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MULTIPHASE REACTION STUDIES IN STIRRED TANK AND FIXED BED REACTORS

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By Shinbeom Lee

Entitled
MULTIPHASE REACTION STUDIES IN STIRRED TANK AND FIXED BED REACTORS

For the degree of Doctor of Philosophy

Is approved by the final examining committee:

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Shripad Revankar

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Head of the Departmental Graduate Program Date
MULTIPHASE REACTION STUDIES
IN STIRRED TANK AND FIXED BED REACTORS

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Submitted to the Faculty
of
Purdue University
by
Shinbeom Lee

In Partial Fulfillment of the
Requirements for the Degree
of
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West Lafayette, Indiana
For the glory of God
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# TABLE OF CONTENTS

<table>
<thead>
<tr>
<th>Page</th>
</tr>
</thead>
<tbody>
<tr>
<td>LIST OF TABLES</td>
</tr>
<tr>
<td>LIST OF FIGURES</td>
</tr>
<tr>
<td>NOMENCLATURE</td>
</tr>
<tr>
<td>ABSTRACT</td>
</tr>
<tr>
<td>CHAPTER 1. INTRODUCTION</td>
</tr>
<tr>
<td>1.1 Biphasic stirred tank reactor</td>
</tr>
<tr>
<td>1.1.1 Reactive mass transfer in the interface</td>
</tr>
<tr>
<td>1.1.2 Measurement techniques for interfacial area</td>
</tr>
<tr>
<td>1.1.3 Estimation methods for interfacial area</td>
</tr>
<tr>
<td>1.2 Trickle-bed reactor</td>
</tr>
<tr>
<td>1.2.1 Flow regimes and detection methods</td>
</tr>
<tr>
<td>1.2.2 Modeling of trickle bed reactors</td>
</tr>
<tr>
<td>1.3 Research objectives</td>
</tr>
<tr>
<td>CHAPTER 2. KINETIC STUDY OF BIPHASIC ALDOL CONDENSATION OF N-BUTYRALDEHYDE USING STIRRED CELL</td>
</tr>
<tr>
<td>2.1 Introduction</td>
</tr>
<tr>
<td>2.2 Theory</td>
</tr>
<tr>
<td>2.2.1 Penetration model and plateau region</td>
</tr>
<tr>
<td>2.2.2 Solubility of nBAL</td>
</tr>
<tr>
<td>2.2.3 Diffusivity and mass transfer coefficient</td>
</tr>
<tr>
<td>2.3 Experimental</td>
</tr>
<tr>
<td>2.3.1 Experimental apparatus</td>
</tr>
<tr>
<td>2.3.2 Analysis of the organic phase composition</td>
</tr>
<tr>
<td>2.3.3 Experimental Procedure</td>
</tr>
</tbody>
</table>
2.4 Results and discussions

<table>
<thead>
<tr>
<th>Section</th>
<th>Page</th>
</tr>
</thead>
<tbody>
<tr>
<td>2.4.1 Determination of the range of experimental conditions</td>
<td>26</td>
</tr>
<tr>
<td>2.4.2 Confirmation of plateau region</td>
<td>29</td>
</tr>
<tr>
<td>2.4.3 Kinetic study</td>
<td>31</td>
</tr>
<tr>
<td>2.4.4 Sensitivity study of activation energy and exponent of $C_{NaOH}$</td>
<td>36</td>
</tr>
<tr>
<td>2.4.5 Comparison with previously published values</td>
<td>39</td>
</tr>
<tr>
<td>2.4.6 Kinetic study in the reaction controlled regime</td>
<td>40</td>
</tr>
<tr>
<td>2.4.7 Analysis of nBAL concentration in aqueous phase film and organic phase films</td>
<td>42</td>
</tr>
</tbody>
</table>

2.5 Concluding Remarks

CHAPTER 3. ALDOL CONDENSATION OF N-BUTYRALDEHYDE IN A BIPHASIC STIRRED TANK REACTOR: EXPERIMENTS AND MODELS

<table>
<thead>
<tr>
<th>Section</th>
<th>Page</th>
</tr>
</thead>
<tbody>
<tr>
<td>3.1 Introduction</td>
<td>47</td>
</tr>
<tr>
<td>3.2 Theory</td>
<td>51</td>
</tr>
<tr>
<td>3.2.1 Reactor modeling using the film model</td>
<td>51</td>
</tr>
<tr>
<td>3.2.2 Interfacial area estimation</td>
<td>55</td>
</tr>
<tr>
<td>3.3 Interfacial area</td>
<td>57</td>
</tr>
<tr>
<td>3.3.1 Experimental apparatus</td>
<td>57</td>
</tr>
<tr>
<td>3.3.2 Calibration</td>
<td>58</td>
</tr>
<tr>
<td>3.3.3 Experimental procedure</td>
<td>60</td>
</tr>
<tr>
<td>3.3.4 Measurements of the Sauter mean diameters</td>
<td>62</td>
</tr>
<tr>
<td>3.3.5 Estimation of the Sauter mean diameter using correlations</td>
<td>65</td>
</tr>
<tr>
<td>3.4 Biphasic stirred tank reactor for nBAL aldol condensation</td>
<td>70</td>
</tr>
<tr>
<td>3.4.1 Experimental setup</td>
<td>70</td>
</tr>
<tr>
<td>3.4.2 Preliminary studies</td>
<td>73</td>
</tr>
<tr>
<td>3.4.3 Experimental results and evaluation of the reactor model</td>
<td>74</td>
</tr>
<tr>
<td>3.5 Concluding remarks</td>
<td>78</td>
</tr>
</tbody>
</table>

CHAPTER 4. ACETOPHENONE HYDROGENATION ON RH/AL$_2$O$_3$ CATALYST: INTRINSIC KINETICS AND EFFECTS OF INTERNAL DIFFUSION

<table>
<thead>
<tr>
<th>Section</th>
<th>Page</th>
</tr>
</thead>
<tbody>
<tr>
<td>4.1 Introduction</td>
<td>81</td>
</tr>
</tbody>
</table>
4.2 Experimental ............................................................................................................82
  4.2.1 Catalysts ......................................................................................................... 82
  4.2.2 Experimental setup ........................................................................................ 83
  4.2.3 Experimental procedure ................................................................................. 84
4.3 Results and discussion..............................................................................................85
  4.3.1 Reaction route validation ............................................................................... 85
  4.3.2 Catalyst screening .......................................................................................... 88
  4.3.3 Intrinsic reaction kinetic study ....................................................................... 91
    4.3.3.1 Reaction controlled regime ....................................................................... 94
    4.3.3.2 Experiments and initial rates .................................................................... 96
    4.3.3.3 Kinetic modeling ...................................................................................... 99
    4.3.3.4 Parameter fitting ..................................................................................... 101
    4.3.3.5 Model evaluation and improvements ..................................................... 102
    4.3.3.6 Discussion ............................................................................................... 105
  4.3.4 Effects of internal diffusion ......................................................................... 107
    4.3.4.1 Catalyst characterization ......................................................................... 107
    4.3.4.2 Experimental results ............................................................................... 108
    4.3.4.3 Modeling of internal diffusion effects .................................................... 110
    4.3.4.4 Model evaluation and discussion ............................................................ 112
4.4 Conclusions ............................................................................................................116

CHAPTER 5. ACETOPHENONE HYDROGENATION ON RH/AL₂O₃ CATALYST:
FLOW REGIME EFFECT AND TRICKLE BED REACTOR MODELING............. 118
  5.1 Introduction ............................................................................................................118
  5.2 Experimental Setup ................................................................................................120
  5.3 Hydrodynamic tests .............................................................................................122
    5.3.1 Experimental procedure ............................................................................... 123
    5.3.2 Experimental results and discussion ............................................................ 124
  5.4 Reactions in low and high interaction regimes ......................................................128
    5.4.1 Experimental procedure ............................................................................... 129
    5.4.2 Experiments in high and low interaction regimes ........................................ 130
<table>
<thead>
<tr>
<th>Section</th>
<th>Page</th>
</tr>
</thead>
<tbody>
<tr>
<td>5.4.3 Experiments for the external mass transfer effect</td>
<td>131</td>
</tr>
<tr>
<td>5.5 Modeling of the trickle bed reactor</td>
<td>138</td>
</tr>
<tr>
<td>5.6 Concluding remarks</td>
<td>145</td>
</tr>
<tr>
<td><strong>CHAPTER 6. CONCLUSIONS AND RECOMMENDATIONS FOR FUTURE WORK</strong></td>
<td>147</td>
</tr>
<tr>
<td>6.1 Summary</td>
<td>147</td>
</tr>
<tr>
<td>6.1.1 Intrinsic reaction kinetic study in the mass transfer regime</td>
<td>147</td>
</tr>
<tr>
<td>6.1.2 Biphasic stirred tank reactor modeling with interfacial area study</td>
<td>148</td>
</tr>
<tr>
<td>6.1.3 Intrinsic reaction kinetics and pore diffusion effects in a slurry reactor</td>
<td>148</td>
</tr>
<tr>
<td>6.1.4 Flow regime effects on trickle bed reactor performance and its modeling</td>
<td>149</td>
</tr>
<tr>
<td>6.2 Recommendations for future work</td>
<td>150</td>
</tr>
<tr>
<td>6.2.1 Drop size distributions</td>
<td>150</td>
</tr>
<tr>
<td>6.2.2 New applications</td>
<td>151</td>
</tr>
<tr>
<td>6.2.3 Mass transfer study with small particles and hydrogen</td>
<td>151</td>
</tr>
<tr>
<td><strong>APPENDICES</strong></td>
<td></td>
</tr>
<tr>
<td>APPENDIX A. Experiments to verify 2EHELI effect on the solubility of nBAL in water</td>
<td>153</td>
</tr>
<tr>
<td>APPENDIX B. Estimation of diffusivity, $D_{AB}$ and mass transfer coefficient, $k_L$</td>
<td>154</td>
</tr>
<tr>
<td>APPENDIX C. 3D-CFD study for power number and interface stability</td>
<td>157</td>
</tr>
<tr>
<td>APPENDIX D. Mass transfer coefficient for the dispersed phase</td>
<td>160</td>
</tr>
<tr>
<td>APPENDIX E. Estimation of the interfacial tension</td>
<td>162</td>
</tr>
<tr>
<td>APPENDIX F. 3D-CFD study for P/V and the disturbance by the borescope</td>
<td>164</td>
</tr>
<tr>
<td>APPENDIX G. nBAL solubility in the reaction temperature and the atmospheric temperature</td>
<td>167</td>
</tr>
<tr>
<td>APPENDIX H. Comparison among statistical correlations from DOEs and the analytical model used in the stirred tank reactor study</td>
<td>170</td>
</tr>
<tr>
<td>APPENDIX I. Simulation and analysis of the overall reaction rate for the stirred tank</td>
<td>173</td>
</tr>
<tr>
<td>APPENDIX J. an Improved interfacial area estimation method: IAC with CFD</td>
<td>176</td>
</tr>
</tbody>
</table>
LIST OF TABLES

<table>
<thead>
<tr>
<th>Table</th>
<th>Page</th>
</tr>
</thead>
<tbody>
<tr>
<td>Table 2.1: Experimental results for intrinsic reaction kinetics using stirred cell; agitator speed 50 rpm</td>
<td>32</td>
</tr>
<tr>
<td>Table 2.2: Activation energies calculated using different models</td>
<td>34</td>
</tr>
<tr>
<td>Table 2.3: Effect of empirical parameter for nBAL ($i_{nBAL}$) on the exponent of $C_{NaOH}$</td>
<td>35</td>
</tr>
<tr>
<td>Table 2.4: Calculations of nBAL concentration in the film of aqueous phase and the comparison of reaction rate and mass transfer rate for 2EHEL in the organic phase</td>
<td>43</td>
</tr>
<tr>
<td>Table 3.1: Temperature dependent viscosities and interfacial tension for the dispersed and continuous phases at 1.9M $C_{NaOH}$</td>
<td>64</td>
</tr>
<tr>
<td>Table 3.2: Viscosities and interfacial tension for dispersed and the continuous phases at 70 °C</td>
<td>65</td>
</tr>
<tr>
<td>Table 3.3: Summary of the correlations used in Figure 7</td>
<td>67</td>
</tr>
<tr>
<td>Table 3.4: $k_{overall}$ from experiments, the models using $d_{32}$ estimations and $d_{32}$ measurements</td>
<td>78</td>
</tr>
<tr>
<td>Table 4.1: Properties of 1% Rh/Al₂O₃ catalyst</td>
<td>93</td>
</tr>
<tr>
<td>Table 4.2: Kinetic modeling results based on L-H kinetics for AP hydrogenation</td>
<td>93</td>
</tr>
</tbody>
</table>
Table 5.1: Initial acetophenone hydrogenation rates (mmol/gcat.s) with different liquid superficial velocities ($v_{liq}$) and reaction time (0-30 min.) at 80 °C, 25 barg, 0.12 M $C_{AP,o}$ and 3 stdl/min $H_2$ flow rate. (T: trickle flow; B: bubbly flow) ...................................................... 133

Table 5.2: Criteria values to verify external mass transfer limitations with 3 stdl/min $H_2$ and 6 ml/min liquid at 80 °C, 25 barg and 0.12 M $C_{AP,o}$ ................................................................. 134

Table 5.3 Initial acetophenone hydrogenation rates (mmol/gcat.s) with different liquid superficial velocities ($v_{liq}$) and reaction time (10-40 min.) at 0.12 M $C_{AP,o}$. (T: trickle flow; B: bubbly flow) ................................................................................................................................. 138
LIST OF FIGURES

<table>
<thead>
<tr>
<th>Figure</th>
<th>Page</th>
</tr>
</thead>
<tbody>
<tr>
<td>Figure 1.1: Types of multiphase reactors and phase contacting patterns [2]</td>
<td>2</td>
</tr>
<tr>
<td>Figure 2.1: Schematic diagram of the biphasic interface</td>
<td>16</td>
</tr>
<tr>
<td>Figure 2.2: Effect of 2EHEL on nBAL solubility; Initial: 14ml organic phase of nBAL/2EHEL, Final: mixed with 36ml water; Est.1: estimated by UNIQUAC, Est.2: estimated by nBAL solubility in water, Exp.: experiment data, See Appendix A for experiment details</td>
<td>19</td>
</tr>
<tr>
<td>Figure 2.3: Equilibrium of nBAL in water and organic phases, calculated using the UNIQUAC activity coefficient model</td>
<td>20</td>
</tr>
<tr>
<td>Figure 2.4: Schematic diagram of the stirred cell system; 1- stirred cell reactor, 2 - nBAL injector, 3 – ice water bath; PI – pressure indicator, TI – temperature indicator, TIC – temperature indicator / controller</td>
<td>23</td>
</tr>
<tr>
<td>Figure 2.5: Plot of ln $C_{nBAL}$ versus time showing 1st order kinetics</td>
<td>28</td>
</tr>
<tr>
<td>Figure 2.6: Plateau region tests at ■ 120 °C, 0.72 M; ▲ 110 °C, 0.38 M; ○ 110 °C, 1.90 M; □ 150 °C, 0.38 M; Δ 150 °C, 1.90M</td>
<td>30</td>
</tr>
<tr>
<td>Figure 2.7: Arrhenius plots for four different NaOH concentrations</td>
<td>33</td>
</tr>
<tr>
<td>Figure 2.8: Plot of ln $k$ versus ln $C_{NaOH}$ for five different temperatures (---: linear regression)</td>
<td>35</td>
</tr>
</tbody>
</table>
Figure 2.9: Comparison of liquid chromatographic profiles at 30 °C and 120 °C, Peaks: 1, 2 = major impurities; 3 = nBAL; 4 = 2EHEL; 5 = hexanal; 6 = 2, 4-dinitrophenylhydrazine (used in derivatization method) .......................................................... 41

Figure 2.10: nBAL concentration distribution in the film of the aqueous phase at various conditions ........................................................................................................................................... 43

Figure 3.1: Schematic diagram of the biphasic interface .............................................................................................................................. 52

Figure 3.2: Schematic diagram of the biphasic stirred tank with the borescope system; 1-stirred tank, 2 – CCD camera, 3 – Personal computer, 4 – fiber optic strobe. ............... 58

Figure 3.3: The calibration of the borescope system with glass beads (1: 250 µm glass beads, 2: 20-30 mesh glass beads). .......................................................................................................................... 59

Figure 3.4: Drop pictures by the borescope system; a) 110°C, 1.9M C<sub>NaOH</sub>, 800 rpm; b) 25°C, 0M C<sub>NaOH</sub>, 600 rpm. .............................................................................................................................. 60

Figure 3.5: Accumulated droplet size distributions at 110°C, 1.9M C<sub>NaOH</sub>; T-top, B-bottom; the numbers in the legend denote agitator rpm. ................................................. 61

Figure 3.6: d<sub>32</sub> measurements of 2EHEL (30%) / Water (70%) system at 25-110 °C, 0-1.9M C<sub>NaOH</sub>, 600-1000 rpm; (a) 0M C<sub>NaOH</sub>, 25-110°C, (b) 1.9M C<sub>NaOH</sub>, 25-110°C, (c) 800 rpm, 0-1.9M C<sub>NaOH</sub> ........................................................................................................................................... 63

Figure 3.7: Parity plot for the Sauter mean diameters estimated by published correlations; the dash lines represent ±15% errors. ............................................................................................................... 66

Figure 3.8: Comparison between the measurements (points) and the estimation (line) by a) Santiago and Trambouze (1971); b) adjusted Godfrey et al. (1989); c) this work. ........ 69
Figure 3.9: Schematic diagram of the biphasic stirred tank for nBAL aldol condensation. (1: stirred tank reactor, 2: injector, 3: condenser, PI: pressure indicator, TIC: temperature indicator / controller) ............................................................................................................................. 71

Figure 3.10: Confirmation of 1st order reaction through ln $C_{nBAL}$ as a function of time. 1: 110 °C, 1.03 M $C_{NaOH}$, 600 rpm; 2: 110 °C, 1.03 M $C_{NaOH}$, 800 rpm; 3: 120 °C, 1.52 M $C_{NaOH}$, 700 rpm; 4: 120 °C, 0.76 M $C_{NaOH}$, 900 rpm; 5: 92 °C, 1.60 M $C_{NaOH}$, 919 rpm. 73

Figure 3.11: Parity plots of the overall rate constant - the experiments versus the estimations by the model using (a) the correlation developed in this work; (b) the correlations from literature; the dash lines represent ±15% errors. ........................................... 75

Figure 3.12: Comparison of the estimation accuracy between the model using $d_{32}$ estimations and $d_{32}$ measurements; the dash lines represent ±8% errors. ....................... 77

Figure 4.1: Schematic diagram of the experimental setup. (1 - 300-ml stirred tank reactor, 2 - injector, PI - pressure indicator, TIC - temperature indicator / controller) ............... 83

Figure 4.2: The reaction scheme of acetophenone hydrogenation on Rh/Al$_2$O$_3$ catalyst: hydrogenation (solid arrow), hydrogenolysis (dashed arrow). ........................................ 86

Figure 4.3: Concentration-time profiles for hydrogenation of (a) acetophenone and (b) phenyl ethanol with 1% Rh/Al$_2$O$_3$ catalyst: (a) 0.6 M $C_{AP,o}$, 2 g$_{cat}$, 2.6 MPa PH2, 80 °C at start, 150 °C after 40 min.; (b) 0.12 M $C_{PE,o}$, 0.3 g$_{cat}$, 2.6 MPa PH2, 80 °C. 87

Figure 4.4: Concentration-time profiles for hydrogenation of acetophenone with (a) 5% Ru/Al$_2$O$_3$ and (b) 1% Pt/Al$_2$O$_3$ at 0.12 M $C_{AP,o}$, 0.3 g$_{cat}$, 2.6 MPa PH2 and 80 °C. ......... 89

Figure 4.5: Concentration-time profiles for hydrogenation of acetophenone with (a) 5% Rh/C and (b) 1% Rh/Al$_2$O$_3$ at 0.12 M $C_{AP,o}$, 0.3 g$_{cat}$, 2.6 MPa PH2 and 80 °C. ............... 90
Figure 4.6: Initial rate of hydrogen consumption and acetophenone hydrogenation at 4.1 MPa PH2, 100 °C, 0.4 M \( \text{CAP}_o \) and 2 g\(_{\text{cat}}\). ................................................................. 95

Figure 4.7: Initial hydrogenation rate versus (a) hydrogen partial pressure at 0.4 M \( \text{CAP}_o \) and (b) initial acetophenone concentration at 2.6 MPa PH2 and different temperatures. 98

Figure 4.8: Comparison of experimental kinetic data with estimation results using model V: (a) 60 °C, 4.1 MPa PH2, 0.04 M \( \text{CAP}_o \) and 0.2 g\(_{\text{cat}}\) \( (R^2 = 0.983) \); (b) 80 °C, 2.6 MPa PH2, 0.08 M \( \text{CAP}_o \) and 0.2 g\(_{\text{cat}}\) with \( (R^2 = 0.959) \); (c) 100 °C, 1.1 MPa PH2, 0.4 M \( \text{CAP}_o \) and 0.8 g\(_{\text{cat}}\) with \( (R^2 = 0.934) \). ................................................................. 104

Figure 4.9: SEM images of catalyst particle edge and the corresponding EDX images collected in map-scan of 1% Rh/Al\(_2\)O\(_3\) catalysts: (a) and (c) for 0.5-mm, (b) and (d) for 2.5-mm particle sizes. ..................................................................................... 108

Figure 4.10: Typical experimental (symbols) and modeling results (solid lines) for acetophenone hydrogenation on 1% Rh/Al\(_2\)O\(_3\): (a) 0.5-mm, (b) 2.5-mm particles at 80 °C, 2.6 MPa PH2, 0.12 M \( \text{CAP}_o \) and 0.3 g\(_{\text{cat}}\). ............................................................................. 109

Figure 4.11: Concentration profiles inside catalyst particles at 80 °C, 2.6 MPa PH2 and 0.12 M \( \text{CAP}_o \) using (a) 0.5-mm diameter particles, \( \eta = 0.467 \), (b) 2.5-mm particles, \( \eta = 0.144 \). ............................................................. 113

Figure 4.12: Sensitivity study for concentration profiles versus reaction time with different tortuosity and porosity at 80 °C, 2.6 MPa PH2 and 0.12 M \( \text{CAP}_o \) using 2.5 mm diameter catalysts with (a) \( \tau = 4, \varepsilon_p = 0.69 \) and (b) \( \tau = 2.7, \varepsilon_p = 0.65 \), solid lines. The dashed lines are the corresponding results with \( \tau = 4, \varepsilon_p = 0.65 \). ......................... 115
Figure 5.1: The reaction scheme of acetophenone hydrogenation on Rh/Al\textsubscript{2}O\textsubscript{3} catalyst: hydrogenation (solid arrow), hydrogenolysis (dashed arrow). ............................................... 119

Figure 5.2: Schematic diagram of the experimental setup. 1: fixed bed reactor, 2: reservoir, 3: back-pressure regulator, 4: mass flow controller, 5: piston pump, 6 and 7: pre-heaters, 8: quencher and condenser, PDI: pressure difference indicator, TI: temperature indicator. ................................................................. 121

Figure 5.3: Confirmation of indirect regime detection method with visual observation using FEP tube with 0.5 mm diameter \(\gamma\)-alumina spheres; 0.033 m/s \(\text{H}_2\) superficial velocity, 0.6 M acetophenone in cyclohexane (liquid phase), 20 bar and 25 \(\text{°C}\). (Stdev: standard deviation) ........................................................................................................ 124

Figure 5.4: Regime boundary map including effects of five different variables at 25 \(\text{°C}\). Legend – gases (N: Nitrogen, H: hydrogen)/liquids (W: water, O: acetophenone in cyclohexane), pressure (bar, PH2), tube material (PC: polycarbonate, FEP: fluorinated ethylene propylene, SS: stainless steel), particle material (G: glass beads, A: \(\gamma\)-alumina). ........................................................................................................ 125

Figure 5.5: Determination of liquid flow rates in low (Trickle) and high (Bubbly) interaction regimes for reaction operating conditions with 0.057 m/s hydrogen superficial velocity at 25 bar, 80 \(\text{°C}\) and 0.6 M \(\text{C}_{\text{AP}}\). (Stdev: standard deviation) ......................... 128

Figure 5.6: Concentration profiles with time for acetophenone hydrogenation in low (dash) and high (solid) interaction regimes: a) 0.6 M \(\text{C}_{\text{AP.o}}\) with 3 gcat, b) 0.04 M \(\text{C}_{\text{AP.o}}\) with 0.3 gcat; 25 bar, 80 \(\text{°C}\) ................................................................. 130
Figure 5.7: Acetophenone concentration profiles with 5 different liquid superficial velocities at 3 stdl/min H₂ flow rate; 80 °C, 25 barg and 0.12 M Cₐₚ,o with 0.3 g 1% Rh/Al₂O₃ catalyst; in parenthesis – liquid volume flow rate, ml/min.............................. 132

Figure 5.8: Acetophenone concentration profiles with 5 different liquid superficial velocities: a) 1.5 stdl/min, b) 4.5 stdl/min H₂ flow rate; 80 °C, 25 barg, 0.12 M Cₐₚ,o and 0.3 g 1% Rh/Al₂O₃ catalyst; in parenthesis – liquid volumetric flow rate, ml/min....... 136

Figure 5.9: Comparison between experiments (symbols, t: trickle flow) and model predictions in trickle (dotted curve) and bubbly flow (solid curve) regimes with 0.12 M Cₐₚ,o: a) 80 °C, 25 barg, 3 stdl/min H₂ and 0.3 gcat; b) 80 °C, 25 barg, 1.5 stdl/min H₂ and 0.3 gcat; c) 100 °C, 25 barg, 3 stdl/min H₂ and 0.24 gcat; d) 80 °C, 10 barg, 1.2 stdl/min H₂ and 0.3 gcat................................................................. 144
NOMENCLATURE

\( A \), Tube cross-sectional area \([m^2]\)

\( A_{surf} \), Interfacial area of stirred cell \([m^2]\)

\( a \), Effective interfacial area \([m^2/m^3]\)

\( a_{kGL,H_2} \), External mass transfer coefficient of hydrogen \([/s]\)

\( a_{GL,H_2} \), Gas-liquid mass transfer coefficient of hydrogen \([/s]\)

\( a_s \), Liquid-solid interfacial area, solid surface area \([m^2/m^3]\)

\( C, C_i, C_{i,j} \), Molar concentration of component \( i \) \([kmol/m^3]\) or Adjustable parameter for different flow regimes

\( C_D \), Drag coefficient (1 for cube, 0.47 for sphere)

\( C^0_i, C^0_{i,j} \), Initial molar concentration of component \( i \) at \( t = 0 \) \([kmol/m^3]\)

\( C_{TI,RC \ or \ mu} \), Adjustable constants for turbulent impact, random collision or viscosity correction term

\( C_x^* \), Concentration of component \( x \) in the liquid in equilibrium with the gas \([mol/L]\)

\( D, D_{AB} \), Diffusivity of A into B \([m^2/s]\)

\( D e_x \), Effective diffusivity of component \( x \) in solvents inside particles \([m^2/s]\)

\( D_e = \frac{D e_p \sigma}{\tau} \)

\( d, d_i \), Drop diameter or diameter \([m]\)

\( d_{10} \), Mean diameter \([m]\) \( \frac{\sum_i n_id_i}{\sum_i n_i} \)

\( d_{32} \), Sauter mean diameter \([m]\) \( \frac{\sum_i n_id_i^3}{\sum_i n_id_i^2} \)

\( d_{imp} \), Impeller diameter \([m]\)
\( d_p \)  
Particle diameter [m]

\( E_a \)  
Activation energy [kcal/mol]

\( F_{c}, F_{c,r} \)  
Conversion factor relating catalyst amount to active Rh site [mmol/gcat]

\[
F_c = \text{Rh dispersion} \times f \times \text{wt} \% \text{Rh} |c| \times 1000 / M_{wRh}
\]

\[
F_{c,r} = \text{Rh dispersion} \times f \times \text{wt} \% \text{Rh} |c(r)| \times 1000 / M_{wRh}
\]

(Note: % value should be applied as fraction, fraction = % value/100.)

