alcohol vapour in the gas flow channel \( (υ_{\text{BnOH}}) \) at STP is calculated by

\[
υ_{\text{BnOH}} = \frac{m_{\text{total}} v_{\text{O}_2}}{M_L (C - 3)}
\]

which was \( 3.1 \times 10^{-3} \) mL/min. So, the toluene vapour partial pressure in the gas phase outlet was \( 1.6 \times 10^{-3} \) bar. Since the benzyl alcohol vapour pressure is \( 0.02 \) bara at \( 373 \) K,\(^{188}\) the benzyl alcohol vapour pressure difference across the membrane was very close to the benzyl alcohol vapour pressure, and the measured benzyl alcohol pervaporation rate appreciated to the maximum rate at \( 373 \) K. The maximum pervaporation rate of benzaldehyde was also measured with the same method, which was \( \sim 6 \) mg/h.

So, the maximum volume ratio of organic vapour in the gas phase during the reaction \( (ϕ_{\text{vapour}}) \) is estimated with the pervaporation rate of benzaldehyde \( (υ_{\text{Bal}}) \).

\[
ϕ_{\text{vapour}} = \frac{υ_{\text{Bal}}}{υ_{\text{O}_2} + υ_{\text{Bal}}} \times 100\%
\]

When the oxygen flow rate \( (υ_{\text{O}_2}) \) is 5 mL/min at STP, \( ϕ_{\text{BnOH}} \) is less than 0.4 vol\%, which is lower than the lower explosive limit (1 vol\%). Due to the small amount of liquid pervaporation as compared to the liquid inlet flow rate (> 1 mg/min), the loss of the liquid due to pervaporation was ignored.
C2. Diffusion Coefficient Calculation

The molecular diffusion coefficient of benzyl alcohol in the fluid ($D_{\text{BnOH}}$) was estimated as the diffusion coefficient of benzyl alcohol in benzyl alcohol with the Reddy-Doraiswamy correlation\(^{163}\)

$$D_{\text{BnOH}} = 1 \times 10^{-16} \frac{T \sqrt{M_{\text{BnOH}}}}{\mu V_{\text{BnOH}}^{2/3}}$$ \hspace{1cm} (C-5)

where $T$ is the temperature in K, $M_T$ is the molecular weight of benzyl alcohol in g/mol, $\mu$ is the viscosity of benzyl alcohol in Pa·s, $V_{\text{BnOH}}$ is the molar volume of benzyl alcohol at normal boiling point in m\(^3\)/kmol\(^1\). The molecular diffusion coefficient of benzyl alcohol in toluene was also calculated, which was found to be ~2.5 times of that in benzyl alcohol.

The molecular diffusion coefficient of oxygen ($D_{O2}$) in benzyl alcohol was calculated based on the Wilke-Chang correlation\(^{164}\)

$$D_{O2} = 1.1728 \times 10^{-16} \frac{T \sqrt{\chi M_{\text{BnOH}}}}{\mu V_{O2}^{6.6}}$$ \hspace{1cm} (C-6)

where $T$ is the temperature in K, $\chi$ is association factor of benzyl alcohol, 1 for non-associated solvents, $M_{\text{BnOH}}$ is the molecular weight of the benzyl alcohol in g/mol, $\mu$ is the viscosity of benzyl alcohol in Pa·s, $V_{O2}$ is the molar volume of oxygen at normal boiling point in m\(^3\)/kmol\(^1\).

The effective diffusion coefficient of species $i$ (oxygen or benzyl alcohol) in the catalyst was calculated based by\(^{165}\)

$$D_{i,\text{eff}} = \frac{\varepsilon_p}{\tau_p} D_i$$ \hspace{1cm} (C-7)

where $\varepsilon_p$ is catalyst pellet porosity, $\tau_p$ is the tortuosity of catalyst pellet assumed to
be $1/\epsilon_p$.

When $\text{Re} < 1$ (in the flat membrane microreactor $\text{Re} < 0.01$), the axial and transverse dispersion coefficients ($D_{i,A}$ and $D_{i,T}$) are both approximately equal to the molecular diffusion coefficient.\textsuperscript{168} So, in the homogeneous model, the dispersion coefficients are calculated by

$$D_{i,A} = D_{i,T} = \frac{\varepsilon_b D_i}{\tau_b}$$  \hspace{1cm} (C-8)

where $\varepsilon_b$ is the bed void fraction, $\tau_b$ is the tortuosity of the bed (1.4 for spheres).\textsuperscript{168}
C3. Reaction rate discrimination

Based on the literature\textsuperscript{11, 40-42}, the reaction rate of benzyl alcohol oxidation on Ru-based catalyst was suggested to be a zero to first order in benzyl alcohol, and a zero to partial order in oxygen. Since the oxygen pressure has shown significant effect on the conversion, the reaction rate with a zero order in oxygen is excluded. Initially, four different reaction rates were considered. The details are shown in Table C-1, and the corresponding reaction rate coefficient (k) was estimated to make the modelling conversion equal to the average conversion at the stable period in Figure 7-3 (40%) under the same reaction conditions (liquid flow rate, 2 μL/min; catalyst contact time, 48 g\textsubscript{cat}.min/g\textsubscript{BnOH}; oxygen pressure, 8.0 bar). Then, these different reaction rates with the estimated k were used in other reaction conditions (different liquid flow rates and oxygen pressures), and the results are compared with the experimental in Figure C-1&2. No obvious difference in terms of fitting the experimental data was observed among these four reaction rate expressions, though the conversions obtained with R2 and R3 at liquid flow rate less than 0.8 μL/min were higher than 100%. Since the oxygen concentration likely varied axially and transversely in the catalyst bed, the ratio of oxygen concentration to benzyl alcohol concentration also varied. To include both oxygen and benzyl alcohol concentrations in the reaction rate, the reaction rate R1, assuming a first order in benzyl alcohol and 0.5 order in oxygen, was used in the following modelling.

\[
R_{\text{BnOH},p} = -kc^{\alpha}_{\text{BnOH},p}c^{\beta}_{\text{O2},p} \quad (C-9)
\]

\[
R_{\text{O2},p} = -\frac{1}{2}k'c^{\alpha}_{\text{BnOH},p}c^{\beta}_{\text{O2},p} \quad (C-10)
\]
**Table C-1** Different reaction rates used in the model and estimated reaction rate coefficients

<table>
<thead>
<tr>
<th></th>
<th>α</th>
<th>β</th>
<th>k</th>
</tr>
</thead>
<tbody>
<tr>
<td>R1</td>
<td>1</td>
<td>0.5</td>
<td>$3.9 \times 10^{-4} [\text{m}^{3/2}/(\text{mol}^{1/2} \cdot \text{s})]$</td>
</tr>
<tr>
<td>R2</td>
<td>0</td>
<td>0.5</td>
<td>$2.7 [\text{mol}^{1/2}/(\text{m}^{3/2} \cdot \text{s})]$</td>
</tr>
<tr>
<td>R3</td>
<td>0</td>
<td>1</td>
<td>0.76 [s(^{-1})]</td>
</tr>
<tr>
<td>R4</td>
<td>1</td>
<td>1</td>
<td>$1.1 \times 10^{-4} [\text{m}^{3}/(\text{mol} \cdot \text{s})]$</td>
</tr>
</tbody>
</table>

**Figure C-1** Comparison of benzyl alcohol conversion at different liquid flow rates obtained from experiment and different reaction rates. Reaction conditions: Ru/Al\(_2\)O\(_3\) catalyst (90-125 \(\mu\)m), 100 mg; neat benzyl alcohol; oxygen pressure, 8 bar; liquid pressure, 9.0 bar; reaction temperature, 373 K.
Figure C-2 Comparison of benzyl alcohol conversion at different oxygen pressures obtained from experiment and different reaction rates. Reaction conditions: Ru/Al2O3 catalyst (90-125 µm), 100 mg; neat benzyl alcohol, 1 µL/min; catalyst contact time, 96 g·cat·min/g_BnOH; liquid pressure, 9.0 bar; reaction temperature, 373 K.
Appendix C. Supporting Information_Aerobic Oxidation of Benzyl Alcohol on Ru/Al2O3 Catalyst in a Flat Membrane Microchannel Reactor

C4. Energy Balance in the Model

The heat balance equation in the unpacked zone of the liquid channel is

\[ u_x \rho C_p \frac{\partial T}{\partial x} - \lambda_{BnO\!H} \left( \frac{\partial^2 T}{\partial x^2} + \frac{\partial^2 T}{\partial y^2} \right) = 0 \]  \hspace{1cm} (C-11)

where \( u_x \) is the liquid velocity in the axial direction, \( \rho \) is the density of benzyl alcohol, \( C_p \) is the heat capacity of benzyl alcohol, \( \lambda_{BnO\!H} \) is the thermal conductivity of benzyl alcohol.

Assuming equality of temperature in both fluid and solid phases, the heat balance equation in the catalyst bed is

\[ u_x \rho C_p \frac{\partial T}{\partial x} - \left( \lambda_{b,A} \frac{\partial^2 T}{\partial x^2} + \lambda_{b,T} \frac{\partial^2 T}{\partial y^2} \right) - Q_R = 0 \]  \hspace{1cm} (C-12)

where \( \lambda_{b,A} \) is axial thermal conductivity of catalyst bed, \( \lambda_{b,T} \) is transverse thermal conductivity of catalyst bed. Due to the low Re, both the axial and transverse thermal conductivities are approximately equal to the effective conductivities of catalyst bed (\( \lambda_b \)).  

\[ \lambda_b = \theta \lambda_s + (1 - \theta) \lambda_{BnO\!H} \]  \hspace{1cm} (C-13)

\[ \theta = (1 - \varepsilon_b)(1 - \varepsilon_p) \]  \hspace{1cm} (C-14)

where \( \lambda_s \) is the thermal conductivity of Al2O3; \( \theta \) is the volume ratio of solid in the bed, calculated based on the catalyst bed void fraction (\( \varepsilon_b \)) and catalyst pellet porosity (\( \varepsilon_p \)).