\( f \)  
Turnover frequency factor based on metal crystallite size [m] or Factor for partial wetting and high interaction regime effects

\( F_c \)  
Conversion factor relating active Rh site to catalyst amount [mmol/gcat]

: 0.0424 for 0.5 mm diameter 1% Rh/Al_2O_3

\( g \)  
The acceleration of gravity [m/s^2]

\( g_g \)  
Gas mass superficial velocity [kg/m^2.s]

\( g_{cat} \)  
Catalyst amount [g]

\( H \)  
Henry constant for hydrogen in cyclohexane [MPa.m^3/kmol]

\( l \)  
Ionic strength of the solution [kmol/m^3]

\( i_i \)  
Empirical parameter of component \( i \) [m^3/kmol]

\( J_i \)  
Molar flux of component \( i \) on the interface [kmol/m^2.s]

\( K_H \text{or} H_2 \)  
Adsorption equilibrium constant of hydrogen [MPa^-1]

\( K_x \)  
Adsorption equilibrium constant of component \( x \) [m^3/kmol]

\( k, k_i \)  
Reaction rate constant of reaction \( i \) [/s]

\( k' \)  
Pre-exponential factor [/s]

\( k'' \)  
Reaction rate constant except \( C_{NaOH} \) [m^3/kmol.s]

\( k_B \text{or} C \)  
Efficiencies of breakage and coalescence

\( k_L \)  
Mass transfer coefficient of aqueous phase [m/s]
<table>
<thead>
<tr>
<th>Symbol</th>
<th>Definition</th>
</tr>
</thead>
<tbody>
<tr>
<td>$k_{L,i}$</td>
<td>Mass transfer coefficient of phase $i$ [m/s] or Liquid-solid mass</td>
</tr>
<tr>
<td></td>
<td>transfer coefficient of component $x$ [m/s]</td>
</tr>
<tr>
<td>$k_{\text{number}}$</td>
<td>Reaction rate constant</td>
</tr>
<tr>
<td>$k_{\text{overall}}$</td>
<td>Overall reaction rate constant $[s^{-1}]$</td>
</tr>
<tr>
<td>$k'_{\text{overall}}$</td>
<td>Overall reaction rate constant $[m^3/s]$ $k'<em>{\text{overall}} = V</em>{\text{org}}k_{\text{overall}}$</td>
</tr>
<tr>
<td>$k_s$</td>
<td>Salting out parameter $[m^3/kmol]$ or Liquid-to-catalyst mass transfer</td>
</tr>
<tr>
<td></td>
<td>coefficient $[m/s]$</td>
</tr>
<tr>
<td>$m$</td>
<td>Distribution coefficient of nBAL between organic phase and NaOH solution</td>
</tr>
<tr>
<td>$m_w$</td>
<td>Distribution coefficient of nBAL between organic phase and water</td>
</tr>
<tr>
<td>$n$</td>
<td>Agitation speed $[/s]$</td>
</tr>
<tr>
<td>$P$</td>
<td>Agitation power $[J/s]$</td>
</tr>
<tr>
<td>$P_x$</td>
<td>Partial pressure of component $x$ $[MPa]$</td>
</tr>
<tr>
<td>$P_c$</td>
<td>Critical pressure $[\text{bar}]$</td>
</tr>
<tr>
<td>$R$</td>
<td>Gas constant, $8.31 \ [J/\text{mol.K}]$</td>
</tr>
<tr>
<td>$R_i$</td>
<td>Reaction rate of component $i$ in film reaction regime $(Ha &gt; 3)$ $[\text{kmol/m}^2.\text{s}]$ or Resistance of phase $i$ $[\text{kmol/s}]$</td>
</tr>
<tr>
<td>$R_{\text{film},i}$</td>
<td>Mass transfer rate of component $i$ by penetration theory dominated</td>
</tr>
<tr>
<td></td>
<td>by reaction in the film of aqueous phase at $x = 0$ $[\text{kmol/s}]$</td>
</tr>
<tr>
<td>$R_{\text{org},i}$</td>
<td>Equivalent maximum mass transfer rate of component $i$ by film</td>
</tr>
<tr>
<td></td>
<td>theory into organic phase $[\text{kmol/s}]$</td>
</tr>
<tr>
<td>$R_{\text{overall}}$</td>
<td>Overall reaction rate $[\text{kmol/s}]$</td>
</tr>
<tr>
<td>$r$</td>
<td>Radius $[\text{m}]$</td>
</tr>
<tr>
<td>$r_i$</td>
<td>Intrinsic reaction rate of component $i$ $[\text{kmol/m}^3.\text{s}]$ or</td>
</tr>
<tr>
<td></td>
<td>Rate of reaction $i$ based on turnover frequency $[s^{-1}]$</td>
</tr>
<tr>
<td>$r_x$</td>
<td>Observed reaction rate of component $x$ $[\text{mol/L.s}]$</td>
</tr>
<tr>
<td>$T$</td>
<td>Temperature $[\text{K}]$</td>
</tr>
<tr>
<td>$T_b$</td>
<td>Boiling point $[\text{K}]$</td>
</tr>
<tr>
<td>$T_c$</td>
<td>Critical temperature $[\text{K}]$</td>
</tr>
</tbody>
</table>
### Symbols

**Latin symbols**

- \( t \): Time [s]
- \( u_t \): Terminal velocity [m/s]
- \( V_{bi} \): Molar volume of component \( i \) at normal boiling temperature [cc/mol]
- \( V_i \): Volume of \( i \) phase [m³]
- \( V_i^0 \): Molar volume of component \( i \) [m³/kmol]
- \( \bar{V}_i^0 \): Partial molar volume of nonelectrolyte at infinite dilution [liter/mol]
- \( \bar{V}_s^0 \): Partial molar volume of electrolyte at infinite dilution [liter/mol]
- \( V_r \): Catalyst bed volume in the tubular reactor [ml]
- \( V_t \): Liquid volume in the liquid reservoir or total liquid volume [m³]
- \( w_i \): Mass fraction of component \( i \)
- \( X \): Conversion of nBAL
- \( x \): Distance from interface in aqueous phase [m]
- \( Z_i \): Charge of ion
- \( z \): Axial length of the tubular reactor [m]

**Greek symbols**

- \( \alpha \): Volume fraction of dispersed phase
- \( \delta \): Film thickness \( \delta = D/k_L \) [m]
- \( \varepsilon \): Organic phase volume fraction or turbulent dissipation rate [m²/s³]
- \( \varepsilon_b \): Bed voidage
- \( \varepsilon_i \): Volume fraction of phase \( i \)
- \( \varepsilon_p \): Particle porosity
- \( \eta \): Kolmogoroff’s length scale with averaged turbulent dissipation rate
  \( \eta = (v_c^3/\varepsilon_{avg.})^{0.25}, \varepsilon_{avg.} = P/(\rho v_c) \) [m]
- \( \eta_i \): Internal effectiveness factor of reaction \( i \)
- \( \mu_i \): Viscosity of component \( i \) or phase \( i \) [mN.s/m²]
Density of component $i$ or phase $i$ [kg/m$^3$]

Catalyst particle density [g/cm$^3$]

Catalyst density in the liquid phase of reactor [g/cm$^3$]

Surface tension [N/m] or Constriction factor

Surface tension of component $i$ [mN/m]

Interfacial tension between components $i, j$ or $i, j$ phases [mN/m]

Tortuosity

Association parameter

Interaction parameter

Viscosity of component $i$ [mN.s/m$^2$]

Liquid flow rate [m$^3$/s]

Kinematic viscosity for continuous phase [m$^2$/s]

Dimensionless groups

Hatta number $\frac{\sqrt{D_k}}{k_L}$

Power number $\frac{P}{\rho n^3 d_{imp}^5}$

Reynolds number $\frac{d_{32} u_{r} \rho}{\mu}$

Impeller reynolds number $\frac{d_{imp}^2 n \rho}{\mu}$

Sherwood number $\frac{k_{sd} d_p}{D_x}$

Weber number $\frac{\rho c n^2 d_{imp}^3}{\sigma}$

Subscripts and superscripts

* Value on interface ($x = 0$)
AP     Acetophenone
aq     Aqueous phase
bulk   Bulk region of aqueous phase
c      Continuous phase or catalyst
CE     Cyclohexylethanol
CH     Cyclohexane
CMK    Cyclohexylmethylketone
d      Dispersed phase
d1     1\textsuperscript{st} carbonyl group hydrogenation: AP \rightarrow PE
d2     2\textsuperscript{nd} carbonyl group hydrogenation: CMK \rightarrow CE
f      Reaction feed chemicals
film   Film region of aqueous phase
int    Interface between organic and aqueous phases
max    Maximum
org    Organic phase
o      Initial value
p      Particle
PE     Phenylethanol
r      Tubular reactor
r1     1\textsuperscript{st} aromatic ring hydrogenation: AP \rightarrow CMK
r2     2\textsuperscript{nd} aromatic ring hydrogenation: PE \rightarrow CE
s      Catalyst surface
t      Total reaction system (dispersed phase + aqueous phase) or Liquid reservoir
w      Water
ABSTRACT


A biphasic stirred tank reactor and a trickle bed reactor were studied to understand complex multiphase reactor behavior arising from mass transfer effects on reactions and to improve modeling accuracy for rational design and optimization.

For the first part involving a stirred tank reactor, the intrinsic reaction rate of n-butyraldehyde aldol condensation was obtained in the industrially relevant range 110-150 °C and 0.76-1.9 M NaOH, which is in the mass transfer regime dominated by reaction in the film. A stirred cell was used to obtain stable interface between the organic and aqueous phases. The mass transfer regime was confirmed by plateau region experiments and calculations of mass transfer. As a result, considering nBAL solubility and diffusivity, the rate was found to be 1st order in both nBAL and NaOH concentrations, along with 13.5±0.4 kcal/mol activation energy. The kinetic parameter sensitivity using different models for solubility, diffusivity and salt effect was also studied. This work demonstrates that, using penetration theory, it is possible to determine intrinsic reaction kinetics in the mass transfer regime, governed by reaction in the film.
Following the first step, reactor modeling for n-butyaldehyde aldol condensation was investigated under the industrially relevant conditions. The interfacial area in the reactor was directly measured using a borescope system under appropriate temperature, NaOH concentration and rpm conditions. To estimate the interfacial area, a semi-empirical correlation was developed, which provided good estimates within ±15% error. The reactor model based on the two-film theory was developed, combining the interfacial area and intrinsic reaction kinetics reported above. The model was verified by reaction experiments in the range 0.05-1.9M NaOH, 80-130 °C and 600-1000 rpm, similar to the industrial conditions. The prediction errors of the reactor model, combining the interfacial area from direct measurements and the correlation were ±8% and ±15%, respectively, suggesting that the model accuracy may be improved with better interfacial area estimation.

For the study of a trickle bed reactor, intrinsic kinetics and internal diffusion effects using various support sizes were investigated for acetophenone hydrogenation. The 1% Rh/Al₂O₃ catalyst was selected by catalyst screening tests using different noble metals and supports in a slurry reactor. Intrinsic reaction kinetic modeling with the Langmuir-Hinshelwood mechanism was conducted from experiments at 60-100 °C, 1.1-4.1 MPa P_H₂ and 0.04-0.4 M C_{AP,o} using powder catalysts. The selected kinetic model included dissociative and non-competitive hydrogen adsorption, along with saturated active sites for organic species, and surface reaction as the rate determining step. With the obtained intrinsic reaction kinetics, internal diffusion effects were investigated using two catalyst particle sizes and diffusion-reaction models. The properties of the egg-shell type catalyst
particles, including metal dispersion, were characterized and utilized in the models. The predictions of the models developed in this work correspond well with the experimental results, explaining the effects of internal diffusion inside catalyst particles on reaction rates and selectivity.

In a trickle bed reactor, flow regime effect and reactor modeling studies were conducted for acetophenone hydrogenation on 1% Rh/Al₂O₃ catalyst, a relatively high pressure and complex reaction scheme typical for pharmaceutical applications. The reactor consisted of a 7.1 mm ID stainless steel tube with 0.5 mm catalyst spheres. From hydrodynamic tests, trickle and bubbly flow regimes were confirmed visually with regime map developed for different gas/liquid, tube/particle materials, pressure and temperature. The operating conditions for each regime were identified using pressure drop fluctuations for the opaque stainless steel reactor. The beneficial effect of bubbly flow on reaction rate was confirmed experimentally in 0.02-0.19 m/s and 2.5-12 mm/s for gas and liquid superficial velocity ranges, respectively under 80-100 °C, 11-26 bar and 0.04-0.6 M C_{AP.o} conditions. The effects of partial wetting and liquid limited reaction were suggested from studies involving gas flow rate, temperature and pressure variation. The reactor model including external/internal mass transfer along with the flow regime effects was developed using an adjustable parameter for partial wetting and flow regime effects. With fitted parameters using a part of the experiments, the model provided good predictions (R² >95%) for all experiments.

The combined experimental and modeling approaches followed in the present work are good examples to demonstrate the effects of mass transfer on reactor performance. This
thesis will help to improve the modeling accuracy for design and scale up with fundamental understanding of multiphase reactors.
CHAPTER 1. INTRODUCTION

Multiphase reaction systems are found in various chemical industries, such as refinery, petro-chemicals, commodity and special chemicals including polymers and pharmaceuticals. Reactors, as the most important unit where raw materials change to desired products, determine pre-process, post-process and even overall process concepts [1, 2]. The value generated by multiphase reactor technology is about $637 billion and 52% derives from fine chemical and pharmaceuticals [3]. Chemical engineers have contributed to improve and maximize the value by designing new reactors and optimizing existing reaction processes. For this, experimental and modeling works are necessary to understand fundamentals and to estimate reactor behaviors, resulting in more effective design and optimization. It is, however, not an easy task because modeling of the reactors requires knowledge of not only intrinsic reaction kinetics but also complex multiphase transport phenomena including mass and heat transfer. Despite significant efforts in this field, many aspects of multiphase reactor behavior remain uninvestigated [4].

Multiphase reactors can be classified, as shown in Figure 1.1, from the view point of phases and phase combinations existing in the reactor [2]. Because there are various types of multiphase reactors, specific reactor systems should be chosen as the target reactors for this study. Biphasic stirred tank reactors and trickle bed reactors are widely used in petrochemical, fine chemical and pharmaceutical industries [5, 6]. The reactors are quite
different in all aspects, and as such are representative examples of most multiphase reaction systems. Therefore, considering mass transfer effects on reactor performance, experimental and modeling studies were carried out for both a biphasic stirred tank reactor and a trickle bed reactor.

1.1 Biphasic stirred tank reactor

A biphasic stirred tank reactor contains two immiscible liquids with liquid phase catalysts, such as acid-base catalysts or organo-metallic catalysts. Industrial applications of the reactor include aldol condensations, hydroformylation, polymerization and pharmaceutical processes [7, 8]. In this reactor, one phase includes reaction feeds and the other phase includes catalysts. Commonly, some feed can be dissolved into the catalyst
phase according to solubility, while catalysts are not transferred to the feed phase. Therefore, the reaction occurs in the catalyst phase. For new chemical synthesis in lab-scale, slow reaction rate is used to determine the reaction mechanism and kinetic equations, ignoring mass transfer effect with uniform feed chemical concentrations in the catalyst phase. In the engineering step for commercial production, however, faster reaction rates are preferred for high productivity. If reactions become fast, feed chemicals react in the zone near the interface between the two liquids because mass transfer rates are not enough to maintain uniform concentrations, called as mass transfer limitation. In this case, knowledge of both interfacial reaction kinetics and interfacial area are necessary to understand the reactor behavior. It is, however, challenging to obtain both, not only because the interfacial reaction kinetics is a combination of mass transfer and intrinsic reaction kinetics but also because measurements and estimations of interfacial area are not straightforward.

1.1.1 Reactive mass transfer in the interface

There are two methods to consider mass transfer in the interface with reaction, the film model and the penetration model. The film model assumes a stagnant film near the interface between the two phases, which is determined by the ratio of diffusivity to convective mass transfer coefficient. The stagnant film explains all mass transfer effects, resulting in uniform chemical concentrations in the bulk region defined except for the film in a phase. In the model, mass transfer is considered as steady state [9]. Without time derivative term, mass balances become ordinary differential equations which are relatively easy to be solved with proper boundary conditions. The model also enables to
combine reactions in the bulk region and provide analytical solutions. Because of steady state assumption, the model may be, however, inappropriate for very fast reaction with no feed species in the bulk region.

The penetration model was developed by Higbie [10] and applied for absorption into turbulent liquid flow with modification by Danckwerts [11]. In the model, materials are transferred by non-stationary diffusion through surfaces and old surface elements are replaced with fresh new elements from the well-mixed bulk region by turbulent fluid motion [12]. With time derivative term, a partial differential equation for mass transfer with reaction was analytically solved with proper boundary conditions, assuming infinite depth stagnant cylinder with zero concentration of penetration material at the infinite boundary [13]. As a complement with the film model [14], the penetration model is appropriate for swallow film penetration of material by low diffusion rate, short exposure time or fast chemical reaction [15].

1.1.2 Measurement techniques for interfacial area

Obtaining the interfacial area value in a stirred tank reactor is difficult due to complex drop breakage and coalescence by turbulent flow from agitation, resulting in drop distribution. For this reason, there are various techniques available to measure interfacial area with advantages and drawbacks for each. They can be categorized as chemical and physical methods. The chemical method obtains interfacial area directly from reaction or absorption experimental data with known intrinsic reaction kinetics using the Danckwert’s penetration theory [16]. As an indirect method, however, it has limitations
because it needs an appropriate reaction system and its intrinsic reaction kinetics which satisfy conditions for the penetration theory. In addition, the accuracy of interfacial area is affected by qualities of physical property data as well as intrinsic reaction kinetics [17].

In the physical method, to obtain the interfacial area value under an operating condition, drop distribution in the reactor vessel is required because interfacial area is, in fact, an averaged value from the drop distribution. There are various techniques to physically measure the drop distribution, including laser diffraction, focused beam reflectance measurement (FBRM) and video probes [18]. The method using a laser diffraction particle size analyzer is widely accepted method with accurate volume distribution. As off-line method, however, sampling is required, which means that the drop distribution can be altered by breakage and coalescence during sampling and measuring in the analyzer [19]. The FBRM provides continuous in situ measurement data from reflectance of a rotating laser using a fiber optic probe (typically 25 mm diameter) composed of a rotating optic, laser diode and detector. The output of the method is a chord length distribution (CLD) which needs mathematical treatment to obtain drop distribution [20]. In the method, thus, there are two difficulties of obtaining the real CLD and its proper treatment. Because of these, the interfacial area from the FBRM may not be accurate enough for model and simulation validation [21]. The video probe method takes pictures of drops inside a tank vessel using a borescope with fiber-optic light guide, a xenon strobe and a camera system, obtaining drop distribution by counting drops from the pictures. Although counting drops is labor-intensive, the video method could provide an accurate interfacial area sufficient for high dispersed liquid-liquid systems [18, 21].
1.1.3 Estimation methods for interfacial area

The most widely used estimation method for interfacial area is semi-empirical correlation based on the Hinze-Kolmogorov concept of turbulence [22]. The correlations provide the sauter mean diameter, $d_{32}$, which is directly converted to interfacial area [23]. The Hinze-Kolmogorov concept provides the prediction of maximum drop diameter, $d_{max}$, from the critical Weber number with isotropic turbulence assumption, if drop diameters are significantly larger than Kolmogorov’s length scale and the drops are inviscid [24]. The $d_{max}$ is connected to $d_{32}$ with assumption that $d_{32}$ is proportional to $d_{max}$, which is generally accepted by most workers [22, 23]. The correlation was extended to consider coalescence effect or turbulent damping when dispersed phase fraction is increased [25]. The semi-empirical correlation obtained through the above procedure is composed of the Weber number, the turbulent damping factor and adjustable constants. In addition, the correction terms composed of viscosities were used to obtain more general correlations for different chemical species and high viscosity cases [26, 27]. Although the correlation is simple and easy to use, the adjustable constants have to be determined to explain different reactor geometry and chemical properties [28]. For more fundamental approach, drop breakage and coalescence should be more rigorously considered to obtain interfacial area.

The interfacial area concentration (IAC) has been developed and used to simulate more accurate flow field in the boiler side of nuclear reactors. It is a transport equation of the interfacial area concentration with source terms to represent drop breakage and coalescence with their times and efficiencies [29]. Although the population balance
equation (PBE) is more general approach for drop distribution with rigorous breakage and coalescence for each drop sizes [30], its calculation cost may be too expensive for obtaining the interfacial area.

1.2 Trickle-bed reactor

A trickle bed reactor is a 3-phase packed bed reactor which is composed of cocurrent downflow of gas and liquid in solid catalyst bed. The applications of the reactor are widespread, including hydrogenation and wet oxidation, from petroleum and chemical to pharmaceutical industries [6, 31]. In particular, the pharmaceutical industry is characterized by heavy molecular weight feed species, complex reaction route and high operating pressure [5]. Three phases coexisting in the reactor lead to complex hydrodynamics which can significantly affect reactor performance including conversion and selectivity. These hydrodynamics can be categorized in four different flow regimes mainly according to different flow rates of gas and liquid in the fixed bed. The flow regimes affect mass, heat transfer and hydrodynamic properties including liquid holdup, wetting efficiency, pressure drop, which subsequently influence reactor performance [32]. A few articles have reported that operation in the pulsing flow regime enhances reaction rate and selectivity at atmospheric pressure [33, 34], while most trickle bed reactors are operated in the trickle flow regime where gas and liquid flow as continuous phases at relatively low superficial velocities. The flow regime effect for a complex reaction with high pressure has not been investigated. To model the trickle bed reactor with flow regime effect, it is necessary to know complex multiphase external mass transfer affected
by flow regimes as well as intrinsic reaction kinetics and pore diffusion effect inside catalyst particles [32].

1.2.1 Flow regimes and detection methods

The flow regimes in trickle beds are typically classified as follows: trickle flow, spray flow, pulsing flow and bubbly flow [35]. While trickle flow obtained in low gas and liquid flow rates is calm and stable, pulsing flow is characterized by alternatively changing gas and liquid rich zones at moderately increased gas and liquid flow rates from trickle flow. Both pulsing and bubbly flows are considered as high interaction regimes, resulting in significantly enhanced heat and mass transfer rates by strong interactions between the gas and liquid phases [36-38]. The regime transition depends on various factors including particle size and distribution, physical properties of fluids, gas and liquid superficial velocities [32]. Among these, the superficial velocities are commonly used as the manipulated variables [39, 40]. The representative detection method of regime transition is visual inspection [41, 42] as an easy and clear method, but transparent tubes are necessary for the method. Because transparent tubes are generally inappropriate for high pressure, temperature and strong organic solvents, which are common in reactions, other alternative methods have been developed as pressure drop [40, 43], pressure drop fluctuation [44, 45], liquid holdup, electrical conductivity, CT (computed tomography), etc [32].
1.2.2 Modeling of trickle bed reactors

As a 3-phase reactor, complex hydrodynamics and flow regimes of gas and liquid flows on solid particles affect external mass and heat transfer coefficients, dispersion coefficients, the uniformity of liquid flow in the bed and external wetting efficiency on catalyst particle surfaces [32]. If the catalyst surface is fully covered by liquid film, i.e. full wetting, the liquid film separates gas region and solid surface, generating two interfaces gas-liquid and liquid-solid. In this case, gas reactants in gas region are transferred to catalyst surface through gas-liquid and liquid-solid mass transfer resistances, while only liquid-solid mass transfer resistance exists for liquid chemicals [46]. With gas and liquid downflows, partially wetted particle surface is more common, although its modeling becomes more complex. A reactor model for the partial wetting suggested three zones for catalyst particle surfaces [47]. One is dry surface which can directly contact the gas phase without liquid film, while another part of the wet surface is affected by the stagnant liquid in the space between solids. The last part of the wet surface is the same as with fully wet surface model. The approach is, however, complicated including portions of three zones, interactions between the zones and the effectiveness factor for the dry zone, which are not easy to obtain. For this reason, the term for the dry zone is often ignored for high wetting efficiency [48].

1.3 Research objectives

Based on the above background, experimental and modeling studies were planned to investigate mass transfer effects on multiphase reactions in stirred tank and trickle bed reactors. For stirred tank reactors, n-butyraldehyde aldol condensation catalyzed by
sodium hydroxide was selected because it is an important reaction in the petrochemical industry and is relatively fast 1st order reaction with pseudo steady state approximation. In any prior works, the intrinsic reaction kinetics was previously not studied in the industrial operating conditions where the reaction is in mass transfer controlled regime. The operating condition was in the available ranges for the penetration model with low rpm and the film model with high rpm. In addition, there was no publication available on intrinsic reaction kinetic study in the mass transfer regime, interfacial area study in high sodium hydroxide concentration and biphasic reactor model with two film resistance and reactions including bulk region for batch operation. For trickle bed reactors, the target reaction was acetophenone hydrogenation. Acetophenone is the simplest ketone having benzene ring and its reaction scheme is complex with six reaction routes, 3-phase hydrogenations and hydrogenolysis under 10-70 bar, relatively high pressure, which is appropriate for flow regime effect on reactor performance in trickle bed reactors for pharmaceutical applications. The flow regime effect for high pressure and complex reaction has not been investigated in any prior works. The study on intrinsic reaction kinetics and pore diffusion effects using a slurry reactor, which is free from external mass transfer, is necessary and it would help to understand the behavior of the trickle bed reactor by comparison. Thus, the research objectives of this thesis are as follows:

- Determine intrinsic reaction kinetic equation for n-butyraldehyde aldol condensation in the mass transfer controlled regime using the penetration theory
• Evaluate and develop a biphasic stirred tank reactor model for aldol condensation with interfacial area estimation model based on interfacial area measurement data obtained under high sodium hydroxide and temperature conditions.

• Determine intrinsic reaction kinetics and develop a pore diffusion model for acetophenone hydrogenation after catalyst screening for flow regime effect study in a trickle bed reactor.

• Investigate and develop a trickle bed reactor model to explain experimental confirmation of flow regime effects on reactor performance under high pressure and the complex reaction for pharmaceutical applications.
CHAPTER 2. KINETIC STUDY OF BIPHASIC ALDOL CONDENSATION OF N-BUTYRALDEHYDE USING STIRRED CELL

2.1 Introduction

Aldol condensation of normal-butyraldehyde (nBAL) is a part of the oxo-alcohol process to produce 2-ethyl-1-hexanol (2EH) which is commonly used to manufacture plasticizers for polyvinylchloride (PVC) [49]. The reaction mechanism of aldol condensation has been well studied as three steps for aldol and one step for condensation of water [50], which can be combined as the two steps shown below [51]. Normal-butyraldehyde is attacked by base catalyst to yield the intermediate (Eq. 2.1), which in turn reacts with another aldehyde and 2-ethyl-1-hexenal (2EHEL) is produced along with the regenerated catalyst (Eq. 2.2). The rate of nBAL consumption can be expressed using the pseudo-steady state approximation (Eq. 2.3), which can be further simplified into a first-order reaction (Eq. 2.4) when the 1st step \( k_1 \) in Eq. 2.1 is the limiting step [52].

\[
\begin{align*}
C_3H_7CHO + OH^- & \xrightarrow{k_1} \frac{k_1}{k_2} C_3H_6CHO^- + H_2O \quad (2.1) \\
C_3H_6CHO^- + C_3H_7CHO & \xrightarrow{k_3} C_7H_{13}CHO + OH^- \quad (2.2) \\
-\frac{dC_{nBAL}}{dt} & = \frac{k_1 k_3 C_{nBAL} C_{OH^-}}{k_2 + k_3 C_{nBAL}} \quad (2.3) \\
-\frac{dC_{nBAL}}{dt} & = k_1 C_{nBAL} C_{OH^-} \quad (2.4)
\end{align*}
\]
Sodium hydroxide (NaOH) in aqueous phase is the most widely used catalyst for the aldol condensation of nBAL [49]. Due to relatively low solubility of nBAL in the aqueous phase (<10 wt %), this reaction is conducted in a biphasic system where nBAL and 2EHEL are present in the organic phase while the catalyst (e.g. NaOH) is in the aqueous phase. The aldol condensation is operated in the industrial scale at temperature 80-140 °C, pressure ~5 bar and 2-4 wt % catalyst (e.g. NaOH) in water [49].

Some efforts have been made in the prior literature to determine the intrinsic reaction kinetics of aldol condensation of nBAL. Most of the kinetic studies were carried out at temperature 20-50 °C to avoid mass transfer limitations [53]. Although the kinetics were obtained in the reaction controlled regime, the temperatures employed in these studies were much lower than industrial conditions. In addition, the reaction order for nBAL in some prior studies was 2, as seen from Eq. (2.3) because they used low nBAL concentration (below 10^{-3} M) for slow reaction rate [54] or used solvent [55, 56] which can alter the kinetics [51, 52]. Beltrame et al. (1973) obtained the 1st order kinetics accounting for the solubility of nBAL in the biphasic system, without dilution or using solvent in the reaction controlled regime [57]. However, the effect of temperature on nBAL solubility was not considered in their study, limiting the value of their results.

As noted above, prior studies regarding the kinetics of aldol condensation reaction were conducted at much lower temperature than industrial operating conditions. In addition, the reported activation energy values vary widely. A further complexity is that the product distribution is different at lower and higher temperatures (see section 2.4.6). Thus, for successful design of chemical reaction systems, the intrinsic reaction kinetics obtained
under the actual operating conditions are preferred to design and optimize the process. Under these conditions, however, the reaction occurs in the mass transfer controlled regime. Thus, it is preferable to conduct intrinsic kinetic study in the mass transfer controlled regime, which makes determination of the intrinsic kinetics difficult.

For nearly immiscible biphasic systems, stirred cell was designed to study mass transfer and also measure interfacial area using first-order reactions [58]. A stirred cell is similar to a stirred tank reactor, except that it is operated at low mixing speed to maintain stable interface between the two phases. Despite some fluctuation of the interface even under low mixing speed, the error caused by it is reported to be acceptably small for well-designed device under appropriate rpm. Further, several studies have shown that the kinetic parameters, such as activation energy, using stirred cell agree well with those obtained using stirred tank in the reaction-controlled regime [59, 60]. Thus, a stirred cell can be used to measure the kinetics of aldol condensation of nBAL, which, as described in section 2.4.1, was confirmed as pseudo first-order reaction under high nBAL concentrations (Eq. 2.4), even under the industrially important higher temperature and mass transfer controlled regime.