\( Q_R \) is the heat of reaction

\[ Q_R = \int_{V_p}^{V_p} R_{BnO\!H} \cdot \Delta H \cdot dV_p \]  \hspace{1cm} \( (1 - \varepsilon_b) \)  \hspace{1cm} (C-15)
where $V_p$ is the volume of catalyst pellet, $R_{BnOH,p}$ is the benzyl alcohol reaction rate per unit volume of catalyst particles, $\Delta H$ is molar reaction enthalpy of oxidation reaction, $\varepsilon_b$ is catalyst bed void fraction.

The heat balance equation in the membrane is

$$\lambda_m \left( \frac{\partial^2 T}{\partial x^2} + \frac{\partial^2 T}{\partial y^2} \right) = 0 \quad (C-16)$$

where $\lambda_m$ is thermal conductivity of the membrane.

Since the temperature of the reactor was experimentally measured to be 373 K with a thermocouple 1 mm away from the liquid channel (stainless steel was used between the thermocouple and the liquid channel), the boundary temperature of the liquid channel was assumed to be constant at 373 K during the reaction. Boundary conditions used are

$$at \ x = 0, \ T = 293 \ K \quad (C-17)$$

$$at \ x = l_2, \ \frac{\partial T}{\partial x} = 0 \quad (C-18)$$

$$at \ y = 0, \ T = 373 \ K \quad (C-19)$$

$$at \ y = d_2, \ \frac{\partial T}{\partial y} = 0 \quad (C-20)$$

The values of the variables used in the model are shown in Table C-2. The simulated temperature profiles for 48 g_{cat·min/g_BnOH} catalyst contact time and 8 bar oxygen pressure (conversion: 40%) are shown in Figure C-3a. Nearly isothermal conditions were observed in the flat membrane reactor and the temperature difference (between the actual and set temperature (373 K)) across the catalyst bed was less than 0.1 K (shown in Figure C-3b). This is in accordance with the
experimentally measured temperature difference between the gas and liquid flow plates (< 3 K). The temperature difference was still less than 0.1 K when using a reaction rate coefficient including an activation energy (Ea) of 51.4 kJ/mol (not shown). These results showed the excellent heat transfer in the flat membrane microchannel reactor, which was due to the small reactor channel dimensions, since the heat transfer coefficient is inversely proportional to the hydraulic diameter. Note that the temperature rise under adiabatic conditions (all the boundaries are insulated except for the inlet and the outlet) was higher than 300 K (shown in Figure C-4). This indicates the importance of heat transfer in the reactor for the highly exothermic oxidation of alcohols.

*Table C-2* List of values of the variables used in the model

<table>
<thead>
<tr>
<th>Symbols</th>
<th>Values</th>
<th>Units</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>$C_P$</td>
<td>2067.9</td>
<td>J/(kg·K)</td>
<td>Heat capacity of benzyl alcohol$^{191}$</td>
</tr>
<tr>
<td>$\Delta H$</td>
<td>-187</td>
<td>kJ/mol</td>
<td>Molar reaction enthalpy of oxidation reaction$^{11}$</td>
</tr>
<tr>
<td>$\theta$</td>
<td>0.2</td>
<td>-</td>
<td>Averaged volume of Al$_2$O$_3$ in the bed</td>
</tr>
<tr>
<td>$\lambda_a$</td>
<td>30</td>
<td>W/(m·K)</td>
<td>Thermal conductivity of Al$_2$O$_3$$^{192}$</td>
</tr>
<tr>
<td>$\lambda_b$</td>
<td>6.0</td>
<td>W/(m·K)</td>
<td>Effective conductivities of catalyst bed</td>
</tr>
<tr>
<td>$\lambda_{BnOH}$</td>
<td>0.154</td>
<td>W/(m·K)</td>
<td>Thermal conductivity of benzyl alcohol$^{193}$</td>
</tr>
<tr>
<td>$\lambda_m$</td>
<td>0.20</td>
<td>W/(m·K)</td>
<td>Thermal conductivity of Teflon AF-2400$^{194}$</td>
</tr>
</tbody>
</table>
Appendix C. Supporting Information_Aerobic Oxidation of Benzyl Alcohol on Ru/Al2O3 Catalyst in a Flat Membrane Microchannel Reactor

Figure C-3 Simulated temperature map and profile in the reactor. a) Temperature map in the reactor; b) temperature difference with respect to the set reaction temperature (373 K) across transverse direction at x = 50 mm. Reaction conditions: Ru/Al2O3 catalyst (90-125 £m), 100 mg; neat BnOH, 2 µL/min; catalyst contact time, 48 g_cat·min/g_BnOH; oxygen pressure, 8 bar; reaction temperature, 373 K; Benzyl alcohol conversion, 40%.
Figure C-4 Simulated adiabatic temperature map in the reactor. Reaction conditions:
Ru/Al$_2$O$_3$ catalyst (90-125 μm), 100 mg; neat BnOH, 2μL/min; catalyst contact time, 48 g$_{cat}$·min/g$_{BnOH}$; oxygen pressure, 8 bar; reaction temperature, 373 K; Benzyl alcohol conversion, 40%.
Appendix C. Supporting Information_Aerobic Oxidation of Benzyl Alcohol on Ru/Al2O3 Catalyst in a Flat Membrane Microchannel Reactor

C5. Concentration Profiles in the Reactor

Figure C-5 Benzyl alcohol concentration profiles in the catalyst particles at x = 50 mm, obtained from modelling. Reaction conditions: Ru/Al2O3 catalyst (90-125 μm), 100 mg; neat benzyl alcohol, 2 μL/min; catalyst contact time, 48 g_cat·min/g_BnOH; oxygen pressure, 8 bar; reaction temperature, 373 K.
Figure C-6 Oxygen concentration map and profiles in the bulk liquid phase within the reactor, obtained from modelling. a) Oxygen concentration map in the reactor, b) Oxygen concentration profiles across the transverse direction at different axial positions. Reaction conditions: Ru/Al₂O₃ catalyst (90-125 μm), 100 mg; neat benzyl alcohol, 1 μL/min; catalyst contact time, 48 g·cat·min/g·BnOH; oxygen pressure, 8 bar; reaction temperature, 373 K; channel depth, 0.5 mm.
Appendix D. Supporting Information_Heterogeneously Catalysed Aerobic Oxidation of Benzyl Alcohol in a Continuous Stirred Membrane Reactor System
D1. Design and Analysis of the Stirred Membrane Reactor with Internal Membrane

A schematic of the stirred membrane reactor with internal membrane is shown in the Figure D-1. Assuming perfect mixing in the reactor (ignoring internal/external mass transfer) and a first order reaction rate in oxygen, the mole balance of oxygen in the reactor can be shown as

\[
(Oxygen\ in) - (Oxygen\ out) + (Generation) = 0 \quad (D-1)
\]

\[
k_m \cdot (c_{O_2,sat} - c_{O_2,out}) \cdot S_m - c_{O_2,out} \cdot v_{out} - k \cdot c_{O_2,out} \cdot V = 0 \quad (D-2)
\]

\[
k_m \cdot S_m \cdot c_{O_2,sat} = k_m \cdot S_m \cdot c_{O_2,out} + c_{O_2,out} \cdot v_{out} + k \cdot c_{O_2,out} \cdot V \quad (D-3)
\]

where \( k_m \) is oxygen mass transfer coefficient in the membrane, \( S_m \) is the membrane area, \( c_{O_2,sat} \) is the oxygen saturation concentration, \( c_{O_2,out} \) is the oxygen concentration at the outlet of the reactor, \( v_{out} \) is the flow rate at the outlet of the reactor, \( k \) is the reaction rate coefficient, \( V \) is the reactor volume.

So, the catalyst utilization factor can be expressed as

\[
U = \frac{k \cdot c_{O_2,out}}{k \cdot c_{O_2,sat}} = \frac{k_m S_m}{k_m S_m + v_{out} + k \cdot V} = \frac{1}{1 + \frac{v_{out}}{k_m S_m + k \cdot V}}
\]  

(D-4)
D2. Estimation of Benzyl Alcohol Pervaporation Rate

Based on the solution-diffusion model\textsuperscript{187}, the driving force for the pervaporation was determined by the vapour pressure difference between the two sides of the membrane.

\[ J_i = \frac{\kappa}{l} \cdot A \cdot \Delta P \]  

where \( J_i \) is the flux of \( i \) through the membrane, \( \kappa \) is the permeability, \( l \) is the thickness of the membrane, \( A \) is the membrane area, \( \Delta P \) is the vapour pressure difference between the two sides of the membrane. When the vapour pressure in the gas phase was equal to the liquid vapour pressure, the liquid permeation due to the hydrostatic pressure difference can be ignored.

The vapour pressure of benzyl alcohol and benzaldehyde at 298 K is \( 1.3 \times 10^{-4} \) bar and \( 1.6 \times 10^{-3} \) bar, respectively.\textsuperscript{188} So, the maximum amount of liquid loss due to the liquid pervaporation was estimated with the benzaldehyde vapour pressure.

\[ \dot{m}_{\text{BnOH}} = \frac{u_{O2}}{P_{atm} - P_{V}} \cdot P_{\text{Bald}} \cdot \frac{M_L}{\dot{V}_{O2}} \]  

where \( u_{O2} \) is the oxygen flow rate (5 mL/min at standard temperature and pressure (STP, 273 K and 1 bar), \( P_{\text{Bald}} \) is the vapour pressure of benzaldehyde at 298 K (1.6 \times 10^{-3} \) bar),\textsuperscript{188} \( P_{atm} \) is the atmospheric pressure (1 bar), \( M_L \) is molar mass of benzaldehyde (106.1 g/mol), \( \dot{V}_{O2} \) is molar volume of toluene vapour at STP (22400 mL/mol). So, The mass flow rate of benzaldehyde vapour in the outlet gas was calculated to be 0.04 mg/h, which is much less than the liquid inlet mass flow rate.
Figure D-2 Picture of the liquid sample: a) slurry in the reactor, b) liquid effluent from the reactor system; c) slurry in the reactor observed with a microscope (diluted with benzyl alcohol)  d) liquid effluent from the reactor system observed with a microscope; e) pure benzyl alcohol observed with a microscope.
Figure D-3 Measured pressure through the pressure sensors $P_1$ (red line) and $P_2$ (white line) during the reaction from ~30 to ~90 h. Reaction conditions: Ru/Al$_2$O$_3$ catalyst, 0.5 g; neat benzyl alcohol, 10 $\mu$L/min; recycle flow rate, 1 mL/min; BPR pressure at the gas vent, 5.0 bar; BPR pressure at the liquid outlet, 6.0 bar; reaction temperature, 373 K; contactor temperature, 298 K; reactor volume, 25 mL.
Figure D-4 Catalyst deposition on the filter after ~7 h experiment with 1 mL/min recycle flow rate and ~20 g/L Ru/Al₂O₃ catalyst in benzyl alcohol.
D3. New Reactor Design with Filters at both Inlet and Outlet

*Figure D-5* New design of continuous stirred reactor with filters at both inlet and outlet.