In order to study the kinetics in mass transfer controlled regime using stirred cell, the reactive and mass transfer behaviors in the interfacial region between the two phases must be understood. In this work, the film and penetration theories were utilized to model mass transfer in the organic and aqueous phases, respectively. The film theory considers the film accounting for majority of the concentration difference between the interface and the bulk region. On the other hand, penetration theory is useful to understand the concentration profile in the interfacial region when fast reaction occurs. In this case, the
mass transfer flux through the interface can be expressed as a function of reaction kinetics and diffusivity of each component [12]. For this reason, penetration theory has been used to estimate interfacial area for various biphasic systems with fast reactions [23, 59, 61].

In this paper, using a stirred cell, intrinsic kinetics of biphasic nBAL aldol condensation were studied in the mass transfer controlled regime. The solubility and diffusivity of components and salting effect were estimated using published data and several models (NRTL, UNIQUAC, etc.). The penetration theory was employed to understand the interfacial region in the aqueous phase side. The activation energy calculated in this work was compared with the values obtained in prior studies. It is noteworthy that this study is the first one investigating the intrinsic kinetics of the nBAL aldol condensation at industrial operating temperatures. Thus, our intrinsic kinetics can be used to design and optimize the nBAL aldol condensation process in the industrial scale. Finally, the method introduced in this study can also be applied for kinetic studies of other reactions which occur in the mass transfer controlled regime.

2.2 Theory

2.2.1 Penetration model and plateau region

For the case of high nBAL concentration, so that Eq. (2.4) applies, the mass transfer with 1st or pseudo first-order reaction on the interface of two-phase systems with cylindrical inclusions can be described by Eq. (2.5), which was first solved by Higbie (1935) [10].
Figure 2.1: Schematic diagram of the biphasic interface.

This model can be applied for the aqueous side interfacial region where the aldol condensation occurs (see Figure 2.1).

\[
\frac{\partial C_{nBAL}}{\partial t} = D_{nBAL,aq} \frac{\partial^2 C_{nBAL}}{\partial x^2} - k C_{nBAL}, \quad \text{where} \quad k = k' e^{-\frac{E_a}{RT}} C_{NaOH}
\]  

(2.5)

With the initial and boundary conditions:

\[
\begin{align*}
C_{nBAL} &= C_{nBAL,aq}^*; \quad x = 0, \quad t > 0 \\
C_{nBAL} &= 0; \quad x > 0, \quad t = 0 \\
C_{nBAL} &= 0; \quad x = \infty, \quad t > 0
\end{align*}
\]

For the case when \( kt \) is sufficiently large (\( kt > 4 \)), Danckwerts (1950) [13] simplified the solution to provide the mass transfer flux of nBAL (\( J_{nBAL} \)) as follows:

\[
R_{nBAL} = J_{nBAL} = -D_{nBAL,aq} \left( \frac{\partial C_{nBAL,aq}}{\partial x} \right)_{x=0} = C_{nBAL,aq}^* \left( D_{nBAL,aq} k \right)^{0.5}
\]  

(2.6)

The convective mass transfer (\( k_L \)) caused by the agitator in the aqueous phase must be also considered for the stirred cell [12]. With this, and accounting for catalyst
concentration and solubility of nBAL which is influenced by composition of the organic phase, Eq. (2.6) becomes

\[ R_{nBAL}a = amC_{nBAL,org} \left( D_{nBAL,aq}k'e^{-\frac{E_a}{RT}C_{NaOH} + k_L^2} \right)^{0.5} \]  \hspace{1cm} (2.7)

The intrinsic reaction kinetics can be obtained under the mass transfer regime dominated by reaction kinetics, so that \( k_L \) in Eq. (2.7) can be ignored. The Hatta number (\( Ha \)) is a dimensionless quantity that can be used as the criterion to determine which model, among the penetration and the film theories, is more appropriate to explain the mass transfer in the interfacial region.

\[ Ha = \left( \frac{D_{nBAL,aq}k'e^{-\frac{E_a}{RT}C_{NaOH}}}{k_L} \right)^{0.5} \]  \hspace{1cm} (2.8)

The Hatta number is a ratio of mass flux enhanced by the chemical reaction to that by mass transfer alone. To exclude the effect of mass transfer on the reaction kinetics, \( Ha \) should be larger than 3 \[16\]. Most published studies employed this criterion to apply penetration model for stirred cell \[16, 60, 62\]. This regime can be experimentally confirmed by the plateau region where the mass flux of nBAL through the interface is not influenced by the agitator speed (rpm) under the condition that the interface is stable. Thus, the mass transfer rate in this region is dominated by chemical reaction. In summary, in the experimentally confirmed plateau region, the intrinsic kinetic parameters can be calculated using Eq. (2.7) once \( m \) and \( D_{nBAL,aq} \) are properly determined.

Note that the penetration theory described above is valid only under the following assumptions and conditions:
a. The reaction is sufficiently fast, so that $c_{nBAL,aq} \text{ at } x = \delta$ is nearly zero.

b. The solubility of nBAL and 2EHEL in the aqueous phase is sufficiently low to avoid mass transfer limitation in the organic phase.

c. Water produced during the reaction is small, so that dilution of $C_{NaOH}$ can be ignored.

### 2.2.2 Solubility of nBAL

As described in the section 2.2.1, the solubility of nBAL in aqueous phase ($m$) is an important factor to determine intrinsic kinetic parameters of aldol condensation of nBAL. As shown in Eq. (2.9), to calculate the solubility of nBAL in aqueous phase (i.e., NaOH in water), both salt effect term and solubility of nBAL in water ($m_w$) must be determined. It is well known that solubility of component decreases with salt formation and this salt effect can be expressed as an exponential function of ionic strength, $I$ and salting-out parameter, $k_s$ [63].

$$m = \frac{c_{nBAL,aq}^*}{c_{nBAL,org}} = m_w \cdot 10^{-k_s I} \quad (2.9)$$

First, to determine $m_w$, two methods, Non-Random Two Liquid (NRTL) [64] and Universal Quasi-Chemical Activity Coefficient (UNIQUAC) [65] can be considered. They are the most widely used activity coefficient models to calculate phase equilibrium. In this work, we employed UNIQUAC to estimate activity coefficients, which were compared with those obtained using the NRTL method (see section 2.4.4). The binary parameters were determined from mutual solubility data of nBAL/water and 2EHEL/water [66]. In addition, vapor-liquid equilibrium (VLE) for nBAL/2EHEL was
predicted by the UNIFAC (UNIQUAC Functional-group Activity Coefficients) group contribution method [67]. The fitting accuracy of the activity coefficient models with published data was 99% or higher. Several experiments were conducted to verify the effect of 2EHEL addition on solubility of nBAL in water. It was found that the solubility of nBAL in water decreases with 2EHEL addition, which is in good agreement with the estimation by UNIQUAC, as shown in Figure 2.2. Additional experimental details and results are discussed in Appendix A.

Figure 2.2: Effect of 2EHEL on nBAL solubility; Initial: 14ml organic phase of nBAL/2EHEL, Final: mixed with 36ml water; Est.1: estimated by UNIQUAC, Est.2: estimated by nBAL solubility in water, Exp.: experiment data, See Appendix A for experiment details.

In the initial stage of the reaction, nBAL is the organic phase and NaOH is dissolved in the aqueous phase. As the reaction proceeds, the nBAL is catalytically converted to
2EHEL, which is nearly immiscible in water. Thus, the 2EHEL produced in the aqueous phase moves to the organic phase where only nBAL was present initially. Further, the solubility of nBAL in aqueous phase can be influenced by the change of composition of the organic phase due to the influx of 2EHEL. In this context, it is important to examine the phase equilibrium of nBAL/2EHEL/water mixture to determine \( m_w \), a distribution coefficient of nBAL:

\[
m_w = \frac{C_{nBAL,w}^*}{C_{nBAL,org}}
\]  

Figure 2.3 shows the concentration of nBAL in the interface of water phase \( (C_{nBAL,w}^*) \), calculated using the UNIQUAC method, as a function of the concentration of
nBAL in organic phase ($C_{nBAL,\text{org}}$). It clearly shows that $C^*_{nBAL,w}$ increases with $C_{nBAL,\text{org}}$ while it decreases with temperature increase. It was also found that $C^*_{nBAL,w}$ is somewhat nonlinear with $C_{nBAL,\text{org}}$ for a given temperature, but was approximated to be linear in this work. The slope was calculated by linear regression and $R^2$ value by the least squares method was 0.99 ±0.01. The slope of each constant temperature curve was evaluated as $m_w$.

As seen in Eq. (2.9), the salt effect can be expressed as an exponential function of ionic strength and salting-out parameter, which can be obtained from Eqs. (2.11) and (2.12) shown below [63]. In addition, the temperature dependence of salting-out parameter, $k_s$, can be found from Eq. (2.13) [63].

\[
I = 0.5(C_{Na^+}Z_{Na^+}^2 + C_{OH^-}Z_{OH^-}^2) 
\]

\[
k_s = i_{nBAL} + i_{Na^+} + i_{OH^-} 
\]

\[
\frac{dk_s}{dT} \approx -\frac{\nu_s^0}{2.3RT} \frac{d\nu_s^0}{dT} 
\]

The empirical parameters of Na$^+$ and OH$^-$ ions ($i_{Na^+}$ and $i_{OH^-}$) can be found from several publications [68, 69]. Santiago and Bidner [68] reported $i_{Na^+} = 0.07678 \text{ m}^3/\text{kmol}$ and $i_{OH^-} = 0.08996 \text{ m}^3/\text{kmol}$ at 27 °C and these values were used in the present work. However, no empirical parameter for nBAL was found from prior work. For this reason, the value for nBAL ($i_{nBAL}$) was estimated to be 0.0098 m$^3$/kmol from the value for normal-butyl formate [68], since it is known that the salting out parameter has a proportional relation with the molar volume [70].
Although $k_s$ was obtained at 27 °C, its temperature dependence must also be considered to better estimate the salting-out parameter. Unfortunately, the temperature dependence has been studied only in the range 0-50 °C [63]. In this work, therefore, the parameter was extrapolated to investigate the higher temperature range (110-150 °C) from the published values by assuming that the derivative of partial molar volume $\left(\frac{dV^\circ}{dT}\right)$ becomes half with every 25 °C increase [63]. With this assumption, the salting-out parameter varies from 0.142 to 0.139 with increasing temperature from 110 to 150 °C. The validity of these values will be discussed more with the parametric study in section 2.4.4.

2.2.3 Diffusivity and mass transfer coefficient

The diffusivity, ($D_{nBAL,aq}$) was calculated using three different models [71-73] while the mass transfer coefficients, $k_L$ was estimated using the correlation of Calderbank and Moo-Young (1961) [74]. The details for both calculations are presented in Appendix B. With these results, using Eq. (2.7), the intrinsic kinetics of nBAL aldol condensation can be obtained under industrial conditions in the mass transfer regime dominated by reaction kinetics.
2.3 Experimental

2.3.1 Experimental apparatus

The experiments were conducted in a 300 mL stainless steel reactor (Parr Instrument Company, Model 4592) with external heating and water-flow cooling coil for temperature stabilization, as shown in Figure 2.4.

Figure 2.4: Schematic diagram of the stirred cell system; 1- stirred cell reactor, 2 - nBAL injector, 3 – ice water bath; PI – pressure indicator, TI – temperature indicator, TIC – temperature indicator / controller.
The reactor height to diameter ratio was 1.6. The four blade and 45°-pitched paddle was used to agitate the solution. The aqueous solution was poured inside the reactor under a nitrogen (99.995% pure) environment. Two thermocouples were positioned to read temperature at the interface between phases. One thermocouple was located near the cooling coil while the other on the opposite side to confirm uniform temperature. The organic feed (nBAL) was injected to the reactor through a nozzle by pressurized nitrogen. The injector holding the organic feed was preheated by heating tape with temperature controller. The nozzle was designed to minimize the fluctuation of the interface level during injection of the organic feed. A thin tube with 1/16” OD was inserted in the middle of the organic phase and small amounts of the organic phase were sampled at specific interval through the tube to analyze its composition. The sampling tube was dipped inside an ice water bath to quench the samples. After each sampling, the tube lining was cleaned with nitrogen to remove any residue of the previous sample.

2.3.2 Analysis of the organic phase composition

The composition of the organic phase was analyzed by GC (HP 5890 II) equipped with an Agilent DB-WAXetr capillary column (50 m x 0.32 mm) and flame ionization detector under the following conditions: helium as carrier gas (2.6 ml/min), inlet temperature 250 °C, oven temperature from 80 to 220 °C, detector temperature 250 °C and injection volume 0.5 μl with split. The composition was also confirmed by HPLC (Agilent 1100) with an Agilent SB-C18 (5 cm x 4.6 mm) column and diode array detector (380 nm). In this case, the aldehyde was analyzed using the derivatization method with DNPH (2, 4-dinitrophenylhydrazine, Sigma-Aldrich, >97%) [75]. Acetonitrile (Sigma-
Aldrich, >99.9%) and hexanal (Sigma-Aldrich, > 98%) were selected as internal standards for the GC and HPLC, respectively. Normal-butyraldehyde (nBAL, Sigma-Aldrich, >99%) and 2-ethyl-1-hexenal (2EH, Sigma-Aldrich, >93%) were used for calibrations of GC and HPLC. The R² values of calibration for the GC and HPLC were calculated to be 99.6 and 99.8 %, respectively. The GC method was confirmed by the data from HPLC, with below 5 % error.

2.3.3 Experimental Procedure

The NaOH solution (50 % in water, Sigma-Aldrich) was diluted with deionized water to specific concentrations which were confirmed by both pH meter (OMEGA PHB-209) and titration method with 1N HCl solution and bromothymol blue as indicator. The titration method was found to be more accurate for strong base (> pH 12), as compared to the pH meter measurement. To conduct the reaction, 180 mL of NaOH aqueous solution was preloaded inside the reactor under a nitrogen (99.995 % pure) environment. Starting at room temperature, the reaction vessel and the injector were heated to target temperature by the PID controller. When the reactor temperature reached the target value, 24 mL of nBAL was injected to the reactor through the nozzle. The reactor was pressurized with nitrogen at ~ 10 bar to prevent azeotropic boiling between nBAL and water during injection of nBAL and initiation of the reaction. The difference between set point and actual reactor temperature was within ± 1.0 °C. In addition, the temperature difference between two thermocouples inside the reactor increased with reactor temperature, but the maximum and average values were 3 and 2 °C, respectively. A small
amount (~0.3 mL) of the organic phase was sampled once every 5 minutes after nBAL injection, for a total 20 minutes. The pH value of the aqueous solution was measured after each experiment and the difference of the pH value between before and after reaction was found to be less than 5 %. It is expected that some water produced by aldol condensation results in a small pH decrease.

The mass balance was checked after each experiment at the reference temperature after cooling, and the error was always less than 1%.

2.4 Results and discussions

2.4.1 Determination of the range of experimental conditions

To obtain intrinsic kinetic parameters for nBAL aldol condensation in the mass transfer regime, the range of experimental conditions where the mass transfer is dominated by the reaction kinetics must be determined. For this reason, several experiments were carried out to set the ranges of temperature, NaOH concentration and the speed of agitator.

It is important to maintain stable interface between the organic and aqueous phases to measure the interfacial reaction rate in the stirred cell. Therefore, the speed of agitator must be sufficiently low to provide a stable interface. The direct observations of interfacial level were conducted in a beaker with similar dimensions as the stirred cell used in this study. The stability or fluctuation of the interface between organic and aqueous phases was monitored while varying the speed of agitator. It was found that the interface was relatively stable for 50-80 rpm, while further increase resulted in fluctuated
level. Note that similar results were obtained from the 3D-CFD simulation, as presented in Appendix C.

It has been reported that the reaction order of nBAL aldol condensation can be 1 or 2, depending on the nBAL concentration [53-57]. In this work, to confirm the reaction order, the concentration of nBAL was measured with time at various temperatures and NaOH concentrations while the mixing speed was maintained constant at 50 rpm, as described in the Experimental section. From the experimental data and the integration of

\[ \frac{dC_{n\text{BAL}_{\text{org}}}}{dt} = K C_{n\text{BAL}_{\text{org}}}^m, \]

where \( K \) includes all the system parameters besides \( C_{n\text{BAL}_{\text{org}}} \), the best fit value of \( m \) was found to be \( m = 1.003 \pm 0.019 \).

According to a pseudo first-order reaction at the interface, the reactor mass balance Eq. (2.14) can be solved to yield Eq. (2.15).

\[
V_{\text{org}} \frac{dC_{n\text{BAL}_{\text{org}}}}{dt} = -A_{\text{surf}} m C_{n\text{BAL}_{\text{org}}} (D k'' C_{\text{NaOH}})^{0.5}; C_{n\text{BAL}_{\text{org}}} = C_{n\text{BAL}_{\text{org}}}^0 \text{ at } t = 0
\]

\[ (2.14) \]

\[
\ln C_{n\text{BAL}_{\text{org}}} = -\frac{A_{\text{surf}}}{V_{\text{org}}} m (D k'' C_{\text{NaOH}})^{0.5} t + \ln C_{n\text{BAL}_{\text{org}}}^0; k'' = k' e^{-\frac{E_a}{RT}}, k = k'' C_{\text{NaOH}}
\]

\[ (2.15) \]
As shown in Figure 2.5, the reaction follows first-order kinetics, exhibiting linear relation of $\ln C_{nBAL_{org}}$ vs. time ($t$). Thus, the reaction order of nBAL can be concluded to be 1, as also observed in prior studies [52, 57]. In this calculation, the volume change of the organic phase was ignored since density of nBAL is similar to that of 2EHEL and the solubility of 2EHEL in water is quite low (<0.15 wt%) [66]. These results suggest that the intrinsic kinetic parameters ($k'$ and $E_a$) can be obtained from $m(Dk)^{0.5}$, which is the slope of $\ln C_{nBAL_{org}}$ versus time ($t$), using properly determined solubility ($m$), diffusivity ($D$) and interfacial area ($A_{surf}$) from the stable interface of the stirred cell. The $A_{surf}$ value was calculated from the cross sectional area of the reactor, accounting for the shaft, cooling tube and thermocouples.
The error of repeatability of $k''$ obtained from the stirred cell experiments was about 5% or lower, although the error of repeatability of each sample point was about 2%. The error is magnified because $k''$ is the slope of the straight line involving four samples.

Finally, using the criterion $Ha > 3$ (see Eq. (2.8)), 110-150 °C and 0.38-1.9 M, as the ranges of temperature and NaOH concentration respectively, were determined to exclude the effect of bulk aqueous phase mass transfer on the reaction kinetics, i.e. $k_L$ in Eq. (2.7) can be ignored.

2.4.2 Confirmation of plateau region

The ranges determined by $Ha$ described above, were also confirmed experimentally by observing the plateau region where the mass transfer rate is dominated by reaction kinetics. From the ranges, four conditions in each corner (110 °C, 150 °C and 0.38 M, 1.9 M) and one condition in the middle of the range (120 °C, 0.76 M) were selected. Then, $m(Dk)^{0.5}$ values in Eq. (2.15) were calculated for varying speed of agitator from 40 to 80 rpm. As shown in Figure 2.6, the $m(Dk)^{0.5}$ values were essentially constant at the selected conditions and the standard deviations were calculated to be 2-7%. These results indicate that the effect of bulk aqueous phase mass transfer on the mass flux in the interface $(m(Dk)^{0.5})$ can be ignored in the tested range.
To further confirm this feature, variation of $k_L$ with the agitation speed was also calculated. The $k_L$ represents the extent of mass transfer in the aqueous phase and can be calculated using Eq. (B.6) in Appendix B. In Eq. (B.6), $P/V_c$ is calculated from the speed of agitator. Using the result of Appendix C, our calculations showed that with increasing the mixing speed from 40 to 80 ppm, the values of $k_L$ and $P/V_c$ increased by 1.7 and 8 times, respectively. These results suggest that under the condition of the selected agitation speed (40-80 ppm), the mass flux in the interface was not influenced by the agitation speed, although the extent of mass transfer was changed significantly.

Based on these results, it can be concluded that under the selected conditions (110-150 °C, 0.38-1.9 M and 40-80 rpm), most of the nBAL aldol condensation occurs in the film region, suggesting that reaction in the bulk region of the aqueous phase can be ignored. Therefore, the assumptions for the film reaction are satisfied experimentally and the
intrinsic kinetic parameters of the reaction can be obtained under the experimental conditions described above.

### 2.4.3 Kinetic study

Experiments were conducted to calculate intrinsic kinetic parameters at 110-150 °C \( (C_{nBd,org}^a : 9.78 - 9.06 \text{ M, respectively}) \), 0.76-1.9 M and 50 rpm, all within the ranges determined in sections 2.4.1 and 2.4.2 (Table 2.1).
Table 2.1: Experimental results for intrinsic reaction kinetics using stirred cell; agitator speed 50 rpm

<table>
<thead>
<tr>
<th>$C_{NaOH}$ (kmol/m$^3$)</th>
<th>Temp. (°C)</th>
<th>$m(Dk)^{0.5}$ (m/s)</th>
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<tr>
<td>0.76</td>
<td>110</td>
<td>4.19E-06</td>
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<tr>
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</tr>
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</table>
Figure 2.7: Arrhenius plots for four different NaOH concentrations.

The reaction rate constant ($k''$) in Eq. (2.15) was estimated for different temperatures using nonlinear least-square fitting of the experimental data. Figure 2.7 shows the Arrhenius plot used to estimate the activation energy ($E_a$) and the pre-exponential factor ($k'$). The $R^2$ values by the least squares method for the data shown in Figure 2.7 were 0.97-0.99. It can be seen that the linear regression plots for various NaOH concentrations are close to each other (see definition of $k''$ in Eq. (2.15)), which clearly indicates that our proposed method is appropriate to determine the intrinsic kinetic parameters. The average values of the activation energy and the pre-exponential factor were determined to be 13.47±0.39 kcal/mol and 1.712x10^8 m^3/kmol.s, respectively. The detailed values for each case are listed under UNIQUAC/Tyn-Calus in Table 2.2.
Table 2.2: Activation energies calculated using different models

<table>
<thead>
<tr>
<th>( C_{NaOH} )</th>
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<th>NRTL</th>
<th>UNIQUAC</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Tyn-Calus</td>
<td>Tyn-Calus</td>
<td>Hayduk-Minhas</td>
</tr>
<tr>
<td></td>
<td>( E_a )</td>
<td>( R^2 )</td>
<td>( E_a )</td>
</tr>
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<td>0.76</td>
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<td>99.9%</td>
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</tr>
<tr>
<td>1.14</td>
<td>13.39</td>
<td>97.9%</td>
<td>14.02</td>
</tr>
<tr>
<td>1.52</td>
<td>13.83</td>
<td>97.6%</td>
<td>14.46</td>
</tr>
<tr>
<td>1.9</td>
<td>12.95</td>
<td>97.4%</td>
<td>13.58</td>
</tr>
<tr>
<td>Average</td>
<td>13.47</td>
<td>98.2%</td>
<td>14.10</td>
</tr>
<tr>
<td>Standard deviation</td>
<td>0.39</td>
<td>0.39</td>
<td>0.40</td>
</tr>
</tbody>
</table>

Using these results, the following reaction rate equation for nBAL aldol condensation is obtained:

\[
r_{nBAL} = 1.712 \times 10^8 e^{\frac{13.47\pm0.39 \text{ kcal/mol}}{RT}} C_{NaOH}C_{nBAL} \text{ (k mol/m}^3\cdot\text{s) } (2.16)
\]

The exponent of NaOH concentration in Eq. (2.15) is expected to be 1, according to the suggested reaction mechanism and the resulting kinetics, Eqs (2.1)-(2.4). In the present study, the exponent of NaOH concentration was also confirmed by assuming that the empirical parameter of nBAL \((i_{nBAL})\) in Eq. (2.12) is 0.0098 as shown in section 2.2.2. As expected, the exponent of NaOH was determined to be \(\sim1 \ (0.98 \pm 0.11)\) (Figure 2.8 and Table 2.3 for \(i_{nBAL}=0.0098\)).
Figure 2.8: Plot of $\ln k$ versus $\ln C_{\text{NaOH}}$ for five different temperatures (---: linear regression).

Table 2.3: Effect of empirical parameter for nBAL ($i_{\text{nBAL}}$) on the exponent of $C_{\text{NaOH}}$

<table>
<thead>
<tr>
<th>Temperature ($^\circ$C)</th>
<th>$i_{\text{nBAL}}$</th>
<th>0.013 (n-butylformate)</th>
<th>0.054 (benzene)</th>
<th>0.0098 (nBAL)</th>
</tr>
</thead>
<tbody>
<tr>
<td>110</td>
<td>0.000</td>
<td>1.07</td>
<td>1.37</td>
<td>1.12</td>
</tr>
<tr>
<td>120</td>
<td>0.013</td>
<td>0.87</td>
<td>1.17</td>
<td>0.93</td>
</tr>
<tr>
<td>130</td>
<td>0.054</td>
<td>0.82</td>
<td>1.13</td>
<td>0.88</td>
</tr>
<tr>
<td>140</td>
<td>0.0098</td>
<td>0.82</td>
<td>1.13</td>
<td>0.88</td>
</tr>
<tr>
<td>150</td>
<td>0.000</td>
<td>1.02</td>
<td>1.32</td>
<td>1.07</td>
</tr>
<tr>
<td>Average</td>
<td>0.92</td>
<td>0.99</td>
<td>1.22</td>
<td>0.98</td>
</tr>
<tr>
<td>Standard deviation</td>
<td>0.11</td>
<td>0.11</td>
<td>0.11</td>
<td>0.11</td>
</tr>
</tbody>
</table>
2.4.4 Sensitivity study of activation energy and exponent of $C_{\text{NaOH}}$

In order to calculate the intrinsic kinetic parameters ($E_a$, $k^*$) based on penetration theory, values for solubility ($m$) and diffusivity ($D_{n\text{BAL},aq}$) of nBAL must be properly determined. In the calculations using Eq. (2.15), $m$ and $D_{n\text{BAL},aq}$ were estimated using UNIQUAC (Abrams and Prausnitz, 1975) and Tyn-Calus (1975) models, respectively. The salting-out parameter ($k_s$) was also estimated from empirical values with appropriate assumptions (see section 2.2.2). The uncertainties of the kinetic parameter values, however, can arise from uncertainties of the input estimates (e.g. $m$, $D_{n\text{BAL},aq}$, $k_s$). Thus, it is important to conduct parametric study to examine the effect of input value variation on changes in the intrinsic kinetic parameters. In the present work, several commonly used models to determine $m$ and $D_{n\text{BAL},aq}$ were introduced. To determine $m$, two models (NRTL and UNIQUAC) were considered while three models (Wilke-Chang, Tyn-Calus and Hayduk-Minhas) for $D_{n\text{BAL},aq}$. In addition, a range of temperature dependence of salting out parameter, $k_s$ was examined to adjust the empirical parameter for nBAL, $i_{n\text{BAL}}$. Finally, changes of the intrinsic kinetic parameters and exponent of NaOH concentration were calculated from different estimates.

The NRTL and UNIQUAC are the most widely used activity coefficient models to calculate phase equilibrium. As shown in Table 2.2, it was found that activation energies using $m$ by NRTL are slightly higher, as compared to those by UNIQUAC. The difference in the average activation energy values using the two models was determined to be ~0.63 kcal/mol, corresponding to ~5 % of the average values, indicating that the activation energy is little influenced by the activity model used for solubility estimation. On the other hand, the exponent was not changed by $m$ value according to Eq. (2.15).
Diffusivity is another important factor to influence the value of activation energy. In the present work, three models [71-73] were introduced to estimate diffusivity of nBAL in the aqueous phase. It was observed that the values using Wilke-Chang and Tyn-Calus are quite close to each other, while higher than those using Hayduk-Minhas. This is because temperature dependence of Hayduk-Minhas model is higher than for the others. Since the results using Tyn-Calus and Wilke-Chang are essentially the same, results using the latter are not presented. It was found that the activation energies using $D_{nBAL,aq}$ by Tyn-Calus model are slightly higher than those by Hayduk-Minhas model, as shown in Table 2.2. The difference of activation energies between the two models was 0.47 kcal/mol, corresponding to ~3.5 % of the activation energy values. Similarly to the solubility case, the exponent was not influenced by the diffusivity values.

The salting-out parameter, $k_s$, influences solubility as seen in Eq. (2.9), however, unlike solubility or diffusivity, the salt effect can influence the exponent of NaOH concentration while the activation energy remains unchanged. As described in section 2.2.2, the empirical parameter for nBAL in Eq. (2.12) cannot be found from prior works. For this reason, $i_{nBAL}$ was evaluated by applying a proportional relation with the molar volume for the value of n-butylformate, which has structure similar to nBAL. Table 2.3 clearly shows that both n-butylformate (0.0134) and the evaluated (0.0098) cases give similar values of exponent (0.99 and 0.98, respectively), which are both close to 1. In Table 2.3, the two other values of the empirical parameter (0.013, n-butylformate; 0.054, benzene) from Santiago and Bidner (1971) [68] were evaluated as the alternative values for nBAL. As shown in Table 3, when the salting-out effect by nBAL was not considered ($i_{nBAL} = 0$), the average of the exponents for different temperatures was ~0.92. Further increase of the
empirical parameter results in increase of the exponent. Thus, the exponent for n-butylformate was calculated to be lower than that for benzene. However, it is likely that the value of the empirical parameter for benzene is not adequate to substitute that of nB(AL) since the molecular properties and structure of benzene are quite different from those of nB(AL). Indeed, the value for benzene may be considered in this calculation as the upper limit of \( i_{nBAL} \). It is noteworthy that the exponent value \( (1.22) \) for \( i_{Benzene} \) was still calculated to be close to 1, suggesting that the effect of the empirical parameter \( (i_{nBAL}) \) on the exponent is not significant. These results also indicate that the assumption to adjust the empirical parameter of nB(AL) was appropriate and the exponent of NaOH can be considered to be 1.

Apart from the empirical parameter, temperature dependence of \( k_s \) must also be considered to better determine \( k_s \). Long and McDevit (1952) [63] reported that the temperature derivative of partial volume \( \left( \frac{dV^o}{dT} \right) \) in Eq. (2.13) for NaCl, which is considered to have similar properties to NaOH, is 0.06 m\(^3\)/kmol.K in the temperature range 25-50 °C and \( \left( \frac{dV^o}{dT} \right) \) decreases by ~60% with every 25 °C increase. From this prior work, \( \left( \frac{dV^o}{dT} \right) \) for NaOH was estimated to be 0.07 m\(^3\)/kmol.K at the temperature range 25-50 °C which is about 50% decrease in value with every 25 °C increase in temperature. In the present work, in addition to 50%, the salting-out parameter was estimated for decrease rates of \( \frac{dV^o}{dT} = 0 \) and 70%, which may provide a possible range of \( \left( \frac{dV^o}{dT} \right) \) variation. With this assumption, \( k_s \) varies from 0.11 to 0.15 m\(^3\)/kmol with increasing temperature from 110 to 150 °C. Finally, the exponents of NaOH concentration for 0, 50 and 70 % of the decrease
rate were determined to be 0.85, 0.98 and 1.10, respectively, indicating that the exponent is not much influenced by the temperature dependence of $k_s$.

2.4.5 Comparison with previously published values

As described in section 2.1, most kinetic studies for nBAL aldol condensation were conducted at relatively low temperature (20-50 °C) and low nBAL concentration, to avoid mass transfer limitation and introduced solvent for homogeneous mixing with nBAL, resulting in 2nd order reaction [54-56]. On the other hand, Beltrame et al. (1973) [57] measured the kinetics in reaction controlled regime without solvent and found 1st order kinetics. They reported activation energy 20.9 kcal/mol, which is higher than our measured value (13.5 kcal/mol) for the range 110-150 °C. However, the reported activation energy 20.9 kcal/mol was obtained at 25~40 °C and some limitations were found such as the effect of temperature on nBAL solubility was ignored, and too low nBAL solubility (3.5 wt% at 20 °C) was used, which is significantly lower than other reported values (7.4 wt% [66]; 9.3 wt% [76]). In the present work, the effect of temperature on nBAL solubility was considered and furthermore, sensitivity studies using various published phase equilibrium models were conducted to better determine the intrinsic kinetic parameters ($k'$, $E_a$) as well as the exponent of NaOH concentration under industrial operating conditions (110-150 °C). Our results suggest that, although obtained under mass transfer regime, our determined activation energy value for nBAL aldol condensation is reasonable and robust in the industrial operation condition under possible changes of variables ($m$, $D_{nBAL,aq}$).
2.4.6 Kinetic study in the reaction controlled regime

To further verify the kinetic parameters obtained in the mass transfer regime, some experiments were carried out in the reaction controlled regime. For better temperature control, a smaller reactor size (100 ml) was employed and the experiments were conducted at lower temperature (25-40 °C), but higher mixing speed (~1500 rpm) to avoid mass transfer effect, as compared to the experiments in the mass transfer regime. After the reaction, the composition of the organic phase was analyzed by HPLC using the derivatization method described in section 2.2.2.

Unfortunately, proper kinetics were not obtained at the lower temperature range due to increased sample treatment time and side reactions. For the reaction at lower temperature, it required more time (5-20 min), to separate organic and aqueous phases from the products for analysis, during which further reaction occurred. In addition, as compared to the reaction at higher temperature, relatively higher amounts of heavies as by-product were produced instead of 2EHEL, even though similar conversion of nBAL was obtained.
Figure 2.9: Comparison of liquid chromatographic profiles at 30 °C and 120 °C, Peaks: 1, 2 = major impurities; 3 = nBAL; 4 = 2EHEL; 5 = hexanal; 6 = 2, 4-dinitrophenylhydrazine (used in derivatization method).

Figure 2.9 shows the profiles of HPLC spectra for the products at 30 and 120 °C. The ratio of heavies (peaks 1 and 2 in Figure 2.9) to 2EHEL (peaks 4 in Figure 2.9) at 30 °C was much larger than that at 120 °C. This means that lower temperature can help to obtain the reaction controlled regime but side reactions cannot be suppressed. In this case, it may be seen that aldol condensation of nBAL is not dominant and is affected by side reactions, which suggests difficulty to obtain intrinsic kinetics of the target reaction.
In the previous work [57], it was noted that product impurities were in trace amount, which does not agree with our observations. It is possible that their analytical procedure, involving high injector temperature (150 °C) for GC, modified the product distribution obtained after reaction. In addition, their GC column was relatively short (2 m, as compared to 50 m in our case), which may have hindered impurity detection. In support of our findings, Nielsen and Houlihan (2004) have also noted that yield of by-products can be up to 50 % over the temperature range 25-110 °C, which is lower than the industrially relevant 110-150 °C range investigated in our work.

2.4.7 Analysis of nBAL concentration in aqueous phase film and organic phase films

To better understand the penetration model employed in this study, the concentration gradient of nBAL in the aqueous phase film was calculated analytically using Eq. (2.6), along with the kinetic parameters obtained in the present work. The film thickness, calculated using $\delta = D/k_L$, is shown in Table 2.4, and gradients of concentration are plotted in Figure 2.10.
Table 2.4: Calculations of nBAL concentration in the film of aqueous phase and the comparison of reaction rate and mass transfer rate for 2EHEL in the organic phase

<table>
<thead>
<tr>
<th>Temperature</th>
<th>°C</th>
<th>110</th>
<th>150</th>
<th>110</th>
<th>150</th>
</tr>
</thead>
<tbody>
<tr>
<td>C_{NaOH} kmol/m³</td>
<td></td>
<td>0.76</td>
<td>0.76</td>
<td>1.9</td>
<td>1.9</td>
</tr>
<tr>
<td>Film thickness, δ mm</td>
<td></td>
<td>0.211</td>
<td>0.215</td>
<td>0.211</td>
<td>0.215</td>
</tr>
<tr>
<td>C_{nBAL} at x=δ kmol/m³</td>
<td></td>
<td>2.28E-03</td>
<td>1.91E-05</td>
<td>3.12E-05</td>
<td>9.13E-09</td>
</tr>
<tr>
<td>C_{nBAL} at x=δ/C_{nBAL} at x=0</td>
<td></td>
<td>0.606%</td>
<td>0.006%</td>
<td>0.01205%</td>
<td>0.000004%</td>
</tr>
<tr>
<td>R_{film,2EHEL} kmol/s</td>
<td></td>
<td>7.83E-08</td>
<td>1.79E-07</td>
<td>7.63E-08</td>
<td>1.76E-07</td>
</tr>
<tr>
<td>R_{org,2EHEL} kmol/s</td>
<td></td>
<td>6.80E-07</td>
<td>9.18E-07</td>
<td>6.88E-07</td>
<td>9.28E-07</td>
</tr>
<tr>
<td>R_{org,2EHEL}/R_{film,2EHEL}</td>
<td></td>
<td>8.69</td>
<td>5.14</td>
<td>9.01</td>
<td>5.26</td>
</tr>
<tr>
<td>Hatta number, Ha</td>
<td></td>
<td>5.11</td>
<td>9.66</td>
<td>9.02</td>
<td>16.94</td>
</tr>
</tbody>
</table>

Figure 2.10: nBAL concentration distribution in the film of the aqueous phase at various conditions.
It may be seen that the concentration of nBAL at $x = \delta$ is much lower than at $x = 0$, which satisfies the assumptions of the penetration theory (section 2.2.1). Even for the case of $110^\circ C$ and 0.76 M NaOH, where lowest $Ha$ (5.11) was evaluated among the operating conditions, the ratio $\frac{C_{nBAL}(x=\delta)}{C_{nBAL}(x=0)}$ was less than 1%. Thus all our results involving mass transfer dominated by reaction kinetics in the aqueous phase film satisfy the criterion $Ha > 3$, as described by Nanda and Sharma (1966) [59].

Although the solubility of 2EH EL in the aqueous phase is sufficiently low to be ignored, mass transfer of 2EH EL is considered since, if it is produced sufficiently fast, there may be mass transfer limitation in the film of organic phase. The generation rate of 2EH EL in the aqueous phase film is given by

$$R_{film,2EHEL} = \frac{1}{2} A_{surf} m C_{nBAL,org} \left( D_{nBAL,aq} k' e^{-\frac{E_a}{RT} C_{NaOH}} \right)^{0.5},$$

which is half of the reaction rate of nBAL, according to the stoichiometry of Eqs. (2.1) and (2.2). On the other hand, the mass transfer rate of 2EHEL into the organic phase is

$$R_{org,2EHEL} = A_{surf} k_{L,org} \left( \frac{1}{C_{2EHEL,org}} - C_{2EHEL,org} \right)$$

To calculate mass transfer coefficient of 2EHEL in the organic phase ($k_{L,org}$), the same model as the aqueous phase (Calderbank and Moo-Young, 1961 [74]; see Appendix B) was utilized since there is no appropriate correlation to estimate the mass transfer coefficient for thin plate shaped organic phase (Diameter/Length > 10) as in our system. In this manner, the mass transfer rates $R_{org,2EHEL}$ were 9 and 5 times faster than $R_{film,2EHEL}$ for 110 and 150 °C (Table 2.4), respectively, suggesting the absence of mass transfer limitation in the organic phase. This feature was also experimentally confirmed by
observing the plateau region where the mass flux of nBAL through the interface between organic and aqueous phases was not influenced by the agitator speed.

2.5 Concluding Remarks

The intrinsic reaction kinetics for biphasic aldol condensation of nBAL in NaOH solution was determined in the mass transfer regime using a stirred cell. Accounting for the solubility and diffusivity of nBAL in aqueous solution of NaOH, the intrinsic reaction kinetic parameters were obtained successfully under the industrial operating conditions, involving mass transfer regime dominated by reaction kinetics, confirmed by the plateau region test in several cases. The reaction rate of nBAL aldol condensation is given by

\[ r_{nBAL} = 1.712 \times 10^8 e^{-\frac{13.47 \pm 0.39 \text{ kcal/mol}}{RT}} C_{NaOH} C_{nBAL} \text{ (kmol/m}^3 \cdot \text{s)} \]

The unity reaction order for nBAL was confirmed by the experimental data. The activation energy values were obtained for four different NaOH concentrations and found to be essentially the same. Also, the exponent of NaOH concentration used as catalyst was verified to follow the reaction mechanism and the basic form of the reaction kinetics, Eq. (2.4). Through sensitivity studies, the important uncertainties in property values using different estimation methods were evaluated and found to be relatively small in magnitude.

The principal limitation of this study is the uncertainty of the estimated physical properties. It is not possible to independently confirm the extrapolated properties which were used in this study, although the resulting intrinsic kinetics is reasonable. Further, the
stirred cell also has the inherent uncertainty whether or not the interfacial area is sufficiently stable, although it was confirmed that its variation is acceptably small.

In summary, for cases where it is difficult to carry out the kinetic study in the reaction controlled regime due to fast reaction, and if lower temperature or dilution applied to overcome this difficulty lead to other unexpected problems as in this work, determining the intrinsic reaction kinetics using a stirred cell in the reaction dominated mass transfer regime is a good alternative method.

CHAPTER 3. ALDOL CONDENSATION OF N-BUTYRALDEHYDE IN A BIPHASIC STIRRED TANK REACTOR: EXPERIMENTS AND MODELS

3.1 Introduction

Biphasic (liquid-liquid) reaction processes provide several advantages over other types, including low pressure drop, intimate catalyst–substrate mixing, high yield and selectivity, and ease of catalyst recycling [77, 78]. For these reasons, biphasic reactions such as olefin alkylation, aromatic nitration and alkaline hydrolysis are widely used in the refinery, petro-chemical, fine chemical and pharmaceutical industries [5, 79, 80]. Despite the prevalence of biphasic reactions in industry, little fundamental work has been performed to clarify coupled reaction-transport events in biphasic systems [81]. Investigation of these issues and development of reactor models which account for hydrodynamic and transport effects is essential for design and scale-up of biphasic reactors.

Successful reactor modeling requires the knowledge of intrinsic reaction kinetics and its coupling with transport phenomena [82]. A biphasic reactor contains two immiscible liquids with liquid phase catalysts. One phase, primarily organic phase, includes the reactant feed and the other phase, mainly aqueous phase, includes catalysts such as acid-base or organo-metallic catalysts [9]. Based on the reaction rate, there are three different regimes to understand the phenomena in the aqueous phase, where the reaction occurs [15, 83]. The Hatta number (Ha), which is the ratio of the reaction rate in the liquid film
to the diffusion rate through the film [84], is used as a criterion to characterize the regimes. If the reaction is fast ($Ha > 3$), all reactions occur in the film region of the aqueous phase and none in the bulk region; this is called the mass transfer controlled regime dominated by the reaction rate in the film [85]. Under these conditions, the Danckwerts penetration model [12] is preferred to estimate the reaction rate in the film [86]. For intermediate rates, reactions occur in both the film and the bulk regions because the reactant concentration in the bulk region is non-zero. In this regime, Quadros et al. [83] used the film model for benzene nitration in a continuous-flow stirred tank reactor. The penetration model may not be appropriate in this case due to the difficulty in determining the bulk phase reactant concentration. In the third regime where the reaction is slow, there is no film resistance and the overall reaction rate is determined only by its intrinsic kinetics; this is called the reaction controlled regime [87]. In the cases of fast and intermediate reaction rates, knowledge of both interfacial reaction kinetics and interfacial area are required to understand the reactor behavior. However, determining the models and parameters describing these phenomena is challenging due to mass transfer disguised kinetics and difficulties in measurement and estimation of the interfacial area.

To evaluate these effects, a suitable reaction must be considered. Base-catalyzed biphasic aldol condensation of normal-butyraldehyde (nBAL) is conducted at the industrial operating conditions of 80-140 °C, ~5 bar, with 2-4 wt% of sodium hydroxide (NaOH) as catalyst [88]. This reaction is a part of the oxo-alcohol process to produce 2-ethyl-1-hexanol (2EH) which is commonly used to manufacture plasticizers for polyvinylchloride (PVC) [88]. The reaction mechanism of aldol condensation has been well studied as three
steps for aldol reaction and one step for condensation of water, leading to the overall reaction expressed as below [89].

\[ 2C_3H_7CHO \xrightarrow{k} C_7H_{13}CHO + H_2O \]  

(3.1)

For an accurate reactor model, it is necessary to obtain intrinsic reaction kinetics under industrially relevant conditions. Applying the pseudo steady state approximation with the 1st step (enolate formation) as the rate-determining [90], the kinetics is simplified as 1st order. Although the reaction at the industrial operating conditions is in the mass transfer controlled regime, the intrinsic reaction kinetic parameters were carefully determined using a stirred cell and the use of penetration theory in our prior work [85].

\[ \tau_{nBAL} = kC_{nBAL}, \quad (k = 1.712 \times 10^8 e^{-\frac{13.47 \times 10.39 \text{kcal/mol}}{RT}} C_{NaOH}) \]  

(3.2)

With these kinetic parameters, the range of the Hatta number for this reaction at the industrial operating conditions is 0.3-1.5, which is in the intermediate reaction regime. In this regime, to develop a reactor model, equations to estimate the interfacial area must also be available. For this purpose, Sauter mean diameter for dispersed phase droplet is widely used as it can be directly converted to the interfacial area through \( a = 6\varepsilon_d/d_{32} \). Semi-empirical correlations based on the critical Weber number and the isotropic turbulence theory by Kolmogoroff have been developed to obtain the Sauter mean
diameter [91]. Typically, correlations composed of the Weber number and the turbulence damping factor are preferred [23]. Although this method is simple and easy to use, the parameters of the correlations have to be determined individually for different cases of reactor geometry and chemical properties [28].

Experimental measurements of the interfacial area are needed to develop and verify correlations. The chemical method has been used to model a biphasic reactor [86]. In this method, the global interfacial area is obtained from specific chemical reaction systems, such as butylformate or diisobutylene, using the Danckwerts penetration theory in a stirred cell or a stirred tank [16, 62]. This method, however, has limitations because it does not directly measure the interfacial area [17]. Also, due to the restriction of this method to $Ha > 3$, it cannot be used for the present cases of nBAL aldol condensation at industrial operating conditions where $0.3 < Ha < 1.5$. As an alternative, the measurement of the interfacial area using a borescope is a direct and in-situ method [18]. In this technique, actual droplet pictures in a biphasic stirred tank are taken by a CCD camera with a borescope. The drop size distribution and interfacial area are obtained directly from counting droplets in the pictures [92]. This method, despite labor-intensive droplet counting, is acknowledged to be the most accurate method to obtain the interfacial area [18, 21]. Because the aldol reaction is at 80-140 °C and temperature can influence the interfacial area, the temperature effect has to be considered. While there are many reports using various methods to determine the interfacial area in a biphasic stirred tank, the relatively high temperature and NaOH concentration (1.14-1.9M) effects on interfacial area have not been investigated. In this light, the objectives of the present study are:
• Evaluate the effect of temperature and NaOH concentration on interfacial area, based on experiments using a borescope system and modeling using a semi-empirical correlation

• Develop a film-based reactor model for the biphasic stirred tank reactor with nBAL aldol condensation

• Determine the effect of the interfacial area modeling on the accuracy of the biphasic reactor model.

3.2 Theory

3.2.1 Reactor modeling using the film model

In a biphasic stirred tank reactor for nBAL aldol condensation, the organic phase is composed of nBAL reactant and 2-ethyl-2-hexenal (2EHEL) product, while the NaOH catalyst is in the aqueous phase. The NaOH concentration is considered constant in the aqueous phase with the assumption that the dilution by water as a reaction byproduct can be ignored. nBAL and 2EHEL are slightly soluble and nearly immiscible in the aqueous phase, while NaOH is immiscible in the organic phase. Thus, the reaction occurs in the aqueous phase, shown schematically by the two film theory in Figure 3.1.
Figure 3.1: Schematic diagram of the biphasic interface.

The film thickness $\delta$ is determined by $D/k_L$. The ultimate goal of the modeling is to estimate the nBAL concentration in the organic phase with time, to indicate reaction progress.

As a batch operation, the mass balance of nBAL in the organic phase is described by Eq. 3.3.

$$V_{org} \frac{dc_{nBAL,org}}{dt} = -R_{overall}; \ C_{nBAL,org} = C_{nBAL,org}^0 \ at \ t = 0$$

(3.3)

In the case of intermediate reaction regime, the overall reaction rate of nBAL is determined by mass transfer in the organic film, mass transfer with reaction in the aqueous film and reaction in the bulk aqueous phase [82]. Following the two film
resistance theory, the overall reaction rate can be described by the molar flux of nBAL at the interface.

\[ R_{overall} = aV_L n_{BAL}\bigg|_{x=0} \]  

(3.4)

The film model assumes a stagnant film, which means no change with time. The nBAL aldol condensation has been confirmed as a pseudo-first order reaction in prior work [85, 90]. Thus, the mass balance of nBAL concentration in the film of the aqueous phase is described by Eq. 3.5.

\[
D \frac{d^2 c_{nBAL, film}}{dx^2} - k c_{nBAL, film} = 0 \begin{cases} 
C_{nBAL, film} = C^*_n_{BAL, aq} \text{ at } x = 0 \\
C_{nBAL, film} = C_{nBAL, bulk} \text{ at } x = \delta \\
C^*_n_{BAL, aq} = m C^*_n_{BAL, org}
\end{cases}
\]  

(3.5)

In the present work, the distribution coefficient, diffusivity, and mass transfer coefficient for the continuous (aqueous) phase, including viscosity and density variations, were obtained from our previous study [85]. From the solution of Eq. 3.5, the nBAL concentration in the aqueous film is:

\[
c_{nBAL, film} = C^*_n_{BAL, aq} \cosh \sqrt{\frac{k}{D}} x + \frac{1}{\sinh H a} (C_{nBAL, bulk} - C^*_n_{BAL, aq} \cosh H a) \sinh \sqrt{\frac{k}{D}} x,
\]  

(3.6)

which is valid for \(0 \leq x \leq \delta\). The molar flux of nBAL in the film is given by

\[
J_{nBAL} = -D \left( \frac{dC_{nBAL, film}}{dx} \right) = \frac{k L H a}{\sinh H a} \left( C^*_n_{BAL, aq} \cosh \left( H a - \sqrt{\frac{k}{D}} x \right) - C_{nBAL, bulk} \cosh \sqrt{\frac{k}{D}} x \right)
\]  

(3.7)

To obtain \(C_{nBAL, bulk}\), the mass balance of \(C_{nBAL}\) in the bulk region of the aqueous phase is described by Eq. 3.8.
\[
\varepsilon_{\text{bulk}} V_t \frac{d C_{\text{nBAL, bulk}}}{dt} = a V_t f_{\text{nBAL}}|_{x=\delta} - \varepsilon_{\text{bulk}} V_t k C_{\text{nBAL, bulk}} \left\{ \begin{array}{l}
C_{\text{nBAL, bulk}} = 0 \text{ at } t = 0 \\
\varepsilon_{\text{bulk}} = \varepsilon_{\text{aq}} - a \delta
\end{array} \right\} \quad (3.8)
\]

From Eqs. 3.7, 3.8, we have

\[
C_{\text{nBAL, bulk}} = \frac{\alpha_1}{\alpha_2} (1 - e^{-\alpha_2 t}) \approx \frac{\alpha_1}{\alpha_2} (kt \gg 1) \quad (3.9)
\]

where \( \alpha_1 = \frac{a H_a k_L}{\varepsilon_{\text{bulk}} \sinh H_a} C_{\text{nBAL, aq}}^* \), \( \alpha_2 = \frac{a H_a k_L}{\varepsilon_{\text{bulk}} \tanh H_a} + k \).

The assumption of Eq. 3.9 \((kt \gg 1)\) is satisfied within 1 second in the actual calculations for nBAL aldol condensation under the reaction conditions. Therefore, with Eqs. 3.7, 3.9, \( R_{\text{overall}} \) is given by:

\[
R_{\text{overall}} = a V_t f_{\text{nBAL}}|_{x=0} = a V_t m C_{\text{nBAL, org}}^* k_L \left( 1 - \frac{1}{\cosh^2 H_a + \alpha_3 k} \right) \frac{H_a}{\tanh H_a}, \quad (3.10)
\]

where \( \alpha_3 = \frac{\varepsilon_{\text{bulk}} \sinh H_a \cosh H_a}{a H_a k_L} \).

The mass transfer rate in the organic phase film is simply described by Eq. 3.11, because no reaction occurs in the organic phase. The mass transfer coefficient for the organic phase as the dispersed phase is presented in Appendix D.

\[
R_{\text{overall}} = a V_t k_L \text{org} (C_{\text{nBAL, org}} - C_{\text{nBAL, org}}^*) \quad (3.11)
\]

Eliminating \( C_{\text{nBAL, org}}^* \) from Eqs. 3.10 and 3.11, the overall reaction rate is rearranged as a function of the nBAL concentration in the bulk organic phase and the operating conditions of temperature, NaOH concentration and the agitation power, as shown in Eq. 3.12.

\[
R_{\text{overall}} = k'_{\text{overall}} C_{\text{nBAL, org}} \quad (3.12)
\]

where
In fact, we can view the two terms on the r.h.s. of Eq. 3.13 as the resistances corresponding to the aqueous and organic phases, respectively to yield

\[
\frac{1}{R_{\text{overall}}} = \frac{1}{R_{\text{aq}}} + \frac{1}{R_{\text{org}}} \tag{3.14}
\]

which is the typical way to note that the overall resistance equals the sum of the resistances in series.

Substituting Eq. 3.12 in Eq. 3.3, the latter becomes a simple first-order linear ordinary differential equation for the nBAL concentration in the organic phase, which has the solution:

\[
\ln C_{n\text{BAL,org}} = -k'_{\text{overall}} t + \ln C_{n\text{BAL,org}}^0 = -k_{\text{overall}} t + \ln C_{n\text{BAL,org}}^0 \tag{3.15}
\]

Using Eq. 3.15, it is possible to determine the change of nBAL concentration in the organic phase, with time. However, this evaluation still requires information for the interfacial area \(a\) as a function of agitation power, reactor geometry and hydrodynamic properties including density, viscosity and interfacial tension.

### 3.2.2 Interfacial area estimation

The dispersed phase in a biphasic stirred tank is composed of several different sized droplets. The Sauter mean diameter, obtained from the number of droplets \(n_i\) with diameter \(d_i\), is useful to determine the interfacial area, representing the droplet distribution of the dispersed phase. The Sauter mean diameter is defined as:
\[ d_{32} = \frac{6\varepsilon_d}{a} = \frac{\sum_i n_i d_i^3}{\sum_i n_i d_i^2} \] (3.16)

If the biphasic system is in turbulence so that drop diameters are much larger than Kolmogoroff’s length scale and the drop is inviscid, by combining the critical Weber number for breakup of droplets by Hinze [24] and Kolmogoroff’s theory of isotropic turbulence [93], the maximum drop diameter \( (d_{max}) \) is described by Eq. 3.17 [23]:

\[ \frac{d_{max}}{d_{imp}} = C_1 We^{-0.6} \] (3.17)

The maximum diameter is substituted with the Sauter mean diameter following several reports that they are linearly proportional [23, 94], although some researchers [91, 95] disagree. Doulah [25] extended this equation to consider that the Sauter mean diameter increases with the holdup of the dispersed phase due to the coalescence and damping of turbulence.

\[ \frac{d_{32}}{d_{imp}} = C_1 (1 + C_2\varepsilon_{org}) We^{-0.6} \] (3.18)

\( C_1 \) is an empirical parameter which includes the effects of reactor geometry, power number and several proportional factors, while \( C_2 \) represents the turbulence damping or coalescence factor. The reported ranges of \( C_1 \) and \( C_2 \) values are 0.04-0.4 and 2-10, respectively [60]. Eq.3.18 is the most frequently used form to estimate the Sauter mean diameter. Further, a viscosity factor has also been introduced to include the effect of different chemical systems, as described by Eq. 3.19 [26, 96].

\[ \frac{d_{32}}{d_{imp}} = C_1 (1 + C_2\varepsilon_{org}) We^{-0.6} \left( \frac{\mu_d}{\mu_c} \right)^{C_3} \] (3.19)
$C_3$ is an exponent for the viscosity factor, which has reported values up to 0.44 for different chemical systems [26]. Models for the interfacial tension used in this study are described in Appendix E.

3.3 Interfacial area

The interfacial area in the biphasic stirred tank reactor for nBAL aldol condensation was evaluated through the droplet pictures taken using a borescope system. A new correlation for the interfacial area was developed after assessing the predictions of literature correlations in comparison with the experimental measurements from this work.

3.3.1 Experimental apparatus

The measurements of the interfacial area were conducted in a 300 ml stainless steel reactor, with height to diameter ratio 1.6. A four blade and 45°-pitched paddle was used to agitate the biphasic solution. The diameter ratio of the impeller to the reactor was 0.5. The visualization system installed in the reactor consisted of a borescope with a sight tube, a CCD camera, a fiber optic strobe and a personal computer as shown in Figure 3.2.
Figure 3.2: Schematic diagram of the biphasic stirred tank with the borescope system; 1- stirred tank, 2 – CCD camera, 3 – Personal computer, 4 – fiber optic strobe.

The borescope had a small diameter (Ø = 2.4 mm; Gradient Lens Corporation Pro Superslim) to prevent the disruption of drop size distribution when installed in the reactor. The sight tube, using a 1/8” or 3/16” stainless tube and a borosilicate sight glass, was used to protect the borescope and help to focus the image. A two million pixel CCD camera (Sony XCDU100) and a fiber optic strobe (Excelitas X-1500) were used to take pictures of droplets which were sent to the personal computer for storage.

3.3.2 Calibration

Although the borescope system provided the pictures of actual droplets, the focus distance between the lens of the borescope and the actual droplets required calibration of the system. For this, two sets of glass beads were used (20-30 mesh and 250 µm). For reference, the particle size distribution was measured by a laser scattering particle size
analyzer as the standard (HORIBA LA-950). The particle size distributions, suspended under water, were then measured by manually counting 200-400 glass beads in many pictures obtained using the borescope system. The results from the two different methods were compared, as shown in Figure 3.3, where it may be seen that the particle size distributions from the borescope were somewhat broader than the actual size. This may be caused by counting particles which are slightly out of focus and appear smaller when they are far away from the sight glass. Despite this limitation [19, 92], the borescope system provides a good estimate of the particle size distribution, hence is suitable for use in the current study.

Figure 3.3: The calibration of the borescope system with glass beads (1: 250 μm glass beads, 2: 20-30 mesh glass beads).
3.3.3 Experimental procedure

To measure the droplet size distribution, unreactive 2EHEL (2-ethylhexenal) was selected as the organic phase because nBAL reacts in the aqueous NaOH solution under the experimental conditions. After each experiment, GC (gas chromatography) analysis conducted after taking pictures confirmed that 2EHEL did not react in the NaOH solution. In fact, 2EHEL was prepared by aldol condensation of nBAL and stored in a glass jar (2EHEL purity > 95%). For the droplet size experiments, a mixture of 140 ml NaOH solution and 60 ml 2EHEL was placed in the reactor, pressurized with nitrogen up to 1.5 barg to prevent azeotropic boiling. After a minimum of 10 min. for each experimental condition, 20-40 pictures of droplets were taken, with most drops appearing perfect spheres as shown in Figure 3.4.