*Figure D-6* Mechanical drawing of the reactor bottom plate with O-ring (unit, mm).
Appendix D. Supporting Information_Heterogeneously Catalysed Aerobic Oxidation of Benzyl Alcohol in a Continuous Stirred Membrane Reactor System

Figure D-7 Mechanical drawing of the reactor gasket and fixing plate (unit, mm).

Figure D-8 Mechanical drawing of the reactor top plate with filters (unit, mm).
The new design of the continuous stirred reactor consists of several layers, which are bottom plate (316 stainless steel), O-ring (Kalrez®, Dupont), gasket fixing plate (alumina), gasket (Kalrez®, Dupont), filters (304 stainless steel; Mott, grade 0.1), and top plate (316 stainless steel) (shown in Figure D-5). The dimensions are shown in Figure D-6~8, and the picture of the reactor is shown in Figure D-9. A PTFE coated magnetic bar (diameter: 6 mm; length: 28 mm) will be placed within the bottom plate to stir the slurry, and the bottom plate will be sealed by the O-ring placed between the bottom plate and the gasket fixing plate. An opening (2 mm away to the chamber) was machined in the bottom plate for thermocouple insertion. The reactor inlet and outlet were machined in the top plate, and the filters were inserted into the top plate for catalyst retaining. The filters will be sealed with the gasket (~2 mm wide contacting area covering the edge of the filters) fixed on the gasket fixing plate. To avoid catalyst accumulation on the filter, the inlet and outlet of the reactor will be switched by using two two-way valves or changing the flow direction of the recycle pump during the reaction.
D4. Reaction Rate Discrimination

\[-R = k c_{\text{BnOH, out}}^{\alpha} c_{\text{O}_2, \text{out}}^{\beta}\]  \hspace{1cm} \text{(D-7)}

Table D-1 Different reaction rates used in the model and estimated reaction rate coefficients

<table>
<thead>
<tr>
<th></th>
<th>(\alpha)</th>
<th>(\beta)</th>
<th>(k)</th>
</tr>
</thead>
<tbody>
<tr>
<td>R1</td>
<td>1</td>
<td>0.5</td>
<td>(3.6 \times 10^{-7} \text{ [m}^{3/2}/\text{(mol}^{1/2} \cdot \text{s}]}))</td>
</tr>
<tr>
<td>R2</td>
<td>0</td>
<td>0.5</td>
<td>(3.0 \times 10^{-3} \text{ [mol}^{1/2}/\text{(m}^{3/2} \cdot \text{s}]}))</td>
</tr>
<tr>
<td>R3</td>
<td>1</td>
<td>1</td>
<td>(1.4 \times 10^{-7} \text{ [m}^{3/2}/\text{mol} \cdot \text{s}])</td>
</tr>
<tr>
<td>R4</td>
<td>0</td>
<td>1</td>
<td>(1.2 \times 10^{-3} \text{ [s}^{-1}])</td>
</tr>
</tbody>
</table>

Table D-2 Comparison of the simulated results obtained with different reaction rates.

Reaction conditions: Ru/Al\(_2\)O\(_3\) catalyst, 0.5 g; reactor volume, ~25 mL; neat benzyl alcohol; recycle flow rate, 1 mL/min; liquid pressure measured at the upstream of the contactor, 6-16 bar; BPR pressure at the liquid outlet, 6 bar; reaction temperature, 373 K; contactor temperature, 298K. \(P_{\text{O}_2}\), oxygen pressure in the contactor; \(v_{\text{IN}}\), the reactor system inlet flow rate; \(X_{\text{exp}}\), the conversion from the experiment; \(X_{\text{sim}}\), the conversion from the simulation.

<table>
<thead>
<tr>
<th>Entry</th>
<th>(P_{\text{O}_2}) / bar</th>
<th>(v_{\text{IN}}) [(\mu)L/min]</th>
<th>(X_{\text{exp}}) [%]</th>
<th>(X_{\text{sim}}) [%]</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td>R1</td>
<td>R2</td>
</tr>
<tr>
<td>1</td>
<td>5.0</td>
<td>10</td>
<td>12.5</td>
<td>12.6</td>
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<td>8.0</td>
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<td>8.8</td>
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<tr>
<td>3</td>
<td>9.1</td>
<td>30</td>
<td>5.0</td>
<td>6.5</td>
</tr>
</tbody>
</table>

Similar to the reaction rate discrimination in Chapter 7, four different reaction rates were initially assumed in the model. The details are shown in Table D-1, and the corresponding reaction rate coefficient (k) was estimated to make the simulated result equal to the average conversion at the stable period in Figure 8-7 under the same reaction conditions (shown in Table D-2, entry 1). Then, the different reaction rates with the estimated k were used in other reaction conditions and the results are compared in Table D-2. Though R1 and R2 present slightly better simulation results in terms of the difference between the experimental and simulated data, no obvious
difference is observed among these reaction rates. To allow for a dimensionless analysis of the various parameters in the reactor, the reaction rate was assumed as a zero order in benzyl alcohol and a first order in oxygen.
Appendix E. Oxidation of Glycerol

As a compound with three hydroxyl groups, glycerol is a potential raw material for various chemicals and intermediates.\textsuperscript{177} Nowadays, the rapid growth of the biodiesel industry is generating a surplus of crude glycerol in the market.\textsuperscript{195} This makes conversion of bio-derived glycerol to value-added chemicals not merely environmentally friendly but also potentially economical. Until now, only several processes for the valorisation of glycerol have already been commercialised, including the chlorination of glycerol to epichlorohydrin, the esterification of glycerol to monoacyle and diachl esters, and the reforming of glycerol to yield syngas. As one of the potential applications, catalytic oxidation of glycerol can synthesize fine chemicals with high value, which makes it more economically promising.\textsuperscript{177} The heterogeneous oxidation of glycerol with molecular oxygen is also deemed to be one of the most attractive routes from the industrial point of view.

However, complex reaction pathways are involved in oxidation of glycerol and lead to a large number of products, such as glyceric acid, trartronic acid, glycolic acid, oxalic acid, and formic acid (shown in Figure E-1). The general consensus about the mechanism has not been proposed, and one of the most challenging problems is how to achieve a high yield or selectivity to a specific product.\textsuperscript{177, 196} This is mainly affected by the synthesis methods of supported catalysts, reaction temperature, pH, partial pressure of oxygen, concentration and physicochemical properties of the catalyst used, mass-transfer conditions and type of reactors.\textsuperscript{197}

In this study, 10 wt% Pd/Al\textsubscript{2}O\textsubscript{3} catalyst, 5 wt% Ru/Al\textsubscript{2}O\textsubscript{3} catalyst, and 1 wt% Au-Pd/TiO\textsubscript{2} catalyst (Au to Pd weight ratio, 5:95) (provided by Johnson Matthey) were
tested in packed bed microreactor for oxidation of glycerol. The experimental set-up was the same as used in Chapter 3, and the same experimental procedure was performed as used in oxidation of cinnamyl alcohol. The effluent from the reactor was analysed with a HPLC (Jasco LC-2000 series) equipped with a refractive index detector (RI-4030) and an ultraviolet-visible detector (UV-4075). The HPLC conditions were MetaCarb 67H (300 × 6.5 mm); oven temperature, 35 ºC; wavelength, 210 nm; mobile phase (0.02 M H₂SO₄ in water) flow rate, 0.5 mL/min.

Six products (highlighted in the Figure E-1), which were oxalic acid (OA), tartronic acid (TA), glyceric acid (GEC), glycolic acid (GOC), lactic acid (LA), and formic acid (FA) were analysed in the HPLC along with glycerol. Except for these products, three unknown peaks were also observed in HPLC profiles (shown in Figure E-2), which could be dihydroxyacetone (DHA), hydroxypyruvic acid (HPA), or other products shown in Figure E-1. The carbon balance was found to be less than 80% at high conversions.
Appendix E. Oxidation of Glycerol

Figure E-1 Reaction network of glycerol oxidation\textsuperscript{177} Reproduced from ref. 177 with permission from The Royal Society of Chemistry.

Figure E-2 HPLC profiles with peaks of unknown products (marked with ?).
Glycerol conversion ($X$) and selectivity ($S_i$) of product $i$ were calculated according to the following equations:

$$X = \frac{c_{alcohol, in} - c_{alcohol, out}}{c_{alcohol, in}} \times 100\%$$  \hspace{1cm} (E-1)

where $c_{alcohol, in}$ and $c_{alcohol, out}$ are the concentration of glycerol at the reactor inlet and outlet, respectively.

$$S_i = \frac{c_i \nu_i}{\sum c_i \nu_i} \times 100\%$$  \hspace{1cm} (E-2)

where $c_i$ is the concentration of $i$ at the outlet, $\nu_i$ is the number of moles of carbon in 1 mole of product $i$.