![Figure 3.4: Drop pictures by the borescope system; a) 110 °C, 1.9M C_{NaOH}, 800 rpm; b) 25 °C, 0M C_{NaOH}, 600 rpm.](image)
The pictures were taken from the top and bottom positions of the reactor to obtain average values for the entire vessel. A minimum of 200 drops for each position, totaling 400 drops, were counted for each experimental condition. In general, high rpm produced smaller droplets and their size at the top position was somewhat larger than at the bottom, as shown in Figure 3.5.

![Droplet size distribution graph](image)

**Figure 3.5:** Accumulated droplet size distributions at 110°C, 1.9M C<sub>NaOH</sub>; T-top, B-bottom; the numbers in the legend denote agitator rpm.

This may be caused by droplet breakup near the agitator and coalescence at the stator zone far from the agitator, which conforms to literature reports for biphasic stirred tanks [97]. The possible disturbing effect of the borescope was checked by CFD (computational fluid dynamics) technique, as described in Appendix F. Because the
borescope caused less than 1% difference in the P/V (power per volume) of the reactor, it was shown that the effect was negligible. Thus, the preliminary tests demonstrated that the borescope system was well suited for the droplet size distribution study to provide the interfacial area.

3.3.4 Measurements of the Sauter mean diameters

The Sauter mean diameters ($d_{32}$) were measured in the range 25-110 °C, 0-1.9 M $C_{\text{NaOH}}$ and 600-1000 rpm. The borescope system allowed measurements of the drop diameters below 110 °C while most prior studies reported in the literature were conducted near 25 °C. Unfortunately, due to the upper temperature limit of the borescope, it was not possible to measure droplets in the higher 120-140 °C region of the industrial operating condition. For all experiments, the ratio of the dispersed phase was maintained at 0.3, similar to the experimental conditions during reaction.
As shown in Figure 3.6, all three variables of rpm, temperature and NaOH concentration influence \(d_{32}\) measurements. As expected, the effect of agitation power (rpm) was much
larger than the others. Interestingly, however, the effects of temperature and $C_{\text{NaOH}}$ were also significant. The Sauter mean diameters of $2EHEL$ in NaOH solution increased with temperature increase and decreased with $C_{\text{NaOH}}$ increase. The physical properties including viscosity, density and interfacial tension are affected by temperature and NaOH concentration, which consequently influence Sauter mean diameters. With temperature increase, the viscosity correction term discussed in the theory section increases, although viscosities for both the dispersed phase and the continuous phase decrease. This is because the viscosity for the continuous phase, using the correlation from Laliberte [98], decreases more significantly than for the dispersed phase due to the effect of temperature as shown in Table 3.1.

<table>
<thead>
<tr>
<th>$T$ (°C)</th>
<th>$\mu_d$</th>
<th>$\mu_c$</th>
<th>$\mu_d/\mu_c$</th>
<th>$\sigma$</th>
</tr>
</thead>
<tbody>
<tr>
<td>25</td>
<td>1.06</td>
<td>1.31</td>
<td>0.80</td>
<td>23.5</td>
</tr>
<tr>
<td>70</td>
<td>0.65</td>
<td>0.56</td>
<td>1.16</td>
<td>21.7</td>
</tr>
<tr>
<td>90</td>
<td>0.55</td>
<td>0.43</td>
<td>1.27</td>
<td>20.8</td>
</tr>
<tr>
<td>110</td>
<td>0.47</td>
<td>0.35</td>
<td>1.35</td>
<td>19.8</td>
</tr>
</tbody>
</table>

The increased viscosity correction term results in larger $d_{32}$. Regarding the Weber number term only, since the interfacial tension from Appendix E decreases slightly with temperature increase, the Sauter mean diameter should decrease slightly as well. However, the experimental result is the opposite (Figure 3.6c). This implies that with respect to the temperature change, the effect of the viscosity correction term is larger than that of the interfacial tension. Concerning the effect of increasing NaOH concentration,
the viscosity of the continuous phase increases and results in smaller drops while the interfacial tension slightly increases and results in larger drops, as shown in Table 3.2.

Table 3.2: Viscosities and interfacial tension for dispersed and the continuous phases at 70 °C

<table>
<thead>
<tr>
<th>C_{NaOH}</th>
<th>\mu_d</th>
<th>\mu_c</th>
<th>\mu_d/\mu_c</th>
<th>\sigma</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>0.65</td>
<td>0.40</td>
<td>1.61</td>
<td>20.6</td>
</tr>
<tr>
<td>1.14</td>
<td>0.65</td>
<td>0.47</td>
<td>1.39</td>
<td>21.3</td>
</tr>
<tr>
<td>1.9</td>
<td>0.65</td>
<td>0.56</td>
<td>1.16</td>
<td>21.7</td>
</tr>
</tbody>
</table>

Because the observed d_{32} decreases with increasing NaOH concentration (Figure 3.6c), d_{32} is affected more by the viscosity correction, rather than the interfacial tension, similar to the effect of temperature. Thus, the viscosity correction term is necessary to explain the effects of temperature and NaOH, and cannot be ignored.

3.3.5 Estimation of the Sauter mean diameter using correlations

In order to develop an accurate reactor model, it is necessary to build an estimation model for the Sauter mean diameter. In this study, with 3-7x10^4 impeller Reynolds number, the biphasic system was in turbulence. In addition, the measured d_{32} was much larger than the Kolmogoroff’s length scale \eta = 20-50\mu m, calculated with the averaged turbulent dissipation rate. Further, with 0.3-1.1 cP \mu_d, drops in this system may be regarded as inviscid. Under these conditions, there are several available correlations obtained from measurements with various chemical systems, measurement methods and operating conditions. Among these, correlations with 0.3 or similar fraction of the dispersed phase
were selected [23, 26, 60, 62, 96, 99-102] and evaluated with the measurement data of this study as shown in Figure 3.7 and Table 3.3.

Figure 3.7: Parity plot for the Sauter mean diameters estimated by published correlations; the dash lines represent ±15% errors.
Table 3.3: Summary of the correlations used in Figure 7

<table>
<thead>
<tr>
<th>Reference</th>
<th>Measurement Method</th>
<th>$\epsilon_{\text{org}}$</th>
<th>$d_{32}$ (mm)</th>
<th>$N$ (s$^{-1}$)</th>
<th>$R^2$ (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Godfrey et al., 1989</td>
<td>Photography</td>
<td>0.1-0.5</td>
<td>-0.3</td>
<td>7-9</td>
<td>61.7</td>
</tr>
<tr>
<td>Calderbank, 1958</td>
<td>Light transmittance</td>
<td>0-0.2</td>
<td>0.05-0.32</td>
<td></td>
<td>57</td>
</tr>
<tr>
<td>Santiago and Trambouze, 1971</td>
<td>Chemical</td>
<td></td>
<td></td>
<td></td>
<td>31.7</td>
</tr>
<tr>
<td>Quadros and Baptista, 2003</td>
<td>Chemical</td>
<td>0.061-0.166</td>
<td>0.01-0.2</td>
<td>5-25</td>
<td>31.9</td>
</tr>
<tr>
<td>Woezik and Westerterp, 2000</td>
<td>Chemical</td>
<td>-0.3</td>
<td>0.03-0.15</td>
<td>15-25</td>
<td>31.9</td>
</tr>
<tr>
<td>Brooks and Richmond, 1994</td>
<td>Photography</td>
<td>0.1-0.5</td>
<td>-0.4</td>
<td></td>
<td>31.7</td>
</tr>
<tr>
<td>Singh et al., 2008</td>
<td>Photography (offline)</td>
<td>0.2-0.5</td>
<td>0.4</td>
<td>1.7-2.5</td>
<td>36.1</td>
</tr>
<tr>
<td>Mlynek and Resnick, 1972</td>
<td>Photography (Insitu)</td>
<td>0.025-0.34</td>
<td>0.14-0.46</td>
<td>2.3-8.3</td>
<td>31.9</td>
</tr>
<tr>
<td>Desnoyer et al., 2003; slow coalescence</td>
<td>Laser granulometer</td>
<td>0.1-0.6</td>
<td>0.15-0.35</td>
<td>10.2-15</td>
<td>31.9</td>
</tr>
</tbody>
</table>
Remarkably, the role of the viscosity correction term was important based on the $R^2$ values from correlations with and without the viscosity correction term, which were ~60% and 30-36%, respectively, as discussed in the previous section. However, because even ~60% $R^2$ is not enough accuracy, and particularly to account for the effects of temperature and NaOH, a new correlation based on Eq. 3.19 was developed by fitting with the measurements of this study (Figure 3.6):

\[
d_{32} / d_{imp} = 0.207 * W_{c}^{-0.53} \left( \frac{\mu_d}{\mu_c} \right)^{0.94}
\]  

(3.20)

The exponent of the Weber number for the full data set was -0.53, which is slightly larger than -0.6 and may arise in relatively larger dispersed phase fraction systems due to the complex breakage mechanism, as studied by Desnoyer et al [102]. The exponent of the viscosity correction term was 0.94, which is larger than 0.25-0.44 in prior studies [26, 96] where the correction term was used solely for different chemical species.

Eq. 3.20 was compared with prior works by Godfrey et al. [26] and Santiago and Trambouze [99], which are representative of correlations with and without the viscosity correction term, respectively, as shown in Figure 3.8.
Figure 3.8: Comparison between the measurements (points) and the estimation (line) by a) Santiago and Trambouze (1971); b) adjusted Godfrey et al. (1989); c) this work.
The latter did not follow the trend with the temperature and NaOH concentration because it did not include a viscosity correction term, while the former showed a relatively better result owing to its viscosity correction term. The correlation developed in this work (Eq. 3.20) showed the best fit with the experimental data obtained from various temperatures and NaOH concentrations. The $R^2$ of this correlation was 89.5%, sufficiently accurate for the reactor modeling, as compared to 31-62% $R^2$ for the other correlations shown in Table 3.3. As shown in Figure 3.7, Eq. 3.20 is accurate within ±15% range and may be used for modeling the biphasic stirred tank reactor for nBAL aldol condensation.

### 3.4 Biphasic stirred tank reactor for nBAL aldol condensation

With the interfacial area investigated in the previous section, the biphasic stirred tank reactor for nBAL aldol condensation was studied both by experiments and modeling. The effect of the interfacial area estimation on predictions of the overall reactor model was evaluated.

#### 3.4.1 Experimental setup

The reaction experiments were carried out in a 300 ml stirred tank reactor system, shown in Figure 3.9. The reactor, including the agitator, was the same used in the previous section. The reactor had a thermocouple, a heating jacket for external heating and a cooling coil inside the reactor to control the reaction temperature. The injector was used to preheat nBAL before the reaction and to inject nBAL into the reactor through a nozzle to initiate the reaction. Heating tape with a temperature controller was used for preheating the injector, while pressurized nitrogen was used for injection and initial venting. To
stabilize the drop distribution of the organic phase as rapidly as possible, the injection nozzle was located near the agitator and modified to create a spray when nBAL was injected. For sampling, 1/8” OD stainless steel tube and 1/16” OD copper tube were used inside the reactor vessel and for the condenser, respectively. The copper tube was dipped inside an ice water bath condenser, to quench the samples. The total volume of sampling tubes was about 0.08 ml, which was about 4% of a 2 ml sample volume.

The organic phase samples were analyzed by GC (Gas Chromatography, HP 5890 II) equipped with an Agilent DB-WAXetr capillary column (50 m x 0.32 mm) and flame ionization detector under the following conditions: helium as carrier gas (2.6 ml/min),
inlet and detector temperatures 250 °C, oven temperature from 80 to 220 °C and injection volume 0.5 µl with split. Acetonitrile (Sigma-Aldrich, > 99.9%) was selected as the internal standard. Normal-butyraldehyde (nBAL, Sigma-Aldrich, > 99%) and 2-ethyl-2-hexenal (2EHEL, Sigma-Aldrich, > 93%) were used for calibration of the GC. The $R^2$ values of all GC calibrations were 99.6%.

The experiments started with preparation of NaOH solution. The NaOH solution (50% in water, Sigma-Aldrich) was diluted with deionized water to obtain specific concentrations, which were confirmed by pH meter (OMEGA PHB-209). The ratio of nBAL to NaOH solution was 0.3, similar to the industrial operating condition. 140 ml of NaOH solution and 60 ml of nBAL were prepared in the reactor vessel and injector, respectively. After preheating both to the target temperature, nBAL was injected into the reactor in less than 2 seconds by 200 psig Nitrogen. The reactor was operated at 140-200 psig pressure to prevent azeotropic boiling between nBAL and water during injection of nBAL and initiation of the reaction. The reactor temperature was controlled within ± 2 °C by a PID controller. Four liquid samples were taken after 10 or 20 seconds at 10 second intervals. The removed samples were stabilized for 0.5-1 minute to separate the organic and aqueous phases. The separated organic phase was collected using a 3 ml disposable syringe, and analyzed by GC. The repeatability error for each concentration point was less than 5%. The mass balance was checked after each experiment and the error was always less than 1%.
3.4.2 Preliminary studies

The solubilities of nBAL and 2EHEL in the NaOH solution may change when samples are transferred from the reactor at high temperature to a stabilized condition at room temperature. To evaluate this effect, the error between the two conditions was calculated with the UNIQUAC activity model [65]. The predicted error between them was less than 3.3% for the organic phase fraction 0.3, indicating that the effect was negligible.

From Eq. 3.12, the overall reaction order of nBAL should be 1st order, as described in the Theory section. This was confirmed by the linear relation of \( \ln C_{\text{nBAL,org}} \) with time according to Eq. 3.15 for a range of conditions, as shown in Figure 3.10.

![Figure 3.10: Confirmation of 1st order reaction through \( \ln C_{\text{nBAL}} \) as a function of time.](image)

1: 110 °C, 1.03 M C\(_{\text{NaOH}}\), 600 rpm; 2: 110 °C, 1.03 M C\(_{\text{NaOH}}\), 800 rpm; 3: 120 °C, 1.52 M C\(_{\text{NaOH}}\), 700 rpm; 4: 120 °C, 0.76 M C\(_{\text{NaOH}}\), 900 rpm; 5: 92 °C, 1.60 M C\(_{\text{NaOH}}\), 919 rpm
As noted previously, the industrial operating temperature for this reaction is 80-140 °C [88]. In our prior study, the intrinsic kinetics were obtained over 110-150 °C due to Hatta number restrictions [85]. Therefore, two design of experiment (DOEs) were prepared to verify the biphasic reactor model, as described by Eq. 3.12. One (DOE 1) was designed by the inscribed central composite method [103] over 80-140 °C, 0.05-2.0 M $C_{NaOH}$ and 600-1000 rpm to cover the full industrial operation range. The range of the other (DOE 2) was 110-130 °C, 0.76-1.52 M $C_{NaOH}$ and 700-900 rpm, selected as being in the narrow kinetic experimental range using the Box-Behnken design [103]. DOE 1 and 2 were composed of 20 and 15 cases, including 6 and 3 repeats of the central points, respectively.

### 3.4.3 Experimental results and evaluation of the reactor model

Experiments for the two DOE sets were carried out in the 300ml stirred tank reactor. Both sets were confirmed as statistically reliable using MINITAB with 2.8% and 3.4% repeatability error for the central points of DOE 1 and 2, respectively. The experiments where the total conversion ranges were less than 20% were excluded. These cases, when the reaction rate is too fast or too slow for the operating conditions of temperature, $C_{NaOH}$ or rpm, can lead to large experimental errors because the extracted data from experiments is the slope of concentrations from successive samples. Following this criterion, 27 points were used to evaluate the model, while 8 points were excluded. Most of the excluded points were obtained at greater than 80% nBAL conversion (below 2 M of $C_{nBAL}$ at 10 seconds) under high temperature and high rpm. The overall reaction constants ($k_{overall}$) were obtained from $C_{nBAL}$ values of each experiment by Eq. 3.15 and used to verify the
$k_{overall}$ values estimated from Eq. 3.13. The average differences in overall reaction rates between the experimental and predicted values were 10% and 4%, for DOE 1 and 2, respectively. All estimations for DOE 1 and 2 were in ±15% error range as shown in Figure 3.11a.

Figure 3.11: Parity plots of the overall rate constant - the experiments versus the estimations by the model using (a) the correlation developed in this work; (b) the correlations from literature; the dash lines represent ±15% errors.
The average estimation error for DOE 2 was lower than that for DOE 1 because DOE 2 is within the experimental conditions for the intrinsic reaction kinetics in our prior work [85]. The results of DOE1 demonstrate that the intrinsic kinetics from 110-150 °C can be extrapolated down to 80 °C and also that the film model, combining the intrinsic kinetics from the stirred cell and the interfacial area correlation from the borescope system without any adjustable parameters, is applicable for the entire industrial operating range.

In order to evaluate the effect of the interfacial area estimation on the accuracy of the reactor model, the predicted results were compared with estimations from different literature interfacial area correlations, using the same reactor model, as shown in Figure 3.11b. It may be seen that the interfacial area correlation has a significant impact on the estimation accuracy of the reactor model. It is seen once again that the interfacial area correlation developed in this work (Eq. 3.20) is more appropriate to describe reactor performance than prior correlations available in the literature.

Although the developed interfacial area correlation was optimized for the biphasic nBAL aldol condensation in the industrial operating range, the estimation error of interfacial areas was around ±15%, as described in the Interfacial Area section. Since the estimation error of the interfacial area directly affects the accuracy of the reaction rate estimation, the reaction rate predictions should be better with the reactor model combined with $d_{32}$ measurement data directly from the borescope system, not through the correlation. To confirm this, 11 additional reaction experiments were conducted under the same operating conditions along with 11 interfacial area measurements obtained with no reaction (see section 3.3.1). The range of the experiments was 90-110 °C, 1.14-1.9 M $C_{NaOH}$ and 600-1000 rpm.
Figure 3.12: Comparison of the estimation accuracy between the model using $d_{32}$ estimations and $d_{32}$ measurements; the dash lines represent $\pm 8\%$ errors.

As shown in Figure 3.12, the estimation error of the reactor model using the direct $d_{32}$ measurements was within $\pm 8\%$, which is significantly lower than the $\pm 15\%$ error range for DOE 1 and 2, as shown in Figure 3.11a, when the developed correlation was used. The predictions of the model using the correlation were compared with estimations of the reactor model using $d_{32}$ measurements, as shown in Table 3.4 where $R^2$ values were 97.3 % and 98.5 %, respectively. Alternatively, these can be expressed as 2.7% and 1.5% $R^2$ errors, respectively. The standard deviation of the rate prediction errors was 5.0 % and 3.3 % for reactor models with $d_{32}$ estimations and $d_{32}$ measurements, respectively. From both evaluations, it is apparent that the errors of the reactor model predictions using $d_{32}$ estimations were nearly double those using $d_{32}$ measurements. This shows that the effect
of the $d_{32}$ estimation error is significant on the biphasic reactor model for nBAL aldol condensation, suggesting that the modeling accuracy may be improved with more effective $d_{32}$ estimations.

Table 3.4: $k_{\text{overall}}$ from experiments, the models using $d_{32}$ estimations and $d_{32}$ measurements

<table>
<thead>
<tr>
<th>Temp °C</th>
<th>$C_{\text{NaOH}}$ M</th>
<th>RPM min$^{-1}$</th>
<th>$k_{\text{overall}}$ ($s^{-1}$)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td>Experiment</td>
</tr>
<tr>
<td>110</td>
<td>1.90</td>
<td>800</td>
<td>0.064</td>
</tr>
<tr>
<td>110</td>
<td>1.90</td>
<td>900</td>
<td>0.075</td>
</tr>
<tr>
<td>110</td>
<td>1.90</td>
<td>1000</td>
<td>0.090</td>
</tr>
<tr>
<td>110</td>
<td>1.90</td>
<td>700</td>
<td>0.053</td>
</tr>
<tr>
<td>110</td>
<td>1.90</td>
<td>600</td>
<td>0.042</td>
</tr>
<tr>
<td>110</td>
<td>1.90</td>
<td>700</td>
<td>0.053</td>
</tr>
<tr>
<td>110</td>
<td>1.14</td>
<td>800</td>
<td>0.068</td>
</tr>
<tr>
<td>90</td>
<td>1.14</td>
<td>800</td>
<td>0.047</td>
</tr>
<tr>
<td>90</td>
<td>1.90</td>
<td>800</td>
<td>0.048</td>
</tr>
<tr>
<td>90</td>
<td>1.90</td>
<td>1000</td>
<td>0.060</td>
</tr>
<tr>
<td>90</td>
<td>1.90</td>
<td>700</td>
<td>0.042</td>
</tr>
</tbody>
</table>

3.5 Concluding remarks

Experimental and modeling studies of nBAL aldol condensation in a 300ml biphasic stirred tank were conducted to estimate the overall reaction rate under industrial operating conditions in the intermediate reaction regime, accounting for the interfacial area, mass transfer in both phases, as well as the intrinsic reaction kinetics.

The interfacial area for reactor modeling was obtained using the experimental measurements and semi-empirical correlation for the Sauter mean diameter. The
borescope system as a physical and in-situ measurement technique was used to obtain the data under 25-110 °C, 0-1.9 M C\textsubscript{NaOH} and 600-1000 rpm at 0.3 of the organic phase ratio. The measurements showed that the effects of temperature and C\textsubscript{NaOH} on the interfacial area were significant in the biphasic system of 2EHEL and NaOH solution. In order to consider these, the semi-empirical correlation was modified with a viscosity correction term and its empirical parameters were determined using the measurement data. The predictions of the developed correlation were significantly better than those of other correlations available in the literature.

The biphasic reactor model was developed based on interfacial area and the intrinsic reaction kinetics. The model was successfully verified with the reaction experimental results through two different DOE sets at 80-140 °C, 0.05-2.0 M C\textsubscript{NaOH} and 600-1000 rpm. The comparison of the experimental results with the biphasic reactor model predictions directly using the interfacial area measurements showed that the model accuracy was significantly improved.

This study shows that the film model, considering film resistance in the organic phase and reactions in both the aqueous film and the bulk region of the aqueous phase, successfully estimates the biphasic stirred tank reactor performance for nBAL aldol condensation in the intermediate reaction regime. This work is expected to be helpful for design and optimization of biphasic reactors for other applications.

Note: Adapted with permission from American Institute of Chemical Engineers (S. Lee and A. Varma, “Aldol condensation of n-butyraldehyde in a biphasic stirred tank reactor:
CHAPTER 4. ACETOPHENONE HYDROGENATION ON RH/AL₂O₃ CATALYST: INTRINSIC REACTION KINETICS AND EFFECTS OF INTERNAL DIFFUSION

4.1 Introduction

Acetophenone (AP) is the simplest ketone having an aromatic ring and is produced from the industrial process for phenol synthesis from isopropylbenzene, selective decomposition of cumene hydroperoxide, or the oxidation of ethylbenzene [104]. Acetophenone is used as a raw material for synthetic resins by reaction with formaldehyde and is also frequently used for organic syntheses, especially for pharmaceutical applications [104, 105]. The hydrogenation of acetophenone provides industrially important products including 1-phenylethanol (PE) for pharmaceutical and fragrance industries and 1-cyclohexylethanol (CE) used in the manufacture of some polymers [105, 106]. The reaction is a typical liquid-phase hydrogenation on transition metal catalysts including rhodium, ruthenium, and copper [107-109]. However, The reaction scheme is also complex, with two competing hydrogenations of aromatic ring and carbonyl group occurring simultaneously with hydrogenolysis of PE [110]. Because the reaction is three-phase, complex, and has valuable products, it was selected as a suitable model reaction for studies of three-phase hydrogenations using slurry or trickle bed reactors for fine chemical and pharmaceutical applications [111, 112].

For model development and scale-up of the trickle bed reactor, it is necessary to first obtain the intrinsic reaction kinetics and couple with internal diffusion effects inside the
catalyst pellet [32]. The reaction kinetics following the Langmuir-Hinshelwood hypothesis (L-H kinetics) provide an understanding of the surface reaction on solid catalysts. However, few papers with L-H kinetics for the reaction have been published. Considering the valuable products, the desired reaction route was selected in this work as $\text{AP} \rightarrow \text{PE} \rightarrow \text{CE}$. For this reason, Ni and Cu catalysts, which prefer the routes $\text{AP} \rightarrow \text{PE} \rightarrow \text{EB}$ (ethylbenzene) and $\text{AP} \rightarrow \text{PE}$, respectively, were excluded [109, 113]. Kinetic studies using Pt and Rh have been reported, but Pt was not considered due to its limited and low-temperature range, which can negatively affect selectivity [110, 114]. Prior work using Rh/C has a more appropriate temperature range, L-H kinetics, and includes intraparticle diffusion modeling [115]. This work, however, assumes non-dissociative hydrogen adsorption within the L-H kinetics, obtains the shell depth of the eggshell type catalyst particle from data fitting rather than by measurements, and assumes an adsorption constant for PE. In the present work, following literature search, catalyst screening was conducted, intrinsic L-H kinetics were determined, and the effects of internal diffusion were studied.

### 4.2 Experimental

#### 4.2.1 Catalysts

The catalysts used in the study were rhodium, ruthenium, and platinum, supported on alumina and activated carbon powder, from Alfa Aesar (1% Rh/Al₂O₃, 11769; 5% Ru/Al₂O₃, 11749; 1% Pt/Al₂O₃, 11797; 5% Rh/C, 44863) and Evonik (5% Rh/C, G106 NW, G106 BW), with average particle diameters 9.8, 13 and 30 µm, respectively, for the
alumina supported catalysts. To evaluate the effects of internal diffusion, rhodium on alumina spheres of 0.5 and 2.5 mm diameters from Alfa Aesar were used (1% Rh/Al₂O₃: 0.5mm, S55366; 2.5mm, 47040). The mean particle diameters were measured by laser scattering particle size analyzer (HORIBA LA-950). More detailed catalyst characterization of 1% Rh/Al₂O₃ is described in Section 4.3.4.1.

4.2.2 Experimental setup

The reaction experiments were conducted in a semi-batch slurry reactor using a 300-mL Parr vessel (Parr 4843) with an injector, as shown in Figure 4.1.

Figure 4.1: Schematic diagram of the experimental setup. (1 - 300-ml stirred tank reactor, 2 - injector, PI - pressure indicator, TIC - temperature indicator / controller)
Mechanical agitation by a 4-blade pitched paddle was applied to ensure no external gas-liquid-solid mass transfer limitations inside the reactor. The reactor temperature was controlled within ±1 °C by a PID controller with an electric heating pot and a cooling tube inside the reactor. Hydrogen was supplied from a high-pressure cylinder with a pressure regulator, which controlled reactor pressure within ±0.5 bar. The hydrogen flow was approximately 0.3 NL/min (NL: normal liter, volume at 0 °C, 1 atm) controlled by the metering valve on the vent line. The injector was composed of a ¾-inch stainless steel tube with heating tape outside the tube to preheat the feed and solvent, which helped to minimize the temperature and concentration fluctuations during reaction initiation.

4.2.3 Experimental procedure

The experiments started with catalyst pretreatment. Solvent (140 mL, cyclohexane, Sigma-Aldrich, >99.9% HPLC grade) was degassed in an ultrasonic cleaner (BRANSON 5510) before being charged along with a known quantity of catalyst into the reactor. The wet catalyst was pre-dried using a rotavapor under vacuum conditions in a 40 °C water bath (BUCHI R-215, B-491 and V-700) before weighing and charging. The reactor was purged 3 times and pressurized to the target value with hydrogen (99.999%, Ultra high purity). For the catalyst pretreatment, the reactor temperature was set 15 °C higher than the reaction temperature. The pretreatment was conducted for 30 min with 800 rpm agitation speed. A total 60 mL of reagent solution (acetophenone, Fluka, >99%, or 1-phenylethanol, Aldrich, >98%; 1-15 mL, depending on the initial concentration, in solvent) was prepared in the injector after degassing, then preheated to the reaction temperature during the pretreatment. After pretreatment, the reactor temperature was
decreased to the desired reaction temperature, and preheated feed solution was injected into the reactor. During the reaction, samples were taken from the reactor using an \( \frac{1}{8}'' \) stainless steel tube and a 3-way valve. Each sample amount was approximately 1 mL, and seven samples were typically taken from the reactor during the course of an experiment. After filtering, each sample was analyzed by gas chromatography (HP 5890 II equipped with an Agilent DB-WAXetr capillary column, 50 m \( \times \) 0.32 mm, and a flame ionization detector) under the following conditions: helium as carrier gas (2.6 mL/min), inlet and detector temperatures 250 °C, oven temperature 150 °C, and injection volume 0.5 µL with 1:100 split. Six possible reaction intermediates or products were purchased and used for GC calibrations: acetophenone (AP, Fluka, >99%), 1-phenylethanol (PE, Aldrich, >98%), cyclohexylmethylketone (CMK, Aldrich), 1-cyclohexylethanol (CE, Aldrich, 97%), ethylbenzene (EB, Sigma-Aldrich, 99.8%) and ethylcyclohexane (EC, Aldrich, >99%). The \( R^2 \) values of all GC calibrations were >99.6%. The repeatability error for each sample was less than 4%. The mass balances were checked after every experiment and were found to close within 5%.