As shown in Figure E-3, a 15% conversion of glycerol was observed on Ru/Al$_2$O$_3$ catalyst in the first sample (0.5 h), and the conversion decreased to zero in 3 h, indicating the catalyst deactivation. The selectivites were observed to be roughly constant but all lower than 30%. The Pd/Al$_2$O$_3$ catalyst shown a higher initial conversion (55%) as compared to Ru/Al$_2$O$_3$ catalyst, but catalyst deactivation was also observed as the conversion decreased to 12% within 3 h (shown in Figure E-4). Tartronic acid was observed to be the main product in the first sample, but the selectivity was lower than 40%. The selectivities fluctuated with the reaction time, which could be due to the integration errors in HPLC profiles caused by overlapping of H$_2$SO$_4$ and oxalic acid peaks.
Appendix E. Oxidation of Glycerol

Figure E-3 Aerobic oxidation of glycerol on 5 wt% Ru/Al₂O₃ catalyst. Reaction conditions: Ru/Al₂O₃ catalyst, 10 mg; 0.3 M glycerol in water, 10 µL/min; oxygen flow rate, 2 mL/min; NaOH to substrate ratio, 2:1; reaction pressure, 4 bar; reaction temperature, 100 ºC. OA: oxalic acid; TA: tartronic acid; GEC: glyceric acid; GOC: glycolic acid; LA: lactic acid; FA: formic acid.
Appendix E. Oxidation of Glycerol

Figure E-4 Aerobic oxidation of glycerol on 10 wt% Pd/Al$_2$O$_3$ catalyst. Reaction conditions: Pd/Al$_2$O$_3$ catalyst, 10 mg; 0.3 M glycerol in water, 10 µL/min; oxygen flow rate, 2 mL/min; NaOH to substrate ratio, 2:1; reaction pressure, 4 bar; reaction temperature, 100 ºC. OA: oxalic acid; TA: tartronic acid; GEC: glyceric acid; GOC: glycolic acid; LA: lactic acid; FA, formic acid.
Appendix E. Oxidation of Glycerol

Figure E-5 Aerobic oxidation of glycerol on 1 wt% Au-Pd/TiO$_2$ catalyst. Reaction conditions: Au-Pd/TiO$_2$ catalyst, 10 mg; 0.3 M glycerol in water, 10 $\mu$L/min; oxygen flow rate, 1 mL/min; KOH to substrate ratio, 4:1; reaction pressure, 4 bar; reaction temperature, 100 °C. OA: oxalic acid; TA: tartronic acid; GEC: glyceric acid; GOC: glycolic acid; LA: lactic acid; FA, formic acid. Averaged conversion from 2-7 h, 82±4%. Conversion at 21 h with the same catalyst packing, 76%.

As shown in Figure E-5, the Au-Pd/TiO$_2$ catalyst presented high activity and good stability in oxidation of glycerol. The conversion of glycerol fluctuated around 82% within 7 h, with ~40% selectivity to tartronic acid. The conversion was observed to be 76% after 21 h, indicating the good catalyst stability. The TOF was calculated to be 178 h$^{-1}$, which was close to that reported in literature with 0.5 wt%Au-0.5 wt% Pd/TiO$_2$ catalysts (prepared with an impregnation method) in batch reactor at 60 °C and 10 bar oxygen pressure (190 h$^{-1}$). However, the selectivities were different, since glyceric acid was reported in the literature as the main product with a selectivity of 71%. The different selectivities might be caused by the different catalyst composition and different reaction conditions (temperature, oxygen pressure, etc.), as well as different reactor configurations. More experimental results on Au-
Appendix E. Oxidation of Glycerol

Pd/TiO₂ catalyst under different reaction conditions can be found in Jolynne Lee’s thesis.
Appendix F. Oxidation of Furfural

As a C5 heterocyclic aldehyde and a promising platform compound, furfural can be produced in large quantities from the lignocellulosic biomass.\(^{178}\) In this part, oxidation of furfural was performed on Au-Pd/TiO\(_2\) catalyst, in an attempt to produce valuable products, such as maleic acid, succinic acid and furoic acid.

Initially, base-free oxidation of furfural (0.3 M in water) was studied using PFA capillary microreactor packed with 1 wt% Au-Pd/TiO\(_2\) catalyst (Au to Pd weight ratio, 5:95; prepared with a sol-immobilized method in the Hutchings research group). The reaction temperature was up to 50 °C and no conversion of furfural was observed. This indicates the importance of base on this reaction. When adding base (NaOH) into the prepared furfural solution, the colour of the mixture changed. The concentrations of each species in the same sample at different times were investigated with HPLC. The HPLC conditions were MetaCarb 67H (300 × 6.5 mm); oven temperature, 30 ºC; wavelength, 215 nm; mobile phase (1 g/L H\(_3\)PO\(_4\) in water) flow rate, 0.5 mL/min.

<table>
<thead>
<tr>
<th>Peak areas of each species in the same sample at different times</th>
<th>a</th>
<th>b</th>
<th>c</th>
</tr>
</thead>
<tbody>
<tr>
<td>Furoic acid</td>
<td>1,418,768</td>
<td>4,480,661</td>
<td>1,520,005</td>
</tr>
<tr>
<td>Furfuryl alcohol</td>
<td>3,916,990</td>
<td>11,587,154</td>
<td>6,487,723</td>
</tr>
<tr>
<td>Furfural</td>
<td>40,556,518</td>
<td>17,231,794</td>
<td>7,621,124</td>
</tr>
<tr>
<td>Estimated con.</td>
<td>/</td>
<td>57.5%</td>
<td>81.2%</td>
</tr>
</tbody>
</table>

a) Fresh prepared sample with 0.1 M furfural and NaOH (base ratio, 1:1)
b) Prepared sample stored at room temperature for 1 night
c) Prepared sample (8.3 μL/min) passed through 20 mg 1 wt% Au-Pd/TiO\(_2\) catalyst bed at 30 °C and 4 bar oxygen pressure (G/L = 200 at STP) and stored at room temperature for 1 night

As shown in Table F-1, more than half of furfural was converted even at room temperature without passing through the reactor. This is probably caused by the
Appendix F. Oxidation of Furfural

Cannizzaro reaction, where disproportionation of aldehyde was induced by base (shown in Figure F-1).

\[
\text{Furfural} \xrightarrow{\text{base}} \text{Furoic acid} + \text{Furfuryl alcohol}
\]

**Figure F-1** Cannizzaro reaction of furfural to furoic acid and furfuryl alcohol.

**Figure F-2** On-line mixing of furfural and base.

**Figure F-3** Samples obtained at different times. a) Fresh prepared sample via on-line mixing; b) Prepared sample via on-line mixing stored at room temperature for 2 h; Effluent collected at the outlet of the Au-Pd/TiO\(_2\) packed reactor from c) 0-1 h and d) 1-1.5 h; e) Fouling inside the reactor.
To reduce the effect of the Cannizzaro reaction, on-line mixing was performed with two syringes and a T-junction (Figure F-2). The colour of the on-line prepared sample (without passing the catalyst bed) still changed when stored at room temperature (shown in Figure F-3 a&b). This suggests the importance of instant sample analysis. Moreover, unknown precipitation was observed during the reaction (Figure F-3 c&d), which made the reaction unable to be performed in flow. Yellow fouling was also observed on the inner wall of the reactor (Figure F-3 e).
Appendix G. Oxidation of “GSK Diol”

Figure G-1 Oxidation of “GSK diol” (27) to the target product, GSK dialdehyde (20). A&B are the by-products.179 Reprinted with permission from ref. 179. Copyright (2013) American Chemical Society.

Selective oxidation of “GSK diol” 27 to dialdehyde 20 (shown in Figure G-1) has been reported as a greener and more efficient step in the synthesis of the prodrugs of HCV replicase inhibitor, as compared to other uncatalysed routes (using a crystalline HCL salt of a N-benzylpyridazinone intermediate).179 The existing of multiple –OH groups in the diol 27 also brings more challenges for selective oxidation of alcohols.

The HPLC conditions used for identification of reactants and products in oxidation of “GSK diol” were mobile phase, mixture of 20 vol% acetonitrile (MeCN) and 80 vol% water with 0.1% phosphorous acid added; mobile phase flow rate, 0.5 ml/min; column, Puisuit C18 ( 250 × 4.6 mm, 5μm particle size); oven temperature, 30 ºC; UV length, 274 nm; sample volume, 10 μL (sample was diluted and then injected into HPLC via an auto-dilution pump).

The peaks of diol 27 and dialdehyde 20 were identified through injecting the standard sample (provided by the Marsden group at University of Leeds). The other two by-products have not been identified in HPLC. The HPLC conditions seem to require further modification, since the peak of dialdehyde 20 was not very sharp (shown in Figure G-2). The repeatability of the HPLC was observed to be acceptable (error < 5%). Thus, no internal or external standard was used for the analysis.
Initially, the reagents were prepared through simultaneously dissolving diol and KOH in a mixture of MeCN and water, and fed into the packed bed microreactor using one single syringe. However, when the prepared reagents were kept in the volumetric flask at room temperature, the conversion was found to increase from 0 to 18.5% after 5 h and further to 82.4% after 72 h. This indicates that some non-catalysed reactions occurred after the mixing of the diol 27 with base. The products were neither the monoaldehyde nor the dialdehyde, according to the NMR/HPLC-MS analysis carried out by our collaborators at University of Leeds. To minimize the effect of the non-catalysed reaction, on-line mixing of diol 27 and KOH (shown in Figure G-3, as used in oxidation of furfural) was adopted in the continuous flow system.
Due to the low solubility of diol 27 in pure MeCN, diol 27 and KOH were separately dissolved in the mixture of MeCN and water (4:1 volume ratio). Two syringes were used for the separate feed of the diol solution (6 mM) and the KOH solution (1.5 equiv.) with the same flow rate. These two solutions were mixed through a T-junction and then mixed with oxygen in another T-junction. The residence time of the mixture in the tube before contacting the catalyst bed was about 20 min. The mixture prepared through on-line mixing was also analysed with the HPLC, and the peak area was very close to half of that of the diol solution before mixing with base (error < 5%). This means that the effect of the un-catalysed reaction on the prepared reactants before reaction could be neglected by using the on-line mixing. However, the collected sample (0.5-1 h duration for each sample) still needs to be analysed immediately.

**G1. Pt-Bi/Al$_2$O$_3$ catalyst**

Oxidation of diol 27 was initially carried out with Pt-Bi/Al$_2$O$_3$ catalyst provided by Johnson Matthey (metal loading is unknown). The reactor used in this section was fabricated with a 30 cm long PTFE tube (ID, 1.0 mm; OD, 1.6 mm; Upchurch) packed with the Pt-Bi/Al$_2$O$_3$ catalyst (63-75 μm.). Blank experiments were conducted with silica beads packed inside the reactor, and negligible conversions (< 5%) were obtained. Note that the water volume reported in the paper was not clear, and the optimized conditions for KOH concentration in Scheme 9 (1.5 equiv.) also contradicted that in the experiment section (2.5 equiv.). Using the conditions in the paper as a reference, the reaction conditions for Pt-Bi/Al$_2$O$_3$ catalyst were Pt-Bi/Al$_2$O$_3$ catalyst, 10 mg; 6 mM diol in mixture of MeCN and water (4:1 volume ratio), 10 μL/min; KOH in mixture of MeCN and water (4:1 volume ratio, 1.5
equiv.), 10 μL/min; air flow rate, 1 mL/min; 1 bar and 45 ºC. The catalyst contact time was kept the same to the optimized conditions with Pt-Bi/C in the experimental section (11.5 g_cat·h/g_diol).

![Figure G-4 HPLC profiles of a) fresh reactants and b) effluent from the reactor](image)

In the paper, the conversion of diol 27 on 5% Pt-5% Bi/Al₂O₃ catalyst was 100% under optimized conditions in batch reactor, with 84% selectivity to dialdehyde 20. In this study, the conversion on the Pt-Bi/Al₂O₃ catalyst was estimated to be ~75% based on the change of the diol 27 peak areas in HPLC. The HPLC profiles of fresh reactants and effluent from the reactor are presented in Figure G-4. As seen in the Figure G-4, very low amount of dialdehyde 20 was formed (the peak was too small to be integrated), and an unidentified species was the main product. To better present the reaction results, the selectivity to the unidentified product was calculated through

\[
S = \frac{\text{Peak area of the unidentified product}}{\text{Initial peak area of diol} \times \text{conversion}} \times 100\% \quad (G-1)
\]

The stability of Pt-Bi/Al₂O₃ catalyst in oxidation of diol 27 was shown in Figure G-5. The conversion was observed to decrease from 80% to 63% in 5 h, with a relatively stable selectivity to the unknown product. After 22h, the conversion was around 40%, indicating the deactivation of Pt-Bi/Al₂O₃ catalyst in oxidation of diol 27.
Appendix G. Oxidation of “GSK Diol”

Figure G-5 Stability of Pt-Bi/Al₂O₃ catalyst in oxidation of diol 27. Reaction conditions: 6 mM diol in mixture of MeCN and water (4:1 volume ratio), 10 μL/min; ~1.5 equiv. KOH) in mixture of MeCN and water (4:1 volume ratio), 10 μL/min; air flow rate, 1 mL/min; Pt-Bi/Al₂O₃ catalyst, 10 mg; reaction temperature, 45 ºC; reaction pressure, 1 bar. X: conversion; S: selectivity to the unknown product.

G2. Pt-Bi/C catalyst

Pt-Bi/C catalyst (prepared with crushed extrudate carbon, provided by Johnson Matthey) was also tested for the oxidation of diol 27. The reaction conditions were Pt-Bi/C catalyst (53-90 μm), 10 mg; 10 mM diol in mixture of MeCN and water (4:1 volume ratio), 5 μL/min; KOH in mixture of MeCN and water (4:1 volume ratio, 1.3 equiv.), 5 μL/min; pure oxygen, 1 mL/min; 2 bar and 45 ºC. The catalyst contact time was ~4.7 gcat·h/gdiol.

The HPLC profiles of the products are shown in Figure G-6, and diol 27 was not detected in the effluent, indicating a 100% conversion. The peak shape of dialdehyde 20 was modified through using mixture of 20 vol% acetonitrile (MeCN) and 80 vol% water with 0.02 M H₂SO₄ rather than 0.1% phosphorous acid as mobile phase. In the
second sample (1-2 h), two unknown products (named unknown 1&2) were observed together with dialdehyde 20 and monoaldehyde A. The unknown 2 was suggested to be acid based on the peak shape. The peak areas of the products are shown as a function of reaction time in Figure G-7. The target product, dialdehyde 20, was observed to decrease from 1 to 2 h. The monoaldehyde A and the unknown 2 were observed to be the main products in the last two samples (3 & 4 h).

**Figure G-6** HPLC profiles of a) fresh reactants and effluent from the reactor collected b) from 0 to 1 h and c) from 1 to 2 h.

**Figure G-7** Peak areas of different products with Pt-Bi/C catalyst as a function of reaction time. Reaction conditions: Pt-Bi/C catalyst, 10 mg; 10 mM diol in mixture of MeCN and water (4:1 volume ratio), 8.8 μL/min; KOH in mixture of MeCN and water (4:1 volume ratio, 1.3 equiv.), 8.8 μL/min; catalyst contact time, 4.7 g cat·h/g diol; pure oxygen, 1 mL/min; 2 bar and 45 °C.
Figure G-8 Peak areas of different products on Pt-Bi/C catalyst as a function of reaction time. Reaction conditions: Pt-Bi/C catalyst, 10 mg; 10 mM diol in mixture of MeCN and water (4:1 volume ratio), 8.8 μL/min; KOH in mixture of MeCN and water (4:1 volume ratio, 1.3 equiv.), 8.8 μL/min; catalyst contact time, 8.2 g\text{cat}·h/g\text{diol}; air or oxygen, 0.875 mL/min;; 1 bar and 45 ºC.

Note that 100% conversion and 85-100% selectivity to dialdehyde 20 were reported in the paper with Pt-Bi/C catalyst and air at atmospheric pressure (1 bar). Our collaborators at University of Leeds also found 88% yield to dialdehyde 20 on Pt-Bi/C catalyst (catalyst details are unknown) with air at 1 bar pressure. So, the reaction was performed with a new packing of Pt-Bi/C catalyst under 1 bar pressure with air or oxygen, The catalyst contact time was kept the same to that used at University of Leeds (~8.2 g\text{cat}·h/g\text{diol}). As shown in Figure G-8, no obvious difference on the peak area of each product was observed when changing the gas from air to oxygen. The monoaldehyde A, which was produced through partial oxidation of diol 27, was observed to be the main product from 3 h. The peak area of dialdehyde 20 was shown to decrease in the first 3 h and then stabilized for the following 4 h. However, the peak area of dialdehyde 20 only accounted for ~10% of the total peak area. The peak area of Unknown 2 was shown to increase in the first 2
h and then gradually decrease with reaction time. Note that the peak area of Unknown 2 was smaller than that under 2 bar oxygen pressure (shown in Figure G-7). If Unknown 2 was an acid, this trend could possibly be caused by the decreased oxygen species on the catalyst surface.

**Figure G-8** Liquid slug observed at the gas outlet of the Teflon AF-2400 tube-in-tube contactor.

Since low diol 27 concentration (10 mM) was used in this study (due to the low solubility in MeCN), the amount of oxygen required for the reaction was also very low, which could be potentially provided through pre-saturation of the solvent. A Teflon AF-2400 tube-in-tube membrane contactor was used to provide oxygen for the reaction. However, large amount of liquid permeation was observed in the gas phase of the contactor (slug flow observed at the gas outlet of the contactor, shown in Figure G-9), even when the pressure difference between the liquid and gas phases was ~0.1 bar. To decrease the liquid pervaporation/permeation, the gas phase was pre-bubbled in MeCN solvent, and the tube-in-tube contactor was placed in ice bath.
Unfortunately, these methods have not shown any obvious effect to decrease the amount of liquid in the gas phase. This indicates the limitations of Teflon AF-2400 membrane for the highly permeable organics, which should be considered before using membrane reactor.
Appendix H. Coating of Teflon AF-2400 Film

Although Teflon AF-2400 membrane possesses several excellent properties, its high cost (~ 3 orders of magnitude higher than that of more common fluoropolymers, such as PTFE) could be one limitation for its wider application. The usage of such expensive membrane material should be decreased through optimized reactor design. Moreover, the Teflon AF-2400 membrane was found to stick to the gasket when using the Kalrez® gasket in the large flat membrane reactor. During replacing the deactivated catalyst, the gasket needed to be peeled off from the reactor, making it likely to break the fragile Teflon AF-2400 membrane. So, the lifetime of the membrane material is still needed to be extended. One possible solution for the above issue is to coat thin Teflon AF-2400 film onto firm support, which can decrease the usage of Teflon AF-2400 membrane material (due to the thinner layer of membrane) and intensify the membrane mechanical properties.

In this study, Teflon AF-2400 film was attempted to be coated onto metal and ceramic supports. As given in the processing guide from Dupont™, Teflon AF perfluorodioxole copolymers are essentially nonpolar and contain no reactive chemical functionality. Thus, the adhesion to various substrates depends primarily on physical, rather than chemical interactions. Specific strategies for adhering Teflon AF are suggested to depend on the nature of the substrate, the degree of which the substrate can be modified, and the degree of adhesion required. Particularly, Teflon AF is reported to adhere relatively well to metals, such as titanium, aluminium, and electrolytic copper.