4.3 Results and discussion

4.3.1 Reaction route validation

Before conducting the study on intrinsic reaction kinetics and the effects of internal diffusion, the reaction route for AP hydrogenation was validated, and the target product and model reaction were selected for future study in a trickle bed reactor. The reaction pathways of AP hydrogenation are shown in Figure 4.2.
All six components shown in the scheme were observed by GC analysis of AP hydrogenation, as shown in Figure 4.3a. Most literature reports of this reaction have observed a similar reaction pathway, although some researchers observed additional intermediates or neglected some observed in this work [110, 114-117]. To confirm this route, separate hydrogenations of AP, PE and CE with 1% Rh/Al₂O₃ were carried out.
Figure 4.3: Concentration-time profiles for hydrogenation of (a) acetophenone and (b) phenyl ethanol with 1% Rh/Al₂O₃ catalyst: (a) 0.6 M C_{AP₀}, 2 g_{cat}, 2.6 MPa P_{H₂}, 80 °C at start, 150 °C after 40 min.; (b) 0.12 M C_{PE₀}, 0.3 g_{cat}, 2.6 MPa P_{H₂}, 80 °C.
In the experiment (Figure 4.3a), the reaction temperature was increased from 80 to 150 °C after 40-50 min to accelerate the reaction, obtain more EB and EC, and evaluate selectivity. Only small amount of EB was produced at 80 °C, with increased levels at 150 °C, similar to reports with platinum catalysts [114]. Hydrogenation of PE was shown to produce CE, EB, and even CMK by regeneration of the carbonyl group, as reported in the literature [115]. In the present study, however, the PE→CMK route was ignored because its yield was below 10% and because also the obtained CMK is converted to CE, as shown in Figure 4.3b. No reaction was observed in our tests with CE as feed at 80-150 °C and 2.6 MPa P\textsubscript{H\textsubscript{2}}, similar to prior report in the literature with platinum [110].

Based on these results, the pathway of AP→PE→CE was selected as the target reaction route in this study, with CE as the target product.

### 4.3.2 Catalyst screening

Rhodium, ruthenium, and platinum were tested to select the appropriate catalyst for this study, with results shown in Figures 4.4 and 4.5. Palladium did not perform as desired because it preferentially followed the AP→PE→EB route, as noted in prior reports [118-120].
Figure 4.4: Concentration-time profiles for hydrogenation of acetophenone with (a) 5% Ru/Al₂O₃ and (b) 1% Pt/Al₂O₃ at 0.12 M C₅₆.₀, 0.3 gcat, 2.6 MPa Pₓ and 80 °C.
Figure 4.5: Concentration-time profiles for hydrogenation of acetophenone with (a) 5% Rh/C and (b) 1% Rh/Al₂O₃ at 0.12 M CAP₀, 0.3 gcat, 2.6 MPa P₇ and 80 °C.
As shown in Figure 4.4, ruthenium preferentially produced CMK, similar to previous reports while platinum produced PE, as did rhodium. [105, 112, 121], Because AP→PE→CE is the desired pathway, ruthenium was eliminated from consideration. The conversion rate on platinum, however, was lower than on rhodium, as shown in the concentration-time profiles in Figures 4.4b and 4.5b. The initial hydrogenation rates below 40% conversion of AP, were 0.057 and 0.1625 mmol/g_cat.s for platinum and rhodium catalysts, respectively. The faster reaction rate on rhodium was considered to be preferable for the study of the effect of mass transfer in a trickle bed reactor.

To select the appropriate catalyst material, rhodium on activated carbon and on alumina support was tested and compared. As shown in Figure 4.5, Rh/C produced significant amounts of EB and EC, in contrast to Rh/Al₂O₃. This may occur due to the higher acidity of alumina, which was shown in the past to have low selectivity towards EB [122, 123]. Alumina support also has excellent mechanical stability and is therefore preferred industrially [124]. Thus, rhodium on alumina support was selected as the catalyst for this study.

4.3.3 Intrinsic reaction kinetic study

With 1% Rh/Al₂O₃ selected from the catalyst screening, the intrinsic reaction kinetic study was conducted. The experimental parameter range was 1.1-4.1 MPa \( P_{H_2} \), 60-100 °C, and 0.04-0.4 M \( C_{AP,o} \) with 0.1-2 g catalyst mass. The temperature range was selected to exclude EB and EC formation from the kinetic study because significant amounts of EB and EC (> 20%) were obtained in tests at 120-150 °C, similar to PE hydrogenation on platinum catalysts described in the literature [114]. Preventing EB and EC formation
avoids more complex L-H kinetics, especially in the adsorption term in the denominator. If all six chemicals were considered, there would be seven adsorption terms, including hydrogen but ignoring the solvent. In this case, there would be too many parameters to fit reliably. Furthermore, including EB and EC was not necessary for the selected reaction route and target product. In the selected temperature range, the mole fractions of EB and EC were each less than 0.1, and typically below 0.05, which could be ignored. Although EB and EC were excluded in the kinetic modeling, they were quantified by GC and are discussed in this work. Therefore, the reaction scheme for kinetic modeling was rearranged with four species (AP, PE, CMK and CE) and four reactions composed of hydrogenations of the carbonyl double bond \( r_{d1} \) and \( r_{d2} \) and the aromatic ring \( r_{r1} \) and \( r_{r2} \).
### Table 4.1: Properties of 1% Rh/Al₂O₃ catalyst

<table>
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<tr>
<th></th>
<th>Mean diameter</th>
<th>Surface area</th>
<th>Pore diameter</th>
<th>Porosity</th>
<th>Rh content</th>
<th>Shell depth</th>
<th>Rh dispersion</th>
<th>Metal crystallite size</th>
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<tbody>
<tr>
<td></td>
<td>mm</td>
<td>m²/g</td>
<td>nm</td>
<td></td>
<td>Total wt%</td>
<td>Shell wt%</td>
<td>Core wt%</td>
<td>mm</td>
<td>%</td>
</tr>
<tr>
<td>Powder (11769)</td>
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<td>177</td>
<td>11.5</td>
<td>0.65</td>
<td>1.00</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>72.3</td>
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<tr>
<td>0.5mm dp (S55366)</td>
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<td>11.3</td>
<td>0.67</td>
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<td>0.04</td>
<td>48.4</td>
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<tr>
<td>2.5mm dp (47040)</td>
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<td>216</td>
<td>9.2</td>
<td>0.66</td>
<td>0.94</td>
<td>1.88</td>
<td>0.29</td>
<td>0.2</td>
<td>48.9</td>
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### Table 4.2: Kinetic modeling results based on L-H kinetics for AP hydrogenation.

<table>
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<tr>
<th>Model No.</th>
<th>Model detail</th>
<th>G</th>
<th>R²(%)</th>
<th>ln k₂₁</th>
<th>ln k₂₂</th>
<th>ln K₂₁</th>
<th>ln K₂₂</th>
<th>E₂₁</th>
<th>E₂₂</th>
<th>k₂₁ or k₂₂</th>
<th>ΔH₂₁ or k₂₂</th>
<th>K₂₁</th>
<th>K₂₂</th>
<th>ΔE₂₁</th>
<th>ΔE₂₂</th>
<th>k₂₁</th>
<th>k₂₂</th>
</tr>
</thead>
<tbody>
<tr>
<td>I</td>
<td>non-competitive, dissociative</td>
<td>1.240</td>
<td>96.2</td>
<td>31.27</td>
<td>22.80</td>
<td>40.32</td>
<td>21.24</td>
<td>62.30</td>
<td>41.96</td>
<td>25.25</td>
<td>-17.88</td>
<td>396.93</td>
<td>2.3E-09</td>
<td>695.62</td>
<td>62.94</td>
<td></td>
<td></td>
</tr>
<tr>
<td>II</td>
<td>competitive, dissociative</td>
<td>1.405</td>
<td>95.1</td>
<td>33.43</td>
<td>24.34</td>
<td>41.41</td>
<td>24.01</td>
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<td>48.07</td>
<td>18.83</td>
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<td>1.89</td>
<td></td>
<td></td>
</tr>
<tr>
<td>III</td>
<td>non-competitive, non-dissociative</td>
<td>1.276</td>
<td>96.0</td>
<td>30.99</td>
<td>19.24</td>
<td>-19.62</td>
<td>21.15</td>
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<td>12.85</td>
<td>25.9</td>
<td>23.54</td>
<td>1.38</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>IV</td>
<td>competitive, non-dissociative</td>
<td>1.332</td>
<td>95.6</td>
<td>32.60</td>
<td>23.23</td>
<td>18.60</td>
<td>23.06</td>
<td>76.48</td>
<td>53.63</td>
<td>16.90</td>
<td>-32.15</td>
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<td>13.70</td>
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</tr>
<tr>
<td>V</td>
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<td>1.214</td>
<td>96.5</td>
<td>29.78</td>
<td>21.95</td>
<td>17.59</td>
<td>19.06</td>
<td>71.94</td>
<td>53.46</td>
<td>47.22</td>
<td>-26.79</td>
<td>11.80</td>
<td>46.16</td>
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<tr>
<td>VI</td>
<td>Model I &amp; saturated, AP+PE data</td>
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<td>95.0</td>
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<td>25.83</td>
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<td>63.57</td>
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<td>46.16</td>
<td>5.14</td>
<td>1.29</td>
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</table>

\[
k_i = k_i' e^{-\frac{E_i}{R} \frac{1}{T}}; K_{H_1 or H_2} = k_{H_1 or H_2} e^{-\frac{\Delta H_{H_1 or H_2}}{R} \frac{1}{T}}
\]

\[
i = d_1, d_2, r_1 or r_2; x = AP, CMK, PE or CE; R = 0.008314 \text{ kJ/mol.K} ;
\]
As summarized in Table 4.1, the surface area, mean pore diameter, and porosity of 1% Rh/Al₂O₃ powder catalyst were 177 m²/g, 11.5 nm, and 0.65, respectively, as characterized by BET (Micromeritics, ASAP 2000), with the porosity confirmed by true density and envelope density measurements by pycnometers (Micromeritics, Accupyc II 1340 and Geopyc 1360). The two porosity values differed by <10% and were similar to 0.66 porosity (with 50-300 m²/g surface area) reported as typical values for γ-alumina powder in the literature [124].

Rhodium dispersion on powdered alumina support was 72.3±2.3%, as determined by the H₂-O₂ titration technique described previously [125, 126]. The operating and surface cleaning temperatures were 200 and 400 °C, respectively, following prior publications on rhodium reduction [127, 128]. The stoichiometry for both oxygen and hydrogen chemisorption on rhodium was taken to be 1.5 [129].

Cyclohexane was selected as the solvent for this work not only because it provided a faster reaction rate with relatively high solubility of hydrogen but also because it was expected to result in relatively low interaction between gas, liquid reagents, and catalyst surface [130-132].

4.3.3.1 Reaction controlled regime

The reaction kinetic study should be conducted in the reaction controlled regime where the measured rate is free from mass transfer limitations in gas-liquid or liquid-solid interfacial films between catalyst particles and from pore diffusion limitations inside the particle [133].
The mass transfer rate in gas-liquid interfaces depends on the agitation power (rpm) in 3-phase stirred tanks [134]. Thus, if the apparent reaction rate does not depend on agitation rate, it can be absent of gas-liquid mass transfer limitation [135, 136]. To confirm this and to set the appropriate agitation speed for this study, reaction tests were conducted at 600-1000 rpm, which corresponds to 300-1300 W/m$^3$ in the slurry reactor using an agitator power number of 1.15, as confirmed in our prior work [85]. An extreme operating condition (4.1 MPa $P_{H_2}$, 100 °C, 0.4 M $C_{AP,o}$ and 2 g catalyst) was selected for the tests.

As shown in Figure 4.6, there was no change of initial hydrogen consumption and initial AP hydrogenation rate between 600 and 1000 rpm. In addition, initial hydrogen
consumption was linear and directly proportional to catalyst amount (0.5-2 g). These results confirmed that gas-liquid mass transfer limitations were absent. As a result, 800 rpm was determined to be a suitable stirring speed for this study.

With respect to the liquid-solid mass transfer effect, we applied the criterion $r_X / (k_X a_p C^*_X) < 0.1$ for absence of liquid-solid mass transfer limitations [137]. For this calculation, $Sh_p = 2$ was used as the lower bound [138]. Molecular diffusivities for organics were estimated by the correlation of Tyn and Calus with the Brock and Bird corresponding states method [72, 139]. Hydrogen solubility and molecular diffusivity in cyclohexane were obtained from the literature [140]. The calculated criteria values for hydrogen and AP were $3.7 \times 10^{-4}$ and $8.5 \times 10^{-4}$, respectively, both much smaller than 0.1, confirming the absence of the liquid-solid mass transfer limitations.

The absence of intraparticle diffusion limitations was confirmed using the Weisz-Prater criterion [141]. Effective diffusivities for the criterion were calculated using molecular diffusivities, porosity 0.65 as noted above, tortuosity 4, and constriction factor 0.8, the latter two obtained for alumina from literature [142, 143]. Calculated Weisz-Prater parameter values for hydrogen and AP were 0.02 and 0.06, respectively, satisfying the criterion of being less than 0.3 for negligible diffusion limitations [133]. Thus, all experimental data used in the intrinsic reaction kinetic study were obtained in the reaction controlled regime, absent of diffusion and mass transfer limitations.

4.3.3.2 Experiments and initial rates

In the experimental range described earlier, 23 experiments of AP hydrogenation for the intrinsic kinetic study were conducted. Experimental designs at 0.4 and 0.04 M $C_{AP,0}$
were composed of 9 experiments each with 60, 80, 100 °C temperature and 1.1, 2.6, 4.1 MPa $P_{H_2}$, according to $3^2$ factorial design [144]. 5 other experiments were added, which consisted of 3 different intermediate initial concentration levels at 80 °C and an intermediate level at 60 and 100 °C. Initial rates of hydrogenation were obtained using differentiation formula for initial points [142]. The initial apparent reaction orders for hydrogen partial pressure and initial AP concentration were first and zero order, respectively, as shown in Figure 4.7.
Figure 4.7: Initial hydrogenation rate versus (a) hydrogen partial pressure at 0.4 M $C_{AP,0}$ and (b) initial acetophenone concentration at 2.6 MPa $P_{H_2}$ and different temperatures.
4.3.3.3 Kinetic modeling

Based on the above results, kinetic modeling following the Langmuir-Hinshelwood hypothesis was considered (L-H kinetics), which is generally accepted for hydrogenation of organic compounds [145].

First, adsorption steps were evaluated to determine whether they were the rate-determining step (r.d.s.). If the adsorption of organics on active sites is the r.d.s., the rate would not be first-order with respect to hydrogen partial pressure, as determined above. Although having first order for hydrogen and zero order for \( C_{AP,0} \) could be satisfied with hydrogen adsorption as the r.d.s., kinetic modeling from the assumption cannot explain the variation of acetophenone concentration with time observed in our experiments (i.e., change of reaction order, starting at zero order but increasing at later reaction time), as shown in Figure 4.5b. Thus, surface reactions were assumed to be the rate determining steps for all reactions in this work.

Four different cases based on prior liquid-phase hydrogenation studies were considered, with either competitive or non-competitive adsorption of hydrogen and organics and with either dissociative or non-dissociative adsorption of hydrogen on catalyst active sites [109, 145-147]. With surface reaction between adsorbed hydrogen and organics being rate limiting, the following rate expressions are obtained:

For hydrogenation of the carbonyl group: 
\[
r_{d1} = k_{d1} K_{AP} C_{AP} K_{HorH_2} P_{H_2} C_v \quad (4.1)
\]

(d1: AP\( \rightarrow \)PE; d2: CMK\( \rightarrow \)CE) 

\[
r_{d2} = k_{d2} K_{CMK} C_{CMK} K_{HorH_2} P_{H_2} C_v \quad (4.2)
\]

For hydrogenation of the aromatic group: 
\[
r_{r1} = k_{r1} K_{AP} C_{AP} K_{HorH_2} P_{H_2} C_v \quad (4.3)
\]
The concentration of vacant sites, $C_v$, is calculated as follows:

Non-competitive, dissociative (model I):

$$
C_v = \frac{1}{(1 + K_{AP} C_{AP} + K_{PE} C_{PE} + K_{CMK} C_{CMK} + K_{CE} C_{CE})(1 + \sqrt{K_H P_{H_2}})^2}
$$

Competitive, dissociative (model II):

$$
C_v = \frac{1}{(1 + K_{AP} C_{AP} + K_{PE} C_{PE} + K_{CMK} C_{CMK} + K_{CE} C_{CE} + \sqrt{K_H P_{H_2}})^3}
$$

Non-competitive, non-dissociative (model III):

$$
C_v = \frac{1}{(1 + K_{AP} C_{AP} + K_{PE} C_{PE} + K_{CMK} C_{CMK} + K_{CE} C_{CE})(1 + K_{H_2} P_{H_2})}
$$

Competitive, non-dissociative (model IV):

$$
C_v = \frac{1}{(1 + K_{AP} C_{AP} + K_{PE} C_{PE} + K_{CMK} C_{CMK} + K_{CE} C_{CE} + K_{H_2} P_{H_2})^2}
$$

An additional consideration in obtaining the rate expressions is that for aromatic ring hydrogenation, the r.d.s. is the insertion of the first two H atoms [148]. Because the experimental data was obtained as concentrations vs. time, integration of mass balances in the slurry reactor was required to determine kinetic parameters from fitting
experimental data. The mass balances of organic components in the slurry reactor are as follows:

\[
\begin{align*}
\frac{dC_{AP}}{dt} &= -\rho_r F_c (\eta_{d1} r_{d1} + \eta_{r} r_{r}) \\
\frac{dC_{PE}}{dt} &= -\rho_r F_c (\eta_{r2} r_{r2} - \eta_{d1} r_{d1}) \\
\frac{dC_{CMK}}{dt} &= -\rho_r F_c (\eta_{d2} r_{d2} - \eta_{r} r_{r}) \\
\frac{dC_{CE}}{dt} &= \rho_r F_c (\eta_{r2} r_{r2} + \eta_{d2} r_{d2}) \\
C_x &= C_{x,0} \text{ at } t = 0 \quad (4.9)
\end{align*}
\]

With powder catalysts, effectiveness factor \( \eta \) for all reactions is 1 owing to no external or internal diffusion limitations, as confirmed in section 4.3.3.1. In addition, a conversion factor \( (F_c) \) was used to obtain intrinsic reaction rates based on TOF (turnover frequency), which accounts for metal dispersion.

4.3.3.4 Parameter fitting

To calculate molar concentrations of organics with reaction time, the mass balance equations were integrated by the 4th-order Runge-Kutta method using MATLAB. The parameters for each model were obtained by minimizing the objective function \( G \) using nonlinear optimization with the Nelder-Mead simplex algorithm [149]. The objective function was:

\[
G = \sqrt{\sum_x \sum_t |C_{s,t} - C_{s,t}^{CALC}|^2} 
\]

(10)
After determining \( G \), the average \( R^2 \) values, which indicate goodness of fit, were used to select the proper kinetic model. Lower \( G \) values and higher average \( R^2 \) values were considered indicative of greater model accuracy.

### 4.3.3.5 Model evaluation and improvements

The evaluation results with optimized parameters for each model are presented in Table 4.2, where the \( G \) and \( R^2 \) values are averaged over the 23 experiments conducted. Models I-IV depend on hydrogen adsorption, as discussed in section 4.3.3.3. With lower \( G \) and higher \( R^2 \) values, both non-competitive models (I and III) fit better than the competitive models (II and IV). This is also supported by the fact that the non-competitive models more easily satisfy zeroth-order dependence on \( C_{AP,o} \), regardless of the hydrogen adsorption term magnitude. Model III leads to a negative value for \( E_{d2} \). Furthermore, for model III, the highest value of hydrogen adsorption term using the fitting parameters was 0.24, while the corresponding value for model I was 0.011; these values should be sufficiently smaller than 1 to obtain the first-order dependence on hydrogen, as shown in Figure 4.7a.

For these reasons, model III was discarded from consideration. Additionally, for model I, the solvent effect was accounted for by insertion of the \( K_{CHCCH} \) term in the denominator for adsorption of organics. The overall value of \( G \), however, did not improve, indicating that the solvent effect could be ignored. Finally, because Model I does not exhibit zeroth-order behavior for \( C_{AP,o} \) unless \( K_{AP}C_{AP} >> 1 \), the \( C_v \) term adjusts to:

\[
C_v = \frac{1}{(K_{AP}C_{AP} + K_{PE}C_{PE} + K_{CMK}C_{CMK} + K_{CE}C_{CE})(1 + \sqrt{K_{H^2}P_{H_2}})^2} \tag{4.11}
\]
leading to model V. The term $K_{AP}C_{AP} >> 1$ essentially means that the active sites for organics adsorption are saturated, as reported in other cases [150, 151]. In addition to having more reasonable parameters, particularly $K_x$, model V also provides somewhat better accuracy than model I (lower $G$ and higher $R^2$ values).
Figure 4.8: Comparison of experimental kinetic data with estimation results using model V: (a) 60 °C, 4.1 MPa $P_{H_2}$, 0.04 M $C_{AP,o}$ and 0.2 g$_{cat}$ ($R^2 = 0.983$); (b) 80 °C, 2.6 MPa $P_{H_2}$, 0.08 M $C_{AP,o}$ and 0.2 g$_{cat}$ with ($R^2 = 0.959$); (c) 100 °C, 1.1 MPa $P_{H_2}$, 0.4 M $C_{AP,o}$ and 0.8 g$_{cat}$ with ($R^2 = 0.934$).
Using the fitted parameters, species concentrations were predicted as functions of time. Figure 4.8 shows a comparison of the experimental and predicted values for three cases that cover the range of $R^2$ values for all 23 experiments (0.90 to 0.995, with 20 of the 23 values between 0.93 and 0.98). It may be seen that the predictions match the experiments well. Thus, model V, which involves non-competitive, dissociative hydrogen adsorption with saturated organics assumption and surface reaction as the rate determining step, was selected in this study.

4.3.3.6 Discussion

Because it is difficult to fully characterize complex reactions on solid surfaces using simple L-H mechanisms, opinions vary regarding adsorption phenomena of hydrogen and organic chemicals. For the rate-determining step, most prior publications selected surface reaction because relatively simple equations resulting from adsorption or desorption as the r.d.s. cannot generally explain the transient reaction behavior [109, 124, 145], and the same conclusion was reached in the present work, as well. Between dissociative and non-dissociative hydrogen adsorption, the former has been preferred for carbonyl group hydrogenation [146, 152-154], while the latter has been selected frequently for aromatic ring hydrogenation [115, 151, 155]. However, most literature on adsorption of hydrogen on noble metals, especially rhodium, reported that the hydrogen adsorption is dissociative and only dissociatively adsorbed hydrogen is catalytically active [156-159]. For this reason, the selection of dissociative adsorption of hydrogen in this work is reasonable. For competitive vs. non-competitive adsorption between hydrogen and other species, most studies have evaluated both cases as candidates [145].
There is a trend in the literature frequently reporting carbonyl and aromatic hydrogenations as having competitive and non-competitive adsorption, respectively [153, 160-162]. Nevertheless, there are opposing opinions for each case [152, 155], and the relative dominance of the two models changes with temperature [163]. For hydrogenation of complex chemicals including aromatic rings, non-competitive adsorption is likely preferred because even for sites saturated by aromatic rings, hydrogen can adsorb due to the size difference between aromatic rings and hydrogen [150]. The saturation coverage of active sites for organics is also possible due to strong adsorption of aromatics [164] and is often used to explain reaction orders [109]. The background knowledge from literature supports the selected model including dissociative hydrogen adsorption that is non-competitive with organics for saturated active sites.

The parameter values for the selected model V were consistent with those found in the literature. The heat of adsorption of hydrogen in this work was found to be -30.7 kJ/mol, which is in the range of reported values, -17 to -50 kJ/mol [115, 151, 155, 165]. The obtained activation energy values are within the ranges reported for carbonyl and aromatic hydrogenations in the literature, 17 to 100 kJ/mol [110, 111, 113, 155]. The adsorption constants in this work are in the order $K_{PE} > K_{AP} > K_{CMK} > K_{CE}$, which are reasonable because (1) aromatic rings adsorb more strongly onto catalytic sites [150, 164], (2) the reaction rate of CMK is slower than those of AP or PE, and (3) CE did not react with hydrogen, as confirmed in our tests. The K values are also in the range reported in the literature [155, 166, 167]. Thus, model V represents the intrinsic kinetics of AP hydrogenation on 1% Rh/Al$_2$O$_3$ well and is sufficiently precise to be used for other studies, including internal diffusion effects.
4.3.4 Effects of internal diffusion

Since catalyst pellets, not powder catalyst, are used in tubular or fixed bed reactors for practical reasons, including low pressure drop and operational convenience, it is important to determine the effects of internal diffusion on the rate of reaction. In this work, 0.5-mm and 2.5-mm diameter catalyst spheres were used to study these features in a slurry reactor, with the intrinsic kinetics obtained in section 4.3.3.

4.3.4.1 Catalyst characterization

To assess internal diffusion effects, characterization of 2.5-mm, 0.5-mm, and powder catalysts was conducted. Table 4.1 shows the results for different sized 1% Rh/Al₂O₃ catalysts. The 2.5- and 0.5-mm catalysts were characterized using the same particle size analyzer, BET, pycnometers, and chemisorption method used for powders in sections 4.2.3 and 4.3.3. From rhodium dispersion ($D$), assuming spherical shape, the metal crystallite size ($d_m$) was calculated from $d_m = 6(v_m/a_m)/D$ with $13.78 \, \text{Å}^3$ for $v_m$, volume occupied by an atom in bulk metal, and $7.58 \, \text{Å}^2$ for $a_m$, area occupied by a surface atom for rhodium [168]. SEM/EDX was used for catalyst distribution on larger supports, confirming typical egg-shell distribution, as shown in Figure 4.9.
The shell depths, as shown in Table 4.1, were determined by averages of several supports. The Rh content values in the shell and core were obtained using the ratio between the values for the shell and core from EDX analyses and known total Rh content.

4.3.4.2 Experimental results

A total of 14 experiments, 7 each for 2.5- and 0.5-mm catalysts, were conducted over the same range of experimental conditions (60-100 °C, 1.1-4.1 MPa P_{H_2} and 0.04-0.4 M C_{AP,o}) with the same experimental and analysis procedures described previously for the intrinsic
As compared with the results for powder catalysts (Figure 4.5b), the reaction rates with 0.5- and 2.5-mm catalysts were significantly lower, as shown in Figure 4.10, as expected based on prior studies [169].

Figure 4.10: Typical experimental (symbols) and modeling results (solid lines) for acetophenone hydrogenation on 1% Rh/Al2O3: (a) 0.5-mm, (b) 2.5-mm particles at 80 °C, 2.6 MPa P\textsubscript{H₂}, 0.12 M C\textsubscript{AP,o} and 0.3 g\textsubscript{cat}. 
The maximum yield fraction of the major intermediate, PE, decreased with increased particle diameter, from 0.61 for powder to 0.34 for 0.5-mm and 0.22 for 2.5-mm catalysts. In these experiments, some EB and EC (see reaction scheme, Figure 4.2) were also observed in the liquid phase, although the sum of their concentrations was always less than 10% and 20% of $C_{AP,o}$ for the 0.5-mm and 2.5-mm catalysts, respectively.

With larger catalysts, the liquid-solid (L-S) mass transfer limitation should be evaluated due to the change in L-S interfacial area, while the gas-liquid mass transfer of hydrogen is not affected by catalyst size. The criterion for L-S mass transfer limitation was determined, as described in section 4.3.3.1, using the mass transfer correlation for larger catalysts [170]. Out of the fourteen experiments, three resulted in values of $r_x/(k_a a_p C_x)$ > 0.1, namely, 0.18 and 0.38 for 2.5-mm catalyst and 0.12 for 0.5-mm, while all remaining cases were < 0.1. These values are only slightly above 0.1, and even with a value of 0.38, the experiment matched model-predicted profiles with an $R^2$ of 0.98. Therefore, the external liquid-solid mass transfer effect was ignored. By comparison, the Weisz-Prater parameter values (which should be < 0.3 to allow neglecting internal diffusion) were 11.8-50.4 for 0.5-mm catalysts and 152-523 for 2.5-mm catalysts at 80 °C, confirming the dominance of internal diffusion in the experiments for both catalysts.