The Teflon AF-2400 solution used in this study was prepared through dissolving
Teflon AF-2400 solid particles into Fluorinert™ FC-40 (the solubility limit is ~0.8 g in 100 mL FC-40.). Mott sintered metal (diameter, 25.4 mm; thickness, 3 mm; Grade 0.2) and metal mesh (thickness, 0.05 mm; hole size, 76 μm; open area, 23%; stainless steel, Industrial netting) as the metal support, were ultrasonically washed with acetone and water before coating. Coating of Teflon AF-2400 film was carried out through either dipping the metal support (sintered metal or metal mesh) into the Teflon AF-2400 solution or dropping the solution onto the surface of the metal support. Due to the low solubility of Teflon AF-2400 in the FC-40 solvent (~0.8%), the coated metal support was dried at 100 ºC and coated 2 more times with the same procedure. Finally, the coated metal support was baked at 165 ºC (10 ºC above the boiling point of the solvent) for 10 min, at 245 ºC for 5 min, and at 330 ºC for 15 min. The heat treatment procedure was carried out based on the information provided by Dupont™.

![Coating of Teflon AF-2400 film on a) Mott sintered metal (grade 0.2) and b) metal mesh (pore diameter, 150 μm).](image)

*Figure H-1 Coating of Teflon AF-2400 film on a) Mott sintered metal (grade 0.2) and b) metal mesh (pore diameter, 150 μm).*

The coated support was observed with a digital microscope (Keyence VHX-600). No clear difference was observed between these two different coating methods (dipping or dropping). As seen in Figure H-1a, Teflon AF-2400 film did not adhere
to the sintered metal. For the coated metal mesh (Figure H-1b), the pores seem to be filled with Teflon AF-2400 film. But the film seems not to be uniform in the pores.

Since coating of Teflon AF-2400 film onto ceramic membrane has been reported in the literature, a ceramic membrane (length, 75 mm; OD, 10 mm; ID, 7 mm; top layer, titania with a thickness of around 14-20 μm; nominal pore size, 5 nm; Pall) was also used as substrate for the coating. The procedure was the same as reported in the literature: one end of the membrane tube was temporarily sealed and the Teflon AF-2400 solution was dropped into the tube from the other end. The solution was kept for 0.5 min before decanted, and the same procedure was repeated 3 times. At the end, the wetted tube was heated at 200 °C for 0.5 h to remove the solvent. The effect of Teflon AF-2400 film on the ceramic membrane was examined through measuring the breakthrough pressure of the coated ceramic membrane. Unfortunately, no improvement on the breakthrough pressure was observed.
Appendix I. Preliminary Assessment of PEEK Membrane

Except for the expensive Teflon AF-2400 membrane, a poly(ether ether ketone) (PEEK) flat membrane as an inexpensive polymer was also investigated as potential membrane materials in the membrane reactor.

The PEEK membrane supported on polypropylene (shown in Figure I-1) was prepared by the Livingston group at Imperial College. Before using the PEEK membrane, the PEEK film (Figure I-1c) was peeled off from the polypropylene.
support (Figure I-1b), since the polypropylene support would dissolve in the organic solvent. Polybenzimidazole (PBI) membrane supported on polypropylene was also provided, which was fragile and could not be peeled off from the support. The details on the thickness of the PEEK membrane were not given, but the thickness of the top dense PEEK layer could be 1~2 μm based on the literature. The PEEK membrane was tested within the flat membrane reactor, and the results are shown in Figure I-2 and compared with those with Teflon AF-2400 membrane in Table I-1.

![Figure I-2](image)

*Figure I-2* Solvent-free benzyl alcohol oxidation on Au-Pd/TiO₂ catalyst with PEEK flat membrane microreactor. Reaction conditions: 1 wt% Au-Pd/TiO₂ catalyst, 100 mg; oxygen pressure, 8.4 bar; oxygen flow rate, 5.0 mL/min; liquid pressure, 9.5 bar; reaction temperature, 120 ºC.

As seen in Table I-1, the conversion of benzyl alcohol and TOFs obtained with PEEK membrane were about 1/4-1/5 of those with Teflon AF-2400 membrane under the same reaction conditions, while the selectivities to benzaldehyde were very close. ~15% dibenzyl ether was also observed after ~24 h reaction when using PEEK membrane. On one hand, the oxygen permeability of Teflon AF-2400 membrane is
Appendix I. Preliminary Assessment of PEEK Membrane

~1000 barrer, which should be much higher than that of PEEK membrane (the value is not available, ~4 barrer for PTFE membrane). The comparable results obtained with these two different membranes could be possibly attributed to the different membrane thickness (shown in Table I-1), since the mass transfer coefficient of the membrane is reversely proportional to the membrane thickness. This suggests that the low permeability of membrane could be potentially offset through decreasing the membrane thickness. On the other hand, the liquid loss due to pervaporation was not considered during the reaction. If the liquid loss through the PEEK membrane was higher than that through the Teflon AF-2400 membrane, it would result to longer catalyst contact time, and thus contribute to the comparable results. So, except for the membrane permeability, another issue that shouldn’t be ignored when selecting the membrane is the pervaporation of the reactants and products (e.g. benzyl alcohol, toluene) through the membrane.

Table I-1 Results of benzyl alcohol oxidation in the flat membrane reactor with different membranes. Reaction conditions: Au-Pd/TiO$_2$ catalyst, 100 mg; oxygen pressure 8.4 bar; oxygen flow rate, 5.0 mL/min; liquid pressure, 9.5 bar for PEEK and 10 bar for Teflon AF-2400; reaction temperature, 120 °C. X, conversion; S$_B$, selectivity to benzaldehyde; S$_T$, selectivity to toluene; TOF$_{total}$, TOF$_O$ and TOF$_D$ are the turnover frequency of the total reaction, the oxidation reaction and the disproportionation reaction, respectively.

<table>
<thead>
<tr>
<th>Thickness [μm]</th>
<th>Flow rate [uL/min]</th>
<th>Teflon AF-2400</th>
<th>PEEK 1~2 (estimated)</th>
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</thead>
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<tr>
<td></td>
<td>X [%]</td>
<td>S$_B$ [%]</td>
<td>S$_T$ [%]</td>
</tr>
<tr>
<td>70</td>
<td>5</td>
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<td></td>
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<td>66</td>
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<td>Catalyst contact time [g$<em>{cat}$/s/g$</em>{alcohol}$]</td>
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</tbody>
</table>
Toluene pervaporation through PEEK membrane was experimentally measured using the same method as in Chapter 6, and the results are shown in Table I-2. The outlet of the gas channel was connected to a vial cooled at 20 °C. Without considering the organic vapour in the outlet gas from the vial, the toluene collected in the vial was up to 70.5 mg/min at 10 bar pressure difference, which corresponded to a permeation rate of 31.3 mg/(cm²·min). In contrast, it was only 0.24 mg/(cm²·min) totally for Teflon AF-2400 membrane under similar conditions (2 mL/min N₂ at STP, 120 °C, and 9 bar pressure difference). The high permeation rate of toluene through the PEEK membrane was probably caused by the high permeability of toluene through the membrane, as well as the small thickness of the membrane. The toluene pervaporation/permeation was decreased when increasing the gas pressure. This could be caused by the increased partial toluene pressure in the gas phase (due to the low gas velocity (constant gas flow at STP)) or the decreased pressure difference for toluene permeation. These results suggest that the organic pervaporation through the membrane, as well as the vapour concentration in the gas phase, should be considered when designing intrinsically safe reactors for aerobic oxidation.

Table I-2 Toluene permeation rate through PEEK membrane at 120 °C. Area of gas-liquid contact through membrane 2.25 cm² (75 mm×3mm); gas flow rate at gas inlet, 3 mL/min at STP, liquid flow rate at liquid inlet, 0.1 mL/min.

| P_L [bar] | 11 | 9.4 | 10.5 | 10.0 |
| P_G [bar] | 1  | 3.6 | 10.2 | 10.9 |
| P_L-P_G [bar] | 10 | 5.8 | 0.3  | -0.9 |
| Liquid collected in the vial (20 °C) [mg/min] | 70.5±13 | 21.7±6.8 | 2.6±0.4 | 1.0±0.1 |
| Pervaporation rate [mg/(cm²·min)] | 31.3 | 9.6 | 1.2 | 0.4 |
Another solution to decrease organic pervaporation through the membrane is to use the membrane at low temperature to decrease the vapour pressure (thus decreasing the driving force for pervaporation). As shown in Figure I-3, a batch reactor mode with recycle operation was applied for heterogeneously catalysed oxidation of benzyl alcohol. The PEEK membrane was employed in a flat membrane contactor (channel length: 75 mm; weight: 30 mm; depth: 1 mm) at room temperature, and the gas phase of the membrane contactor was pressured with air or oxygen without considering the liquid pervaporation (Ru/Al₂O₃ catalyst was used to avoid the toluene formation). The liquid reactants were pumped through the membrane contactor (to be saturated with gas at room temperature) and the 5 wt% Ru/Al₂O₃ catalyst packed reactor (to react at 100 ºC), and then recycled back to the vessel.

**Figure I-3** Schematic of the experimental set-up for aerobic oxidation of benzyl alcohol using PEEK membrane contactor and packed bed reactor (batch reactor mode with recycle operation). Gas and liquid channels in membrane contactors: length: 75 mm; weight: 30 mm; depth: 1 mm; packed PTFE tube reactor (1/8 inch PTFE tube: length, 30 cm; ID, 1.6 mm).
Appendix I: Preliminary Assessment of PEEK Membrane

Figure I-4 Oxidation of benzyl alcohol using PEEK membrane contactor and recycle operation. Reaction conditions: 10 mL neat BnOH; Ru/Al₂O₃ catalyst, 100 mg; recycle flow rate 0.1 mL/min; gas pressure in the contactor, 7.5 bar; liquid pressure, 8 bar; reaction temperature, 100 ºC.