4.3.4.3 Modeling of internal diffusion effects

To assess the effects of simultaneous internal diffusion and reaction, we conducted a modeling study of species concentrations inside catalyst particles. For spherical catalyst particles, the mass balances of AP, PE, CMK, CE, and H$_2$ are expressed by the following equations [169]:
\[
\frac{d^2 C_{H_2}}{dr^2} + \frac{2}{r} \frac{dC_{H_2}}{dr} - \frac{\rho_p F_{ce}}{De_{H_2}} (r_{d1} + 3r_{d1} + r_{d2} + 3r_{d2}) = 0
\]
(4.12)

\[
\frac{d^2 C_{Ap}}{dr^2} + \frac{2}{r} \frac{dC_{Ap}}{dr} - \frac{\rho_p F_{ce}}{De_{Ap}} (r_{d1} + r_{d1}) = 0
\]
(4.13)

\[
\frac{d^2 C_{Pe}}{dr^2} + \frac{2}{r} \frac{dC_{Pe}}{dr} - \frac{\rho_p F_{ce}}{De_{Pe}} (r_{d2} - r_{d1}) = 0
\]
(4.14)

\[
\frac{d^2 C_{CMK}}{dr^2} + \frac{2}{r} \frac{dC_{CMK}}{dr} - \frac{\rho_p F_{ce}}{De_{CMK}} (r_{d2} - r_{d1}) = 0
\]
(4.15)

\[
\frac{d^2 C_{CE}}{dr^2} + \frac{2}{r} \frac{dC_{CE}}{dr} + \frac{\rho_p F_{ce}}{De_{CE}} (r_{d2} + r_{d2}) = 0
\]
(4.16)

along with the boundary conditions:
\[
\frac{dC_i}{dr} \bigg|_{r=0} = 0; \quad C_x = C_{x,r_p}
\]
(4.17)

The effectiveness factors for the various reactions are given by:
\[
\eta_i = \frac{4 \pi \int_0^{r_p} r^2 F_{ce} r_i (r) dr}{4 \pi r^3_p F_{ce} r_i}
\]
(4.18)

In addition, to describe \( C_{H_2}(r) \) in the various rate expressions, we have:
\[
H \cdot C_{H_2} = P_{H_2} f or \ r_i
\]
(4.19)

where the Henry’s law constant \( H \) is obtained by fitting literature data [140]:
\[
H = 0.1 \cdot \exp(-12.58 + \frac{1446.2}{T} + 2.7261 \cdot \ln(T))
\]
(4.20)

Since both particles have egg-shell-distributed catalyst on support and different metal dispersion than powder, conversion factors \( F_c \) and \( F_{ce} \) were applied to the TOF-based intrinsic kinetics equations described in section 4.3.3.3. For \( F_{ce} \), the radius-dependent catalyst loading was considered separately in the core and shell zones, with the same metal dispersion in each zone. Since TOF changes with metal size, a correction factor, \( f \)
= 1.15 for both particles, was used to adjust the reaction rate based on the results for Rh available in the literature [136]. The effective diffusivities, $D_{e}$, were obtained using the same methods and values as those described in section 4.3.3.1.

We also calculated the heat transfer parameter values to assess whether intraparticle heat effects were present. The estimated values for all 4 reactions over 60-100 °C varied from $2.0 \times 10^{-4}$ to $7.2 \times 10^{-4}$ due to low effective diffusivities in liquids and high thermal conductivities of pellets, much lower than 0.1, indicating the absence of intraparticle heat effects [169].

Equations (4.12)-(4.17) were solved by the method of lines [171]. These numerical results were combined with initial-value ordinary differential equations (4.9) for the stirred batch reactor, as described in section 4.3.3.3, and solved by the 4th-order Runge-Kutta method.

4.3.4.4 Model evaluation and discussion

The stirred batch reactor model coupled with internal diffusion-reaction matched the experimental results well, with $R^2$ values 0.962 over all 14 cases described in section 4.3.4.2. The comparison of experimental and model-predicted profiles for the two catalyst particle sizes for one experimental condition is shown in Figure 4.10. This demonstrates that combining the results of intrinsic reaction kinetics from powder catalysts with diffusion-reaction in larger particles, accounting for metal dispersion and the catalyst distribution profile, is appropriate. The $\eta$ values for AP reactions (d1 and r1) were 0.08-0.26 for 2.5-mm particles and 0.2-0.7 for 0.5-mm due to larger internal diffusion limitations with larger particles.
Figure 4.11: Concentration profiles inside catalyst particles at 80 °C, 2.6 MPa $P_{H_2}$ and 0.12 M $C_{AP.o}$ using (a) 0.5-mm diameter particles, $\eta=0.467$, (b) 2.5-mm particles, $\eta=0.144$. 
Figure 4.11 demonstrates the decrease in observed reaction rates caused by internal diffusion limitations. For 2.5-mm catalyst, due to larger diffusion limitation, AP exhibits a larger internal concentration gradient (Figure 4.11b) and is essentially absent in the particle core for all cases. In the 0.5-mm particles with a higher effectiveness factor on the other hand, the concentration gradient is smaller, and some AP remains in the core, as shown in Figure 4.11a. Hydrogen dissolved in the liquid within the particles also reduces the internal reaction rate due to its concentration profile, although it is not completely depleted inside the particle owing to faster diffusion. Even in the most limiting case (1.1 MPa $P_{H_2}$ and 60 °C with 2.5-mm particles), there remains ~2% of the surface concentration in the core. Since they are products of reactions, the concentrations for both intermediate (PE and CMK) and final (CE) products are higher within the particles than on the surface (or in the bulk liquid). It is noteworthy that concentration vs. radius profiles of intermediates PE and CMK exhibit maxima in the larger 2.5-mm particles, but not for the smaller 0.5-mm ones. These features also explain the lower maximum PE yield for larger particles.

In this work, for both 1% Rh/Al$_2$O$_3$ particle sizes, we used 4 for the value of tortuosity, selected as the middle of the 2-6 range for alumina reported in the literature [143, 172], and 0.65 for porosity, averaged from the measured value range of 0.62-0.69 as described in section 3.3.1. To evaluate the robustness of model estimations with different tortuosity and porosity in the available range, a sensitivity study was conducted with 0.69 porosity or 2.7 tortuosity values, as shown in Figure 4.12.
Figure 4.12: Sensitivity study for concentration profiles versus reaction time with different tortuosity and porosity at 80 °C, 2.6 MPa $P_{H_2}$ and 0.12 M CAP.o using 2.5 mm diameter catalysts with (a) $\tau = 4$, $\epsilon_p = 0.69$ and (b) $\tau = 2.7$, $\epsilon_p = 0.65$, solid lines. The dashed lines are the corresponding results with $\tau = 4$, $\epsilon_p = 0.65$. 
The effect of tortuosity was larger than that of porosity due to the broader considered tortuosity range [173]. Changes caused by the porosity and tortuosity, however, were not significant, indicating that their values within the considered ranges are satisfactory to describe the experimental results.

4.4 Conclusions

In this work, acetophenone hydrogenation was conducted with 1% Rh/Al₂O₃ catalysts selected by screening different noble metals and supports for AP→PE→CE reaction selectivity and rapid reaction rates. The intrinsic reaction kinetic model based on the L-H mechanism was obtained with experiments conducted in the ranges of 60-100 °C, 1.1-4.1 MPa P$_{H_2}$ and 0.04-0.4 M C$_{AP.o}$ with powder catalysts. The kinetic model including dissociative and non-competitive hydrogen adsorption with saturated active sites for organic species was selected after detailed analysis and comparison of the parameters with literature values. With the obtained reaction kinetic model, internal diffusion effects were studied with different sized spherical catalyst particles, characterized to have egg-shell type catalyst distributions. Catalyst properties were characterized through various physical and chemical techniques. The diffusion-reaction models accounting for egg-shell distributions and metal dispersion fit the experimental results well, explaining the effects of internal diffusion inside catalyst particles on reaction rates and yields. The results of this work are suitable for use in future trickle-bed reactor modeling studies.
CHAPTER 5. ACETOPHENONE HYDROGENATION ON RH/AL₂O₃ CATALYST: FLOW REGIME EFFECT AND TRICKLE BED REACTOR MODELING

5.1 Introduction

Trickle-bed reactors are three phase fixed-bed reactors, with gas and liquid in cocurrent downflow over solid catalyst bed [32, 174]. These reactors are used for various applications including hydrogenation and wet oxidation in the refinery, fine chemical and pharmaceutical industries. Applications in the pharmaceutical industry are typified by high molecular weight feed chemicals, complex reaction schemes, various intermediate chemicals and relatively higher pressure than atmospheric. This industry is focused on high productivity using continuous operations [46]. Owing to complex hydrodynamics influencing reactor performance, however, it is not easy to design and scale up trickle bed reactors [174].

The flow regimes in trickle beds are typically classified as follows: trickle flow, spray flow, pulsing flow and bubbly flow [35], determined by the gas and liquid flow rates, physical properties and the nature of reactor packing. While trickle flow obtained in low gas and liquid flow rates is calm and stable, pulsing flow is characterized by alternatively changing gas and liquid rich zones in moderately increased gas and liquid flow rates from trickle flow. Both pulsing and bubbly flows are considered as high interaction regimes, resulting in significantly enhanced heat and mass transfer rates by strong interactions between the gas and liquid phases [36-38]. Despite several experimental and modeling
studies, due to the complexity involved, estimations of regime transition boundary between pulsing and trickle flow are not sufficiently accurate [32]. Showing significant enhancement of reactor performance in pulsing flow, reactions in trickle and pulsing flows were conducted with various flow rates in different heights of catalyst beds for H$_2$O$_2$ decomposition [34] and with different catalyst positions in the bed for phenylacetylene hydrogenation [33]. In the later case, the reactor was modeled using mass balances with pulse-base separated mass transfer for pulsing flow operation [175].

Acetophenone (AP) hydrogenation is a typical three phase reaction which is composed of hydrogen gas and AP liquid on solid catalysts. As the simplest ketone having an aromatic ring, this reaction has complex reaction scheme which consists of competitive and continuous hydrogenations and hydrogenolysis of phenyl ethanol (PE) as shown in Figure 5.1.

Figure 5.1: The reaction scheme of acetophenone hydrogenation on Rh/Al$_2$O$_3$ catalyst: hydrogenation (solid arrow), hydrogenolysis (dashed arrow).
In our prior work, considering the reaction route composed of valuable products and faster reaction rate, the 1% Rh/Al$_2$O$_3$ catalyst with AP $\rightarrow$ PE $\rightarrow$ CE reaction route was selected and studied at 60-100 °C, 1.1-4.1 MPa $P_{H_2}$ and 0.04-0.4 M $C_{AP,0}$ using a slurry reactor as reported in chapter 4.

In this work, the effect of operation in high interaction regime on reactor performance was investigated with AP hydrogenation having complex reaction scheme and relatively high pressure for pharmaceutical applications. For this, to set proper reaction conditions for different flow regimes, hydrodynamic tests were first conducted for the regime transition boundary from low to high interaction regime. With results from reaction experiments, considering external mass transfers and wetting efficiency, a reactor model was developed to predict trickle bed reactor behavior in different flow regimes.

5.2 Experimental Setup

The hydrodynamic and reaction experiments were conducted in a trickle-bed reactor using a 7.1 mm ID x 25 cm length stainless steel tube to conserve hydrogen in 10-25 barg pressure and facilitate radial isothermality, connecting straight to a 300 ml liquid reservoir for liquid recycle, as shown in Figure 5.2.
Figure 5.2: Schematic diagram of the experimental setup. 1: fixed bed reactor, 2: reservoir, 3: back-pressure regulator, 4: mass flow controller, 5: piston pump, 6 and 7: pre-heaters, 8: quencher and condenser, PDI: pressure difference indicator, TI: temperature indicator.

The reactor tube included a jacket for heating medium to maintain target temperature and the liquid reservoir had cooling tube, quencher and condenser using cooling media to minimize liquid loss by gas vent. The heating and cooling medium were supplied by Thermo Scientific AC 200 and AC 150 circulators with A 25 refrigerator, respectively. Pressure drop by fixed bed in the reactor tube was measured by Omega DPG-100DWU and 005 DWU pressure differential indicators, recorded in a console box using Horner QX-351. Liquid in the reservoir was agitated by Chemglass Optichem digital hotplate stirrer. A 1/8” OD x 1 m length stainless steel tube with a jacket and 1 liter cylinder were used as the quencher and condenser, respectively. The reactor pressure was controlled by Equilibar EB1LF2 backpressure regulator. Brooks 5850E mass flow controllers and SSI
Prep-100 dual piston pump were used to supply gas and liquid to the reactor, respectively. Micromotion LF Series Coriolis flow meter and transducer were installed after the piston pump to measure and record liquid flow rate. Gas and liquid inlet flows were electrically preheated using two Omega CSI32K PID controllers. Thermocouples and pressure gauges were installed to measure temperature and pressure at several points including the reactor inlet and outlet.

5.3 Hydrodynamic tests

The purpose of the hydrodynamic study was to determine the reaction operation conditions for trickle (low interaction) and high interaction regimes, prior to conducting reaction experiments in different flow regimes. The regime transition depends on various factors including particle size, distribution, physical properties of fluids, gas and liquid superficial velocities [32]. Among these, the superficial velocities are commonly used for the manipulated variables [39, 40]. The detection methods of the regime transition include visual inspection [41, 42], pressure drop [40, 43], pressure drop fluctuation [44, 45], liquid holdup, electrical conductivity, CT (computed tomography), etc [32]. The most frequently used experimental conditions in the literature are air and water as operating fluids at atmospheric temperature and pressure in 1-2 inch diameter transparent tubes filled with 1-5 mm diameter particles, through visual inspection as the detection method. The superficial velocity ranges for gas and liquid are 0.03-0.3 m/s and 2-20 mm/s, respectively [32, 39]. In the reaction of this work, however, hydrogen and acetophenone in cyclohexane were introduced in a 7.1 mm ID stainless steel tube packed with 0.5 mm 1% Rh/Al₂O₃ spheres and the operating conditions were 10-25 barg pressure
and 60-100 °C temperature. Because the visual inspection could not be applied for the reaction system due to the opaque reactor material, an appropriate indirect detection method was required. Hydrodynamic studies under high pressure and temperature with hydrogen and organic liquid system are scarce. Therefore, hydrodynamic tests for regime transitions were conducted systematically, ranging from the frequently used conditions to the reaction conditions.

5.3.1 Experimental procedure

Transparent polymer tubes (6.5 mm ID x 25 cm; polycarbonate – PC or fluorinated ethylene propylene - FEP) or the stainless steel tube were filled with 20-30 mesh glass beads (0.6-0.8 mm diameters) or 0.5 mm diameter γ-alumina spheres. Nitrogen or hydrogen flowed with the target flow rate equivalent to 0.01-0.25 m/s linear superficial velocities, which were calculated with gas densities using the Redlich-Kwong equation of state modified by Soave [176]. When the pressure of the reactor system was stabilized at the target pressure, deionized water or 0.6 M acetophenone (AP, Fluka, >99%) in cyclohexane (CH, Sigma-Aldrich, >99.9% HPLC grade) as liquid phase was introduced to prewet the bed with 30 ml/min liquid flow rate (high interaction regime) for 15-30 minutes. The liquid flow was stopped for 30 seconds and set at 2.5 ml/min (trickle flow). Following this, the liquid flow rate increased with 3 or 3.5 ml/min increment up to 35 ml/min. Each case typically took 1-10 minutes to stabilize and then the pressure drop data was collected every second for 2 minutes to obtain the average and the standard deviation of pressure drops for each case, visually observing regime transition for transparent tubes. The maximum repeatability error was ± 10% and typically below 5%. In addition, video
clips for each flow regime were taken during the experiments by a borescope system, composed of a borescope (Gradient Lens Corporation Pro Superslim), CCD (Sony XCDU100) and halogen lamp (Edmund 21DC).

5.3.2 Experimental results and discussion

At the boundary between trickle to high interaction regime, the vigorous interaction between gas and liquid started at the bottom of the tube and grew to the top with increase of liquid flow rate, similar to reports for pulsing flow regime transition [42]. In the present work, only bubbly flow regime was observed with quickly passing gas bubbles in liquid instead of alternatively changing gas and liquid zones as in the pulsing flow regime, as confirmed by video clips in the high interaction regime. No pulsing flow was likely owing to relatively small tubes and particles used in this work.

Figure 5.3: Confirmation of indirect regime detection method with visual observation using FEP tube with 0.5 mm diameter γ-alumina spheres; 0.033 m/s H₂ superficial velocity, 0.6 M acetophenone in cyclohexane (liquid phase), 20 bar and 25 °C. (Stdev: standard deviation)
Figure 5.3 shows that standard deviation of pressure drop increased moderately in trickle flow and became relatively stable at higher value in bubbly flow, while the pressure drop kept increasing as reported for high pressure operation [177]. The stabilization point of pressure drop fluctuation (standard deviation of pressure drop) matched the regime transition point confirmed by visual observation. Therefore, pressure drop fluctuation was used as detection method of the regime transition for the opaque stainless steel tube in the present work.

Following the procedure, 26 regime transition experiments were conducted with different gases and liquids, pressures, tube and particle materials at 25 °C, as shown in Figure 5.4.

Figure 5.4: Regime boundary map including effects of five different variables at 25 °C. Legend – gases (N: Nitrogen, H: hydrogen)/liquids (W: water, O: acetophenone in cyclohexane), pressure (bar, $P_{\text{bar}}$), tube material (PC: polycarbonate, FEP: fluorinated ethylene propylene, SS: stainless steel), particle material (G: glass beads, A: $\gamma$-alumina).
Run (1) of Figure 5.4 as the starting point (conditions: $N_2$, water, 1 bar, PC tube, glass beads) was compared with prior works conducted with larger particles and similar operating conditions [43, 178]. Although smaller particles and tubes can allow more wetting and earlier inception of high interaction [32, 179], the transition points (11-12 mm/s) with 0.6-0.8 mm glass beads were not quite different from the prior works (7-12 mm/s) with 3 mm glass spheres [178]. For varying gas flow rates, the regime transitions in this work were either independent or became later with higher gas flow rates, while most data in pulsing flow were significantly affected and became earlier as reported in the literature [39]. This may be caused by characteristics of bubbly flow regime in the present work, different from pulsing flow regime in the prior works. As the next step, experiments with hydrogen at high pressure (25 barg) instead of nitrogen and atmospheric pressure were conducted as shown in (2) and (3) of Figure 5.4. Hydrogen as low density gas and high pressure can result in earlier and later boundary, respectively [178]. Although the results appear to follow the trend, it is not obvious because the differences caused by hydrogen and higher pressure were not significant within the experimental error range. Run (4) of Figure 5.4 showed similar boundary despite differences of particle size, bed voidage and particle materials between $\gamma$-alumina (0.5 mm $\phi$, 0.44 $\varepsilon_b$) and glass beads (0.6-0.8 mm $\phi$, 0.40 $\varepsilon_b$) having smaller bed voidage with larger diameters due to the particle size distribution. Significant effect of bed voidage, reported for pulsing flow transition [40, 180] may not apply for bubble flow regime transition or glass beads with particle size distribution, considering the relatively insignificant effect of particle size from the literature [181]. Different from ignorable effects in (1) – (4) of Figure 5.4, significant earlier boundary shift as shown in (5) and (6)
of Figure 5.4, occurred with the change from water to the organic liquid, as explained by earlier boundary with lower interfacial tension, 1/3 for cyclohexane than that of water [180]. As seen by comparing (2) and (3), (5) and (6), and (7) and (8), there was no pressure effect between 5 and 25 barg. Considering 2.0 kg/m³ hydrogen density at 25 barg and 25 °C, this feature was supported by the report indicating no pressure effect below 2.3 kg/m³ gas density [178]. With the indirect detection method described above, comparing (5) and (7), and (6) and (8), regime boundaries in the SS tube were similar to the FEP tube except for a little earlier boundary. This may be caused by relatively high hydrophilicity of SS tube (72° contact angle for SS [182]; 111.6° for FEP [183]), amplified by high wall surface to tube volume ratio of small diameter tubes used in this work. As the final step, 7 additional experiments at 80 °C were conducted. However, there was no difference between regime transitions at 25 °C and 80 °C, although later boundary was expected at higher temperature from literature [184].
Figure 5.5: Determination of liquid flow rates in low (Trickle) and high (Bubbly) interaction regimes for reaction operating conditions with 0.057 m/s hydrogen superficial velocity at 25 bar, 80 °C and 0.6 M C_{AP}. (Stdev: standard deviation)

From Figure 5.5 for 25 barg and 80 °C as the representative reaction operating condition, 2.5 mm/s (6 ml/min) and 12.5 mm/s (30 ml/min) liquid superficial velocities were selected for reactions in trickle flow and bubbly flow regimes, respectively. In this case, the regime transition boundary was 4.7 mm/s (12 ml/min), similar to the value at atmospheric temperature, (8) in Figure 5.4.

5.4 Reactions in low and high interaction regimes

As introduced in section 5.1, the acetophenone hydrogenation reaction tests were carried out in the tubular reactor system with different liquid linear velocities to assess the effect of high interaction regime, over the entire operating range.
5.4.1 Experimental procedure

The 1% rhodium on 0.5 mm diameter $\gamma$-alumina spheres were packed near the bottom position of the SS tube reactor, and filled with the same support without rhodium for the rest of the tube. For example, the reactor for 0.12 M $C_{AP,o}$ was packed in the following order: 1.2 g $\gamma$-alumina at the bottom, 0.3 g catalyst, followed by 7.8 g $\gamma$-alumina on the top part of the tube. First, hydrogen (99.999% Ultra High Purity) was flowed at the target gas flow rate, then the reactor pressure and temperature were increased to the target operating conditions using the back pressure regulator and heaters, respectively. After stabilization, 140 ml cyclohexane was placed in a 300 ml stirred liquid reservoir and circulated to the tubular reactor with 30 ml/min liquid flow rate for 30 minutes to wet the bed. Following this, the reaction was started by injection of 60 ml $C_{AP,o}$ acetophenone in cyclohexane into the liquid reservoir with magnetic agitation. During the reaction period, the reactor temperature was maintained by hot fluid circulation. To minimize liquid loss by gas flow, hydrogen effluent from the reactor was quenched in the liquid reservoir and condenser by -5 to -10 $^\circ$C cooling medium. The samples from 10 to 160 minutes reaction time were taken at the liquid reservoir and characterized by GC (Gas Chromatography, HP 5890 II) equipped with an Agilent DB-WAXetr capillary column (50 m x 0.32 mm) and flame ionization detector under the following conditions: helium as carrier gas (2.6 ml/min), inlet and detector temperatures 250 $^\circ$C, oven temperature 150 $^\circ$C and injection volume 0.5 $\mu$l with split. The $R^2$ values for all GC calibrations were >99.6%. The repeatability error for each sample was less than 5%, average 2.4% for all cases. The mass balance was checked after experiments and the error was verified to be below 5%.
5.4.2 Experiments in high and low interaction regimes

Following the above procedure, the reaction tests were carried out with 2.5 mm/s and 12 mm/s liquid linear velocities, corresponding to trickle and bubbly flow regime, respectively, over the entire AP concentration range as shown in Figure 5.6.

Figure 5.6: Concentration profiles with time for acetophenone hydrogenation in low (dash) and high (solid) interaction regimes: a) 0.6 M $C_{AP,0}$ with 3 gcat, b) 0.04 M $C_{AP,0}$ with 0.3 gcat; 25 bar, 80 °C
In the high interaction regime, the AP concentration with time decreased faster (blue arrow in Figure 5.6a) than in the low interaction regime, indicating that the reaction rate was enhanced. However, because at the early reaction time, the bottom temperature with 0.6 M $C_{AP,o}$ and 3 gcat was increased up to 10 °C, the effect of bubbly flow was amplified by the non-isothermal reactor condition. It is likely that the larger reaction heat from relatively higher $C_{AP,o}$ and more catalyst could not be sufficiently removed through the reactor wall. With lower 0.04 M $C_{AP,o}$ and lesser 0.3 gcat to remove the non-isothermal effect, Figure 5.6b showed smaller effect of the high interaction regime with stable bottom temperature during the entire reaction time. In addition, isothermal operation was also confirmed for 0.12 M $C_{AP,o}$ and 0.3 gcat. The high interaction regime effect in Figure 5.6b suggests that bubbly flow enhances the reaction rate for acetophenone hydrogenation. However, because the external mass transfer rate becomes larger with faster liquid flow, the enhancement level may be affected by the increase of liquid flow rate. Thus, to clarify the effect of different flow regimes, it is necessary to separate the external mass transfer effect.

5.4.3 Experiments for the external mass transfer effect

To minimize the effect of external mass transfer change caused by different liquid flow rates, three options can be considered. The first is to conduct additional tests in the intermediate liquid flow rates. In this case, when liquid flow rate is gradually increased from the trickle flow under otherwise identical operation conditions, the result would be a gradual increase of reaction rate. Near the regime boundary, however, a sudden increase of reaction rate by increased liquid flow rate may be expected. The second option is to
use a transparent tube with the catalyst in different positions, as was done in our prior work [33]. This option is, however, difficult to apply due to safety issue because most transparent tubes cannot survive at high temperature and pressure with hydrogen and organic mixture. The third possibility is to use induced pulsing as in our different prior work [185]. In this case, the trickle flow is changed to pulsing flow without increase of average liquid flow rate by the artificial liquid pulsing having specific interval. However, due to artificially induced pulsing, its hydrodynamic study has more complexity including the specific interval which affects reactor performance [186]. Thus, the first option was selected with experiments at three different intermediate liquid flow rates between the trickle and bubbly flow, as shown in Figure 5.7 and Table 5.1.

Figure 5.7: Acetophenone concentration profiles with 5 different liquid superficial velocities at 3 stdl/min H₂ flow rate; 80 °C, 25 barg and 0.12 M C_{AP.o} with 0.3 g 1% Rh/Al₂O₃ catalyst; in parenthesis – liquid volume flow rate, ml/min.
Table 5.1: Initial acetophenone hydrogenation rates (mmol/gcat.s) with different liquid flow rates ($\nu_{\text{liq.}}$) and reaction time (0-30 min.) at 80 °C, 25 barg, 0.12 M C_{AP,o} and 3 stdl/min H$_2$ flow rate. (T: trickle flow; B: bubbly flow)

<table>
<thead>
<tr>
<th>$\nu_{\text{liq.}}$ (cc/min)</th>
<th>Regime</th>
<th>Reaction time, min</th>
<th>mmol/gcat.s</th>
<th>Average</th>
<th>Difference</th>
<th>$\Delta \nu_{\text{liq.}}$</th>
<th>Ratio of $\nu_{\text{liq.}}$</th>
</tr>
</thead>
<tbody>
<tr>
<td>6</td>
<td>T</td>
<td>0</td>
<td>-0.0320</td>
<td></td>
<td>-0.0178</td>
<td></td>
<td></td>
</tr>
<tr>
<td>9</td>
<td>T</td>
<td>10</td>
<td>-0.0129</td>
<td></td>
<td>-0.0093</td>
<td>-0.0014</td>
<td>3</td>
</tr>
<tr>
<td>15</td>
<td>B</td>
<td>20</td>
<td>-0.0140</td>
<td></td>
<td>-0.0132</td>
<td>-0.0033</td>
<td>6</td>
</tr>
<tr>
<td>21</td>
<td>B</td>
<td>30</td>
<td>-0.0124</td>
<td></td>
<td>-0.0180</td>
<td>-0.0012</td>
<td>6</td>
</tr>
<tr>
<td>30</td>
<td>B</td>
<td>30</td>
<td>-0.0510</td>
<td></td>
<td>-0.0146</td>
<td>-0.0006</td>
<td>9</td>
</tr>
</tbody>
</table>
The liquid flow rates were increased sequentially to ~1.5 times of the previous liquid velocity. The initial acetophenone reaction rates in Table 5.1 were obtained for 0-30 min. reaction time by differentiation formula [142]. The rates over 0-30 min. were averaged and compared with results from different liquid flow rates. Figure 5.7 shows that the reaction rate becomes gradually faster with increase of liquid flow rate, except for the interval between 3.7 and 6.2 mm/s liquid superficial velocities, which include the regime transition boundary confirmed in section 5.3. In this interval, the reaction rate was sharply increased, double for the other intervals as shown in Table 5.1. Because this behavior cannot be explained without the regime transition effect, it may be concluded that the bubbly flow regime enhances reaction rates for acetophenone hydrogenation. In addition, comparing the result with the same 0.5 mm diameter catalyst in the slurry reactor from chapter 4 which is free from external mass transfer, it may be seen that the reaction rates in trickle bed reactor are significantly decreased by external mass transfer and can be improved by increased liquid flow rates and high interaction regime operation. For this case, the criteria for external gas-liquid and liquid-solid mass transfer limitations [137] were calculated using mass transfer correlations, introduced in the next section, as shown in Table 5.2.

<table>
<thead>
<tr>
<th></th>
<th>gas-liquid</th>
<th>liquid-solid</th>
</tr>
</thead>
<tbody>
<tr>
<td>hydrogen</td>
<td>hydrogen</td>
<td>ACPH</td>
</tr>
<tr>
<td>0.2087</td>
<td>0.8369</td>
<td>0.9236</td>
</tr>
</tbody>
</table>

Table 5.2: Criteria values to verify external mass transfer limitations with 3 stdl/min H₂ and 6 ml/min liquid at 80 °C, 25 barg and 0.12 M C₅₆esity.0.
Because all values are above 0.1 and values for liquid-solid limitations are larger than that for gas-liquid limitation, this reaction is confirmed as liquid limited reaction. In this case, higher wetting efficiency and mass transfer rate by high interaction regime operation can enhance reaction rates [187-189], explaining the enhancement in bubbly flow regime.

To investigate the effect of gas flow rates, the same experiments with 5 different liquid velocities were conducted at 1.5 and 4.5 stdl/min hydrogen flow rates (stdl: standard liter per minute at 25 °C and 1 bar) as shown in Figure 5.8.
Figure 5.8: Acetophenone concentration profiles with 5 different liquid superficial velocities: a) 1.5 stdl/min, b) 4.5 stdl/min H₂ flow rate; 80 °C, 25 barg, 0.12 M C_{AP,o} and 0.3 g 1% Rh/Al₂O₃ catalyst; in parenthesis – liquid volumetric flow rate, ml/min.
Similar enhancement by the high interaction regime was again confirmed at the regime boundary (3.7-6.2 mm/s). Such enhancements with different gas flows occurred at the same liquid superficial velocity range because regime transitions from trickling to bubbly flow do not depend on the gas flow rate as discussed in section 5.3 (Figure 5.4). In Figure 5.8, reaction rates decreased with higher gas flow rates. This can be more clearly seen by comparing the slurry reactor and trickle bed reactor results. At 1.5 stdl/min hydrogen flow, average reaction rates were 4.1% faster, while for 4.5 stdl/min were 3.8% slower, than for the case of 3 stdl/min. In general, faster gas superficial velocity results in increase of gas-liquid mass transfer rate, leading to faster reaction rate under mass transfer limitation. In this case, however, the effect of gas velocity was reverse, which may be explained by partial wetting because wetting efficiency is decreased by increase of gas velocity [187]. Considering liquid limited reaction, lower wetting efficiency in 4.5 stdl/min case can cause inefficient liquid contact to catalyst surface, resulting in slower reaction rate than the other cases.