As shown in Figure I-4, the benzyl alcohol conversion within 50 h was rather low, despite using air or oxygen in the gas phase. This is probably caused by the low solubility of oxygen in the liquid reactant at room temperature. Assuming all the dissolved oxygen in the liquid was fully converted in each cycle, it would need ~1910 h with air or 380 h with oxygen to fully convert the neat benzyl alcohol at 7.5 bar gas pressure. The average TOF of the batch reactor mode with PEEK membrane contactor (recycle flow rate, 0.1 mL/min at 7.5 bar gas pressure, catalyst contact time, 1731 g·cat·s/g·alcohol) was ~0.9 h⁻¹, which was comparable to that in continuous stirred reactor (3.0 h⁻¹, catalyst contact time, 2880 g·cat·s/g·alcohol) with 0.01 ml/min reactor system inlet flow rate and 1 mL/min recycle flow through Teflon AF-2400 membrane contractor at 5 bar oxygen pressure.
Appendix J. Test of Different Pumps for Feeding of Slurry Flow

Slurry flow reactors, in which solid catalysts are delivered together with the reactants, have been demonstrated with good mass and heat transfer performance. However, the complex interaction among catalyst particles, fluid, and reactor surface makes it challenging to handle particles in reactors without particles trapping. Nevertheless, how to efficiently feed the solid-containing fluid into reactors is still a tough issue. In this study, several types of pumps were attempted for efficient delivery of slurry flow (mixture of liquid and solid) in heterogeneously catalysed oxidation of benzyl alcohol with oxygen.

J1. Home-Made Set-up for Slurry Flow Feeding

A home-made set-up was designed for slurry flow feeding (shown in Figure J-1) and initially tested with water in vessel B. An HPLC pump (Knauer P2.1S) was used to inject water into vessel A and push the air in vessel A into vessel B. The water in vessel B was then pushed into the following reactor coil (volume, 10 ml; fabricated...
Appendix J. Test of Different Pumps for Feeding of Slurry Flow

with 1/16 inch PFA tube (1 mm ID); Vapourtec). When gas was not introduced into the liquid flow (pure liquid in the reactor coil), the flow rate measured with a flow meter (Dolomite Mitos, 0.2-5 mL/min) was very close to the set value in HPLC pump with low fluctuation (shown in Figure J-2).

![Figure J-2](image)

**Figure J-2** Water flow rate measured at the outlet of vessel B with flow meter. Flow rate from HPCL pump, 1 mL/min, pressure in the vessel, 0.10-0.13 bar.

When introducing gas (1 mL/min oxygen) into the liquid flow with a T-junction at the upstream of the reactor, the pressure drop along the reactor coil was observed to increase from 0.1 bar to 2.5 bar and then fluctuate around 1.8 bar. To make the pressure in the vessel reach steady state faster, a gas line was connected to the vessel A. When the vessel pressure was close to the reactor pressure drop, the HPLC was switched on to pumping water into the vessel A. However, the measured liquid flow rate was observed to fluctuate dramatically (0-2 mL/min) due to the fluctuating system pressure which was probably caused by the changes of hydrodynamics within the coil (non-uniform gas-liquid slugs) as well as the compressibility of gas in both the vessels and the reactor. When the pressure in the vessels was lower than the
reactor pressure drop, the measured flow rate was observed to be lower than the set value in the HPLC pump, and temporally even zero.

To avoid the effect of gas in the vessels, de Bellefon and co-workers\textsuperscript{92} pushed slurry flow directly with liquid with the help of a polypropylene bag. Due to the chemical compatibility, the polypropylene bag cannot be used in the oxidation of benzyl alcohol. A PFA gas sampling bag (500 mL) was attempted, but it is not as elastic as a polypropylene bag, making it difficult to fit with the vessel.

**J2. Vapourtec Peristaltic Pump for Slurry Flow Feeding**

A Vapourtec V-3 peristaltic pump was also tested for the feeding of slurry flow. Initially, the pump was tested with water at 8 bar back pressure and the flow rate was measured with a flow sensor (Dolomite Mitos, 0.2-5 mL/min). As shown in Figure J-3, even the 5-second average flow rate presents ±10\% relative error. This results to non-uniform lengths of liquid/gas slugs within the tube coil (shown in Figure J-4).

![Figure J-3 Measured flow rate of water from Vapourtec peristaltic pump with Dolomite flow sensor a) every second and b) 5-second average. Set flow rate 0.5 mL/min.](image_url)
Appendix J. Test of Different Pumps for Feeding of Slurry Flow

Figure J-4 Observed gas-liquid-solid slug in PFA tube with Vapourtec peristaltic pump.

Despite the non-uniform slug flow in the reactor coil, the effect of the slurry flow rate on catalyst delivery was investigated. The prepared slurry (5 g/L Au-Pd/TiO$_2$ catalyst (particle size 63-75 μm) in benzyl alcohol) was contained in a 250 mL vessel placed on a hotplate. A PTFE coated magnetic bar (length, 40 mm; ID 10 mm) was used to stir the slurry, and uniform cloudy slurry was seen at a stirring speed of ~800 rpm. Due to the low mechanical strength, the catalyst particles also seem to become powder during the stirring. The vessel was connected to the inlet of the pump with a 1/16 inch PFA tube (ID, 1 mm). The solid catalyst in the specific volume of slurry collected at the outlet of the pump was centrifuged and washed with acetone for three times, and then dried and measured to calculate the actual catalyst concentration at the outlet of the pump. The measured catalyst concentration was shown in Figure J-5. As compared to the catalyst concentration in the prepared slurry (5 g/L), the lowest flow rate was required to be ~1.0 mL/min to assure the same catalyst concentration at the outlet of pump. This suggests that catalyst with large particle sizes could not be taken out from the vessel at low liquid flow rate, which resulted to the low measured catalyst concentration. Note that the required lowest flow is highly depended on the specific properties of solids and solvents (density, viscosity, etc.).
Appendix J. Test of Different Pumps for Feeding of Slurry Flow

**Figure J-5** Effect of flow rate on the catalyst delivery. Au-Pd/TiO$_2$ catalyst particle, 63-75 μm; catalyst charge in the vessel, 5.0 g/L in benzyl alcohol; stirring speed = 800 rpm.

**Figure J-6** Schematic of experimental set-up for aerobic oxidation of benzyl alcohol using G-L-S flow and recycle operation. PUMP: Vapourtec peristaltic pump; MFC: mass flow controller; Reactor coil fabricated with 1/16 inch PFA tube (ID 1.0 mm): 10 mL, heated with Vapourtec E-series system.

Finally, the Vapourtec peristaltic pump was used for oxidation of benzyl oxidation in G-L-S flow without considering the flow type in the reactor coil. The
schematic of the set-up is shown in Figure J-6. 1 g/L Au-Pd/TiO₂ catalyst (1 wt%, Au to Pd weight ratio, 50:50; particle size < 90 μm; prepared with a traditional impregnation method) in benzyl alcohol was used as the slurry, which was stirred at a speed of 800 rpm on a hotplate (Stuart US152). The slurry was delivered with the pump and combined with oxygen (1 mL/min at STP, controlled by a mass flow controller (Brooks GF40 series)) through a T-junction before entering the reactor coil (10 mL; fabricated with PFA tube (ID 1 mm); Vapourtec). The reactor coil was heated at 120 ºC with the Vapourtec E-series system. The pressure at the outlet of the reactor coil was atmospheric pressure, and the pressure drop within the coil was ~ 0.3 bar. The outlet of the reactor coil was connected to the vessel to recycle the liquid, and the gas was vented from a 1 mm hole in the vessel cap.

As shown in Figure J-7, the conversion of benzyl alcohol was observed to be 16% after ~30 h reaction, with a 72% selectivity to benzaldehyde. A blank test was also carried out in the same reactor coil without catalyst, and 2% conversion was obtained within 30 h. This is possibly caused by the slight deposition of catalyst in the reactor coil during the previous run. The turnover frequency of the overall reaction (TOF_total) and the oxidation reaction (TOF_O) with G-L-S flow was also calculated based on the amount of catalyst contained in the reactor coil (assumed as 5 mL slurry), which was 2800 h⁻¹ and 1400 h⁻¹, respectively. The TOFs obtained in the G-L-S flow were much lower than those obtained in a batch reactor at 2 bar pressure with 2.5 wt% Au-2.5 wt% Pd/TiO₂ catalyst (26,400 h⁻¹ and 8,800 h⁻¹).²⁶ This was possibly caused by the low oxygen pressure, as well as the non-uniform slug flow that could reduce the mass transfer efficiency.
Appendix J. Test of Different Pumps for Feeding of Slurry Flow

Figure J-7 Oxidation of benzyl alcohol using G-L-S slurry flow reactor. Reaction conditions: home-made 1 wt% Au-Pd/TiO₂ catalyst (Au to Pd weight ratio, 50:50), particle size <90 µm; catalyst concentration, a) 1 g/L, b) 0 g/L; neat benzyl alcohol, 20 mL; slurry flow recycle rate, 1 mL/min; oxygen flow rate, 1 mL/min at STP; reaction temperature, 120 °C; volume of reactor coil, 10 mL.

J3. Elveflow Pressure Pump for Solid-Containing Flow Feeding

An Elveflow OB-1 pressure pump was also tested for solid-containing flow feeding. The flow rate of the pressure pump was controlled through changing the vessel pressure autonomously based on the measured flow rate via a flow sensor (ID,
Appendix J. Test of Different Pumps for Feeding of Slurry Flow

1.0 mm; Elveflow FS5) at the outlet of the pump. As shown in Figure J-8, the pressure pump presents excellent performance with fluctuation less than ± 10% per second. The calibration of the pressure pump was also carried out with pure benzyl alcohol and the curve is shown in Figure J-9.

**Figure J-8** Measured slurry flow rate from Elveflow pressure pump with Elveflow FS5 flow meter. Au-Pd/TiO$_2$ catalyst particles, 10-38 μm; catalyst concentration, 1 g/L in benzyl alcohol; set value, 234.5 μL/min; measured average value, 234.8 μL/min.

**Figure J-9** Calibration of Elveflow pressure pump with pure benzyl alcohol at atmospheric pressure.
After the calibration, the pressure pump was attempted for oxidation of benzyl alcohol in G-L-S flow. The schematic of the set-up is shown in Figure J-10. 1 wt% Au-Pd/TiO$_2$ catalyst powder (Au to Pd weight ratio, 5:95; different batch from Johnson Matthey, which shown deactivation in packed bed reactor) was dispersed in benzyl alcohol to form a slurry in the vessel (~125 mL, stirred at ~800 rpm by a magnetic bar). The slurry was pushed out from the vessel by the pressure pump, and the flow rate was measured by the flow sensor at the outlet of the vessel. The slurry was combined with oxygen through a T-junction and then directed to a heated reactor coil (10 mL, home-made with 1/16 inch PFA tube (ID 1.0 mm)). The outlet of the reactor coil was connected to a vessel (~125 mL), and the residual oxygen was vented through a back pressure regulator (Swagelok, K series).
Appendix J. Test of Different Pumps for Feeding of Slurry Flow

Figure J-11 Picture of the set-up used for observing the flow with a microscope and a high-speed camera.

Figure J-12 Flow at room temperature and atmospheric pressure without reaction: a) at the outlet of vessel containing slurry observed with microscope and b) within the tube reactor observed with high-speed camera.
Before starting the reaction, the flow was initially observed with microscope (Veho)/high-speed camera (Photron FASTCAM SA Series) at room temperature and atmospheric pressure (1 bar). The picture of the set-up is shown in Figure J-11 and the observed flow is shown in Figure J-12. The slurry flow at the outlet of the vessel (close to vertical) was seen to containing both liquid and solid. The solid particles were observed to be at the bottom of the tube and move together with the flow (shown in Figure J-12a). After introducing the gas, uniform G-L-S flow was formed (shown in the insert picture in Figure J-12b). The slurry slug was pushed forward by gas flow, with solid particles mainly circulating in the lower part of the slurry slug. Further study on the flow was not continued, since the slurry flow rate was observed to fluctuate dramatically after long time testing. The stability of the G-L-S flow was seriously affected.

![Image](image.png)

**Figure J-13** Measured slurry flow rate from Elveflow pressure pump with flow sensor after several hours testing. Au-Pd/TiO$_2$ catalyst powder; catalyst concentration, 5 g/L in benzyl alcohol; set value (red line), a) 84 μL/min and b) 133 μL/min.

The measured flow rate after long time testing with catalyst powder was shown in the Figure J-13. As compared to the initial results, the relatively error per second was much higher than ±10%. Since the working principle of the flow sensor is based on
locally heating the fluid and measuring its temperature through a glass capillary, the flow sensor is very sensitive to surface quality. Considering catalyst powder was used in the test, rather than solid particles, the serious fluctuation could possibly be caused by the catalyst fouling on the capillary wall, which affected heat transfer for both heating and temperature measurements and led to a measurement deviation and/or an offset. The flow sensor was also washed with aqua regia. The performance of the pressure pump was improved for a short period. Since catalyst powder could be formed due to the catalyst abrasion during stirring, how to provide the accurate feedback of flow rate to the pressure pump still requires better design.

**Figure J-14** Oxidation of benzyl alcohol in G-L-S flow with gas flow rate (at STP) a) 0.5 mL/min and b) 1.0 mL/min. Reaction conditions: 1 wt% Au-Pd/TiO₂ catalyst powder, 5 g/L in benzyl alcohol; slurry flow rate, 0.5 mL/min; pressure at the outlet of the reactor coil, 1 bar; pressure drop < 0.3 bar; reactor coil, 10 mL; reaction temperature, 120 °C.

Before observing the fluctuation problem, limited sets of experiments was performed with the G-L-S flow. In each set of experiments, 25 mL slurry was prepared and fed into the vessel A. When the vessel A is empty, one cycle of the
reaction was finished. The slurry collected in vessel B was analysed with GC and then fed back to the vessel A to start another cycle. The results under different gas flow rates were shown in Figure J-14. Since slug flow was formed in the reactor, the residence time of the reactants in the reactor coil could be different under different gas flow rates. Surprisingly, the conversion and selectivities were similar within four reaction cycles. This could be possibly caused by the different mass transfer efficiency under different flow rates. Moreover, blank tests without catalyst loading were also carried out before and after the experiment using catalyst powder, and the conversion per cycle was observed to be 0.2% and 1.0%, respectively. Since catalyst fouling was seen on the inner tube of the reactor coil, the differences between these two blank tests were probably caused by the catalyst fouling. Note that the conversion per cycle was only ~5% when catalyst was used. This suggests that the catalyst fouling within the reactor coil also has a remarkable effect on the reaction. Though the reactor was washed with aqua regia, catalyst fouling was observed to form quickly during the reaction.

Without considering the effect of catalyst fouling, the increasing rate of the conversion was observed to slow down after five cycles, which might be caused by the catalyst deactivation. To decrease the effect of catalyst deactivation, the results of each set of experiment were averaged within the first 4 cycles and compared in Table J-1. The TOFs were calculated based on the average residence time of slurry in the reactor coil (the ratio of the coil volume to the total flow rate of gas and slurry at STP) without considering the change of the gas flow rate, the pressure drop and the effect of catalyst fouling within the reactor coil. The TOFs obtained with different pumps were also summarized in Table J-2.
Appendix J. Test of Different Pumps for Feeding of Slurry Flow

**Table J-1** Oxidation of neat benzyl alcohol using G-L-S flow reactor. Reaction conditions: Reactor coil (10 mL; home-made with 1 mm ID PFA tube); reaction temperature, 120 °C; 1 wt% Au-Pd/TiO$_2$ catalyst powder, 5g/L; pressure at reactor outlet, atmospheric pressure (1 bar); pressure drop, ~0.3 bar. X, conversion per cycle; $S_B$, selectivity to benzaldehyde; $S_T$, selectivity to toluene; TOF$_{\text{total}}$, TOF$_O$, and TOF$_D$ are the turnover frequency of the total reaction, the oxidation reaction and the disproportionation reaction, respectively.

<table>
<thead>
<tr>
<th>Set</th>
<th>Oxygen flowrate [mL/min]</th>
<th>Slurry flow rate [mL/min]</th>
<th>X [%]</th>
<th>$S_B$ [%]</th>
<th>$S_T$ [%]</th>
<th>TOF$_{\text{total}}$ [h$^{-1}$]</th>
<th>TOF$_O$ [h$^{-1}$]</th>
<th>TOF$_D$ [h$^{-1}$]</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>0.5</td>
<td>0.5</td>
<td>4.8</td>
<td>58</td>
<td>42</td>
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<td>1071</td>
<td>5625</td>
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<tr>
<td>2</td>
<td>1.0</td>
<td>0.5</td>
<td>5.4</td>
<td>59</td>
<td>40</td>
<td>10044</td>
<td>1607</td>
<td>8437</td>
</tr>
</tbody>
</table>

**Table J-2** TOFs obtained with different pumps and reactor types. Reaction conditions: reaction temperature, 120 °C; 1 wt% Au-Pd/TiO$_2$ catalyst; pressure at reactor outlet, atmospheric pressure (1 bar); pressure drop, ~0.3 bar. X, conversion per cycle; $S_B$, selectivity to benzaldehyde; TOF$_{\text{total}}$, TOF$_O$, and TOF$_D$ are the turnover frequency of the total reaction, the oxidation reaction and the disproportionation reaction, respectively.

<table>
<thead>
<tr>
<th>Pump</th>
<th>Peristaltic pump</th>
<th>Pressure pump</th>
</tr>
</thead>
<tbody>
<tr>
<td>Au:Pd weight ratio</td>
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<td>5:95</td>
</tr>
<tr>
<td>Catalyst shape</td>
<td>&lt; 90 μm</td>
<td>powder</td>
</tr>
<tr>
<td>Catalyst concentration</td>
<td>1 g/L</td>
<td>5 g/L</td>
</tr>
<tr>
<td>10 mL reactor coil</td>
<td>Vapourtec</td>
<td>Home-made</td>
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<tr>
<td>Batch reactor mode</td>
<td>continuous recycle</td>
<td>once through</td>
</tr>
<tr>
<td>Flow type</td>
<td>non-uniform G-L-S slug flow</td>
<td>G-L-S slug flow</td>
</tr>
<tr>
<td>X [%]</td>
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<td>22</td>
</tr>
<tr>
<td>$S_B$ [%]</td>
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<td>59</td>
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<tr>
<td>TOF$_{\text{total}}$ [h$^{-1}$]</td>
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</tr>
<tr>
<td>TOF$_O$ [h$^{-1}$]</td>
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<td>1607</td>
</tr>
<tr>
<td>TOF$_D$ [h$^{-1}$]</td>
<td>1400</td>
<td>8437</td>
</tr>
</tbody>
</table>
Appendix K. Fabrication of Kovar-Glass Microreactor

Figure K-1 Failed bonding of Kovar plate with glass.

Figure K-2 Flatness of the polished Kovar plate.

Kovar material is one type of nickel-cobalt ferrous alloy with excellent thermal conductivity. It can be mechanically milled to have millimetre-scale channels and directly bonded with Pyrex glasses to be used as reactors under high temperature and
pressure. These also make it promising to be easily scaled-up. The information on Kovar-Glass microreactor was obtained from Dai’s PhD thesis. The variation in flatness of the well-polished Kovar plate (140 × 17 × 3 mm) was reported to be between 0.885 and 2.065 μm and the surface roughness, Ra (the arithmetic mean roughness), between 0.02 and 0.15 μm.

Initially, a piece of Kovar plate (dimensions 42 mm (L) × 15 mm (W) × 5 mm (T)) was polished by Kemet International Limited and attempted to be bonded with Corning 7740 glass using AML anodic bonder (Pyrex 7056 used in Dai’s thesis was not available in our group.) Unfortunately, satisfactory bonding was not realized (shown in Figure K-1). The flatness of the polished Kovar Plate was measured with DektakXT surface profiler and the results are shown in Figure K-2. The variation in flatness seems to be larger than 2 μm, suggesting that better polishing is required.