At 10 barg and 100 °C, the same experiments with 5 different liquid velocities were carried out to obtain experimental features for modeling. For the 100 °C case, 0.24 g catalyst amount was used to compare with our prior work in the slurry reactor as reported in chapter 4.
Table 5.3 Initial acetophenone hydrogenation rates (mmol/gcat.s) with different liquid flow rates ($v_{\text{liq.}}$) and reaction time (10-40 min.) at 0.12 M C$_{\text{AP,o}}$. (T: trickle flow; B: bubbly flow)

<table>
<thead>
<tr>
<th>$v_{\text{liq.}}$ (cc/min)</th>
<th>Regime</th>
<th>average</th>
<th>difference</th>
<th>Regime</th>
<th>average</th>
<th>difference</th>
</tr>
</thead>
<tbody>
<tr>
<td>10 barg, 80 °C, 1.2 stdl/min and 0.3 g</td>
<td>T</td>
<td>-0.0114</td>
<td></td>
<td>T</td>
<td>-0.0207</td>
<td></td>
</tr>
<tr>
<td>9</td>
<td>T</td>
<td>-0.0124</td>
<td>-0.0010</td>
<td>T</td>
<td>-0.0218</td>
<td>-0.0011</td>
</tr>
<tr>
<td>15</td>
<td>B</td>
<td>-0.0135</td>
<td>-0.0011</td>
<td>B</td>
<td>-0.0254</td>
<td>-0.0035</td>
</tr>
<tr>
<td>21</td>
<td>B</td>
<td>-0.0138</td>
<td>-0.0004</td>
<td>B</td>
<td>-0.0259</td>
<td>-0.0005</td>
</tr>
<tr>
<td>30</td>
<td>B</td>
<td>-0.0142</td>
<td>-0.0004</td>
<td>B</td>
<td>-0.0266</td>
<td>-0.0007</td>
</tr>
<tr>
<td>25 barg, 100 °C, 3 stdl/min and 0.24 g</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Table 5.3 shows the same high interaction regime effect for both cases. However, in the 10 barg case, the differences by liquid velocities became smaller than all other cases, which make the average differences near the regime boundary less clear. This may occur because lower operating pressure decreases hydrogen solubility in the liquid phase and results in larger hydrogen limitation.

5.5 Modeling of the trickle bed reactor

In modeling the trickle bed reactor, all features from the experiments should be reflected in the model predictions. In our prior work [175], it was assumed that the catalyst surface was fully covered by liquid film, i.e. full wetting. In this assumption, hydrogen in gas region is transferred to catalyst surface through gas-liquid and liquid-solid mass transfer resistances, while only liquid-solid mass transfer resistance exists for liquid chemicals [46]. As a basic concept in reactor modeling, partial wetting should be considered to
describe slower reaction rate with larger gas flow rate based on prior literature that in liquid limited reaction, partial wetting results in lower overall effectiveness factor while it helps to improve reaction rates in gas limited reaction [189, 190]. A prior partial wetting model considered that dry surface can directly contact the gas phase without liquid film, while one part of the wet surface is affected by the stagnant liquid in the space between solids [47]. This three zone approach is, however, too complicated including the difficulty to obtain overall effectiveness factor for the dry zone. For high wetting efficiency, the dry zone is often ignored [48]. In the present work, the correlation for wetting efficiency [187] was modified by ignoring the effect of liquid flow rate, accounting only for the gas flow rate effect. It is possible that liquid flow rate effect on wetting is already included in the liquid-solid mass transfer coefficient selected in the present work because, when obtaining mass transfer rate experimentally, it is difficult to separate wetting efficiency and mass transfer. For high interaction regime, it has been reported that wetting efficiency and mass transfer rate are increased as a result of high interaction between gas and liquid, resulting in faster reaction rate for liquid limited reaction [187, 188]. The enhancement was considered by one adjustable parameter, $C$, because, for 3/8” tube with 0.5 mm sphere particles, there is no mass transfer correlation for the high interaction regime. The proper fitted adjustable parameter can explain enhanced reaction rate for liquid limited reactions by better wetting efficiency and mass transfer in the high interaction regime. Thus, to account for the partial wetting and mass transfer in different flow regimes, the effective wetting efficiency, $f$ was applied as

$$ f = 0.75 g^{-0.083} + C. $$
For reactor modeling, we applied the assumptions: 1) The liquid is in ideal plug-flow; 2) The tube reactor is isothermal; 3) Vaporization of organic species is negligible. Based on these assumptions and with the factor, \( f \), we obtained mass balances for the tubular reactor:

\[
\frac{dC_{AP,r}}{dz} = -\frac{A_f}{v_{liq.}} a_{s,k_{L,AP}} (C_{AP,r} - C_{AP,s})
\]

\[
\frac{dC_{PE,r}}{dz} = -\frac{A_f}{v_{liq.}} a_{s,k_{L,PE}} (C_{PE,r} - C_{PE,s})
\]

\[
\frac{dC_{CMK,r}}{dz} = -\frac{A_r f}{v_{liq.}} a_{s,k_{L,CMK}} (C_{CMK,r} - C_{CMK,s})
\]

\[
\frac{dC_{CE,r}}{dz} = -\frac{A_r f}{v_{liq.}} a_{s,k_{L,CE}} (C_{CE,r} - C_{CE,s})
\]

\[
(5.1)
\]

\[
\frac{dC_{H_2,r}}{dz} = -\frac{A_f}{v_{liq.}} a_{s,k_{GL,H_2}} (C_{H_2,r} - C_{H_2,s})
\]

\[
C_{j,r} = C_{j,t} \text{ at } z = 0
\]

\[
a_{s,k_{L,AP}} (C_{AP,r} - C_{AP,s}) = \frac{m_{cut} F_c}{V_r} (\eta_{d1} r_{d1} + \eta_{i1} r_{i1})
\]

\[
a_{s,k_{L,PE}} (C_{PE,r} - C_{PE,s}) = \frac{m_{cut} F_c}{V_r} (\eta_{r2} r_{r2} - \eta_{d1} r_{d1})
\]

\[
a_{s,k_{L,CMK}} (C_{CMK,r} - C_{CMK,s}) = \frac{m_{cut} F_c}{V_r} (\eta_{d2} r_{d2} - \eta_{d1} r_{d1})
\]

\[
- a_{s,k_{L,CE}} (C_{CE,r} - C_{CE,s}) = \frac{m_{cut} F_c}{V_r} (\eta_{2} r_{2} + \eta_{d2} r_{d2})
\]

\[
ak_{GL,H_2} (C_{H_2,r} - C_{H_2,s}) = \frac{m_{cut} F_c}{V_r} (3 \eta_{1} r_{1} + \eta_{d1} r_{d1} + 3 \eta_{2} r_{2} + \eta_{d2} r_{d2})
\]

\[
\frac{1}{ak_{GL,H_2}} = \frac{1}{a_{GL} k_{G,H_2}} + \frac{1}{a_{s} k_{L,H_2}}
\]

\[
a_{s} = \frac{6(1 - \epsilon_b)}{d_p}
\]

\[
(5.2)
\]

and for the liquid reservoir:

\[
\frac{dC_{j,t}}{dt} = \frac{v_{liq.}}{V_t} (C_{j,r(at = L)} - C_{j,t})
\]

\[
C_{j,t} = C_{j,0} \text{ at } t = 0
\]

\[
(5.4)
\]
The reactor model is based on our prior work [175], accounting for external mass transfer limitations for gas-liquid and liquid-solid films. Because of fully liquid recycled system, species concentrations in the liquid reservoir change with time. Thus, it was additionally assumed that the tubular reactor is in pseudo steady-state at each time step of the liquid reservoir, removing time derivative terms in the tubular reactor model. Connecting both models, the behavior of the reaction system was predicted. The adjustable parameter, $C$, was obtained by fitting with experiments at 5 different liquid flow rates, at 80 °C, 25 barg and 3 stdl/min hydrogen flow rate. The obtained values for $C$ were -0.1 for trickle flow and 0 for bubbly flow. The correlations of Yoshikawa et al. [191] and Metaxas and Papayannakos [162] were used in this work for liquid-solid mass transfer and gas-liquid mass transfer, respectively. Because most correlations were developed in typical cases including 1” tube and 3 mm catalysts, there are few correlations available for small tube and catalyst diameters in this work (3/8” tube and 0.5 mm catalysts). Among these, that of Yoshikawa et al. with 0.5 mm particles was remarkably the only one which provided the mass transfer coefficient less than that for the slurry reactor. This limit should be satisfied because it is unreasonable if the external mass transfer coefficient for the trickle bed reactor exceeds that for the slurry reactor which is free from external mass transfer limitations, as confirmed in the experiments. For gas-liquid mass transfer, most correlations for the typical system provided much smaller values implying severe gas limitations. The one by Metaxas and Papayannakos was obtained for benzene hydrogenation with 0.25 mm catalyst particles and they reported that, upon comparison between their work and other correlations, gas mass transfer in hydrogenation was faster [162]. The underlying reason is that there are relatively few gas-liquid mass transfer
studies with hydrogen and organic liquids, for small size catalyst particles. With the selected correlations, the reactor model was numerically solved, including the intrinsic reaction kinetics and the internal pore diffusion calculation module, reported in chapter 4. The model provided good predictions for the experiments, explaining features from the experiments well, as shown in Figure 5.9.
Figure 5.9: Comparison between experiments (symbols, t: trickle flow) and model predictions in trickle (dotted curve) and bubbly flow (solid curve) regimes with 0.12 M $C_{AP,0}$: a) 80 °C, 25 barg, 3 stdl/min $H_2$ and 0.3 gcat; b) 80 °C, 25 barg, 1.5 stdl/min $H_2$ and 0.3 gcat; c) 100 °C, 25 barg, 3 stdl/min $H_2$ and 0.24 gcat; d) 80 °C, 10 barg, 1.2 stdl/min $H_2$ and 0.3 gcat.
The average $R^2$ value for all 25 experiments was >95%, suggesting that the modeling approach in the present work is sufficiently accurate to predict the trickle bed reactor performance for acetophenone hydrogenation well in both high and low interaction regimes.

5.6 Concluding remarks

Acetophenone hydrogenation on 1% Rh/Al$_2$O$_3$ catalyst was conducted in a fully liquid recycled trickle bed reactor with 7.1 mm ID stainless steel tube and 0.5 mm diameter catalyst spheres, confirming different flow regime effects in 0.02-0.19 m/s and 2.5-12 mm/s range for gas and liquid superficial velocities under 80-100 °C, 10-25 barg and 0.04-0.6 M $\text{C}_{\text{AP},0}$ operating conditions. From hydrodynamic tests for different flow regimes, trickle and bubbly flow regimes were confirmed visually using similar diameter transparent tubes, systematically from nitrogen and water with glass beads to hydrogen and the reaction feed with γ-alumina spheres in 0.01-0.25 m/s and 0.7-15 mm/s for gas and liquid superficial velocities, respectively, at 25-80 °C and 1-26 bar range, providing the regime transition map. Using transparent tubes, variation of pressure drop fluctuations was shown to follow the visual observations for transitions between the trickling and bubbly flow regimes, thus the former was utilized for the opaque stainless steel reactor. Using operating conditions in the two flow regimes, it was confirmed that the operation in bubbly flow regime enhanced reaction rate in the above ranges for the liquid limited reaction. The effects of partial wetting and gas limited reaction were also confirmed by slower reactions for high gas flow rate and smaller regime transition effect for lower pressure, respectively. In the reactor model, an adjustable parameter accounting for
partial wetting and regime transition effect was included, and external mass transfer was considered along with internal pore diffusion using TOF based intrinsic kinetics reported in chapter 4. The developed reactor model provided good match with experiments, including the regime transition effect. The present work provides an example of combined experimental and modeling study, considering complicated hydrodynamics with regime transitions and complicated reactions for pharmaceutical applications.

Note: Adapted with the manuscript (S. Lee, N. Zaborenko and A. Varma, “Acetophenone hydrogenation on Rh/Al₂O₃ catalyst: Flow regime effect and trickle bed reactor modeling,” In the final stage of submission.)
CHAPTER 6. CONCLUSIONS AND RECOMMENDATIONS FOR FUTURE WORK

6.1 Summary

Experimental and modeling studies of multiphase reactors beyond previous works reported in the literature were conducted. Biphasic stirred tank and trickle bed reactors with n-butyraldehyde aldol condensation and acetophenone hydrogenation, respectively were selected and investigated as representatives of multiphase reactors. In general, this work showed two examples to develop reactor models for multiphase reactors, considering mass transfer effects on overall reaction rates. Specifically, the main conclusions of the present work are as follows.

6.1.1 Intrinsic reaction kinetic study in the mass transfer regime

Intrinsic reaction kinetics for base catalyzed nBAL aldol condensation was investigated under industrial operating conditions which are in mass transfer controlled regime. The reaction kinetics were obtained in the mass transfer regime dominated by reaction rate in the film using a stirred cell, confirming the mass transfer regime by plateau region tests and Hatta number calculations. Reasonable parameter values for the intrinsic reaction kinetics were obtained, with sensitivity studies for calculated properties of solubility and diffusivity considering salt effect and temperature dependency. It is noteworthy that this work is the first to obtain intrinsic reaction kinetics in the specific mass transfer regime.
The approach developed in this work will be useful for the other cases where it is difficult to obtain intrinsic reaction kinetics in the reaction controlled regime.

6.1.2 Biphasic stirred tank reactor modeling with interfacial area study

The biphasic stirred tank reactor model for nBAL aldol condensation based on two film theory was developed. Interfacial area in the tank was measured using a borescope system as the most reliable in-situ method, enabling measurements at high temperature and sodium hydroxide concentrations in a 300 ml lab-scale tank reactor. From the measurements, it was found that salt and temperature effects on interfacial area are significant, which draws reverse trends from conventional Weber number based correlations. A new semi-empirical correlation was developed with the viscosity correction term included to incorporate salt and temperature effects. Combining the interfacial area correlation and the intrinsic reaction kinetics from this work, the new reactor model for batch operation of the biphasic reactor provided good predictions for experimental data. In addition, it was shown that the accuracy of interfacial area estimation is a key factor to improve the model prediction quality for biphasic stirred tank reactors.

6.1.3 Intrinsic reaction kinetics and pore diffusion effects in a slurry reactor

To develop a model for trickle bed reactors, intrinsic reaction kinetics and pore diffusion effects were investigated using a slurry reactor. From screening tests, acetophenone hydrogenation on Rh/Al₂O₃ catalyst particles was selected to study flow regime effects of trickle bed reactors for pharmaceutical applications. The intrinsic kinetics based on
Langmuir-Hinshelwood mechanism with dissociative, non-competitive hydrogen adsorption and saturated surface assumption for organics provided the best fit with experiments from a slurry reactor with powder catalysts, relating the reliability of the obtained model with prior reports. From experiments with two different sized catalyst particles, significant pore diffusion effects were evaluated and modeled with boundary value ordinary differential equation sets. Catalyst characterization for metal dispersions and eggshell distributed rhodium contents allowed to apply turnover frequency based intrinsic reaction kinetics and pore diffusion effect modeling along with the kinetics.

6.1.4 Flow regime effects on trickle bed reactor performance and its modeling

Bubbly flow regime effects on reactor performance were investigated for a trickle bed reactor for acetophenone hydrogenation on 0.5 mm diameter Rh/Al₂O₃ particles in relatively high pressure operation for pharmaceutical applications. Hydrodynamic study for flow regime boundary was carried out from atmospheric conditions with glass beads in transparent tubes to the reaction conditions with alumina in stainless steel tubes, providing flow regime map for trickle and bubbly flow regimes. Reaction experiments in bubbly flow regime showed that, with increase of liquid-solid mass transfer rates, bubbly flow regime operation enhanced reaction rates. In addition, partial wetting in high gas flow rates decreased reaction rates under liquid limited reaction. Combining intrinsic reaction kinetics and pore diffusion models, the trickle-bed reactor model with a new correction factor for partial wetting and regime transition provided good predictions for experiments, explaining flow regime and partial wetting effects well.
6.2 Recommendations for future work

6.2.1 Drop size distributions

The results of biphasic stirred tank reactor modeling showed the need for more accurate interfacial area values to improve reactor models. Because interfacial area is obtained from actual drop size distributions, a study on drop size distribution is a good next step. Population balance equation (PBE) is a suitable method to develop drop size distributions. This method is composed of a number density equation for each drop size range with source terms for birth and death of the drops in the specific range by drop breakage and coalescence [30]. Since appropriate approaches for source terms were introduced, various breakage and coalescence have been suggested to improve PBE modeling [22]. In these prior works, convective term was often ignored to simplify calculation of PBE, obtaining ordinary equation set. The convective term, however, should be included to consider different turbulence intensities and flow rates according to the location inside a stirred tank. This part, convective term, can be considered by computational fluid dynamics techniques (CFD). To consider 3D multiphase turbulent flow field, CFD has been improved by suggesting new models or source terms including multiphase interactions for turbulent and flow field estimations [192]. This is necessary for more accurate scale up of stirred tank reactors [193]. Although its calculation cost becomes expensive, PBE with CFD, as the most rigorous model for drop size distributions, may provide good insight to improve interfacial area estimation methods with well-developed source terms for drop breakage and coalescence.
6.2.2 New applications

In the biphasic stirred tank study in the present work, nBAL aldol condensation was used as a model reaction which was a simple self-aldol condensation with 1st order reaction based on pseudo steady state approximation. With aldol condensation, a biphasic stirred tank reactor was successfully investigated in terms of intrinsic reaction kinetics under the mass transfer controlled regime and reactor modeling with interfacial area study. To apply this study to general biphasic reaction which may be more complex, there is, however, a need to expand this reactor study with new applications which are more complicated including higher order reaction, heavier feed chemical and complex reaction scheme. For this, hexanal aldol condensation, azocoupling and the Suzuki coupling can be new applications for biphasic stirred tank reactor study, which are well-known reactions in pharmaceutical and fine chemical fields. Hexanal aldol condensation as higher aldehyde may allow to study on different property effects including solubility, interfacial tension and diffusivity [53]. Azocoupling and Suzuki coupling may provide an opportunity to study complex reactions in terms of reaction order, feed chemical and reaction schemes [194-196].

6.2.3 Mass transfer study with small particles and hydrogen

External mass transfer of trickle bed reactors consists of gas-liquid, liquid-solid and gas-solid mass transfers. Among these, gas-liquid mass transfer depends on hydrodynamic properties including tube and particle diameters, flow rates, viscosity and densities of fluids [32]. Most gas-liquid mass transfer correlations from literature are for 1-3 mm diameter particles and air or nitrogen as gas, which are inappropriate for hydrogenations
with small size catalyst particles [197, 198]. Small particle sizes are usual for pharmaceutical applications to obtain high selectivity of reaction intermediates as the desired product. Although a correlation using 0.25 mm particles was used for 0.5 mm particles for the present work [162], it would be important to thoroughly study mass transfer for 0.1-1 mm particle size range with hydrogen, and comparing results with the literature.
APPENDICES
APPENDIX A. EXPERIMENTS TO VERIFY 2EHEL EFFECT ON THE SOLUBILITY OF NBAL IN WATER

To verify that 2EHEL decreases the solubility of nBAL in water, some simple experiments were conducted. At reference temperature, the mixture of nBAL/2EHEL was prepared and then water was added. After vigorous mixing and archiving equilibrium, the organic phase samples before and after adding water were characterized by GC and the results were compared with estimates by the NRTL/UNIQUAC methods. The results from the experiments were near the estimates considering 2EHEL effect (Figure 2.2). Therefore, the tertiary solubility estimation method, including the effect of 2EHEL which decreases the solubility of nBAL, was verified well.
APPENDIX B.  ESTIMATION OF DIFFUSIVITY, $D_{AB}$ AND MASS TRANSFER COEFFICIENT, $K_L$

Diffusivity is typically a function of viscosity and temperature. Based on the Stokes-Einstein equation, many modified correlations have been published [199]. Among these, three correlations were considered because there is none that is widely accepted. The Wilke and Chang (1955) [71] correlation is the oldest one but it is often used due to its acceptable accuracy:

$$D_{AB} = \frac{2.02 \times 10^{-12} \phi^{0.5} T}{\mu_B} \quad (B.1)$$

This correlation includes the association number $\phi$ (2.6 for water at 20°C) which can cause uncertainty in extrapolation. Subsequently, two other correlations have been recommended in the literature as alternatives to estimate diffusivity. One is by Tyn and Calus (1975) [72]:

$$D_{AB} = 8.93 \times 10^{-12} \frac{v_{BB}^{0.267} T}{v_{bA}^{0.433} \mu_B} \left( \frac{\sigma_B}{\sigma_A} \right)^{0.15} \quad (B.2)$$

$$\sigma = \frac{P_c^2 \tau_c^3 (0.132 \alpha_c - 0.279)(1 - T_{br})^{11/9}}{1000}, \alpha_c = 0.9076 \left[ 1 + \frac{T_{br} \ln \left( \frac{P_c}{1.013} \right)}{1 - T_{br}} \right]$$

$$T_{br} = T_b / T_c$$
This correlation does not use association number $\emptyset$, but needs interfacial tension ratio, which was estimated by the Brock and Bird corresponding states method [199]. The other is the correlation by Hayduk and Minhas (1982) [73]:

$$D_{AB} = 1.59 \times 10^{-13} T^{1.52} \mu_B^{-1.02} \quad (B.3)$$

which has a larger temperature dependence than the other correlations.

The viscosity of NaOH solution for diffusivity estimate was calculated using the correlation by Laliberte (2007) [98]:

$$\mu_{aq} = \mu_w^{w_w} \cdot \Pi \mu_i^{w_i} \cdot \mu_w = \frac{T-27.15}{(0.05594(T-273.15)+5.2842)(T-273.15)+137.37} \quad (B.4)$$

$$\ln \mu_{NaOH} = \frac{440.2(1-w_w)^{0.008976}-423.67}{(0.015949(T-273.15)(107.6(1-w_w)^{4.6489}+1)} \quad (B.5)$$

The viscosities for nBAL and 2EHEL were estimated using the DIPPR (Design Institute for Physical Properties: AIChE) liquid viscosity model from ASPEN Properties V7.3.

$$k_{L,aq} = 0.00325 \left( \frac{(P/V_c)\mu_{aq}}{\rho_{aq}} \right)^{0.25} \left( \frac{\mu_{aq}}{\rho_{aq} d_{nBAL,aq}} \right)^{-2/3} \quad (B.6)$$
The mass transfer coefficient, $k_L$ was calculated using the Calderbank and Moo-Young (1961) [74] correlation. The power number, $Po$ used for four blade 45°-pitched paddle was 1.15, obtained from the 3D-CFD calculation (Appendix C).

The density of NaOH solution for mass transfer coefficient evaluation was calculated using the correlation of Laliberte and Cooper (2004) [200]:

$$\rho_{aq} = \frac{1}{\frac{w_W}{\rho_W} + \sum \frac{w_i}{\rho_i}} \rho_{NaOH} = \frac{(385.55w_{NaOH}+753.47)e^{(0.000001(T+269.73)} \frac{0.10938+0.0006953(T+273.15)}{w_{NaOH}}}{w_{NaOH}} \quad \quad (B.7)$$

The other densities were estimated using DIPPR liquid density model from ASPEN Properties V7.3.
APPENDIX C. 3D-CFD STUDY FOR POWER NUMBER AND INTERFACE STABILITY

Three dimensional computational fluid dynamics (CFD) technique was used to calculate the power number for the agitator of the 300ml stirred cell and for comparison with experiment data for stability of the interface between the organic and aqueous phases using the ANSYS FLUENT 12.1 (2009) [201]. The power number for 4-blade 45° pitched paddle is in the range 0.5~2.5 [202] but it is difficult to find or obtain the exact value for the small impeller \(d_{\text{imp}} = 0.035 \text{ m}\). Also, according to literature of the multiphase simulation technique, simulation on the interface of biphasic systems is possible even though some authors have reported that this technique is not sufficient to estimate the velocity vector field inside the agitated vessel [7, 203].

The mesh for the 300ml stirred cell was prepared using GAMBIT 2.4.6 (2004). The mesh was primarily hexahedral but tetrahedral mesh was also used around some parts of the cooling tube. The total mesh number was 361,000. The models used for this CFD study were RANS (Reynolds Averaged Navier Stokes), standard k-ε with standard wall function for turbulence, MRF (multiple reference frame) for impeller, Euler-Eulerian for multiphase model and continuity equation [201]. The calculations were made for the same values as in the experimental study; organic phase: 24 ml, aqueous phase: 180 ml.
The density and viscosity for each phase were obtained using ASPEN properties V7.3 at 120 °C.

Using the CFD technique, the momentum around the agitator can be evaluated, and then the power and power number can be calculated. The $P/V_c$ (W/m$^3$) values are shown in Table C.1.

<table>
<thead>
<tr>
<th>rpm (1/min)</th>
<th>$P/V_c$ (W/m$^3$)</th>
</tr>
</thead>
<tbody>
<tr>
<td>40</td>
<td>0.108</td>
</tr>
<tr>
<td>50</td>
<td>0.208</td>
</tr>
<tr>
<td>60</td>
<td>0.355</td>
</tr>
<tr>
<td>70</td>
<td>0.558</td>
</tr>
<tr>
<td>80</td>
<td>0.826</td>
</tr>
</tbody>
</table>

Through the simulations, the power number, $Po$ was about 1.15 under all operating conditions for the agitator used in the stirred cell. The stability of the interface between organic and aqueous phases was also studied, as shown in Figure C.1. These results were similar to the experimental observations, as described in section 2.4.1.
Figure C.1: The aqueous phase volume fraction contours using 3D-CFD study for the interface in the stirred cell: a) 50 rpm, b) 100 rpm, c) 150 rpm; Aqueous phase (light gray), Organic phase (dark gray)
APPENDIX D. MASS TRANSFER COEFFICIENT FOR THE DISPERSED PHASE

The dispersed phase in a stirred tank forms droplets inside the tank. The mass transfer coefficient for the dispersed phase is affected by the movements inside a droplet. The movements can be classified as a rigid drop having no movement, a laminar circulating drop and a turbulent circulating drop. According to the movement in the drop, different correlations have been developed [204]. To determine the type of drop movement, the diameter number ($d^*$) can be used as shown in Eq. D.1 [205].

\[
d^* = d_{32} \left( \frac{\mu_a^2}{\rho_{aq} g (\rho_{aq} - \rho_{org})} \right)^{-1/3}
\]  

(D.1)

The diameter numbers in this study were 16±3. With the assumption that the droplets are clean, the droplet movement in this study is evaluated as a laminar circulating drop. For this case, Kronig and Brink [206] developed the generally accepted relation as Eq. D.2.

\[
k_L = \frac{17.9 D}{d_{32}}
\]  

(D.2)
Johnson and Hamielec [207] experimentally confirmed that this equation provides good estimations when the Reynolds number is below 70 for the case of n-butanol and cyclohexanol. The Reynolds number range in the present study was 51±13, with the terminal velocity calculated from the correlation (Eq. D.3) by Hu and Kintner [208].

\[
  u_t = \sqrt{\frac{4gd_2(\rho_{aq}-\rho_{org})}{3\rho_{aq}C_D}} \tag{D.3}
\]

In this study, therefore, Eq. D.2 was used for the mass transfer coefficient for the dispersed phase. The Sauter mean diameter and the averaged value of nBAL and 2EHEL were used for the drop diameter and the physical properties for the organic phase, respectively.
APPENDIX E. ESTIMATION OF THE INTERFACIAL TENSION

The interfacial tension between organic and aqueous phases has a significant role in predicting the interfacial area through stability of the organic system as the dispersed phase in this study. There are various empirical correlations to predict the interfacial tension between two liquid phases. Among these, Good and Elbing [209] suggested Eq. E.1 to estimate the interfacial tension from the surface tensions of the two phases (1 – organic, 2 – aqueous).

\[
\sigma_{12} = \sigma_1 + \sigma_2 - 2\Phi_{12}\sqrt{\sigma_1\sigma_2}
\]  

(E.1)

\(\Phi_{12}\) is an interaction parameter which can be obtained from dipole moments using the figure available in Good and Elbing [209], and the dipole moments for nBAL and 2EHEL obtained from Yaws [210], the interaction parameter averaged for nBAL and 2EHEL was 0.85.

The continuous aqueous phase in this study contains significant amounts of NaOH which affect the surface tension of water. Dutcher et al. [211] provide a correlation to estimate the surface tension of inorganic multicomponent aqueous electrolyte solutions, including NaOH, which leads to Eq. E.2.
\[ \sigma_{aq} = \sigma_w + x_{NaOH}(a_{ws} + b_{ws}T) \{ a_{ws} = 69.691, b_{ws} = 0 \text{ for } NaOH \} \quad (E.2) \]

\[ \sigma_w = 235.8 \left( \frac{647.15 - T}{647.15} \right)^{1.256} \left( 1 - 0.625 \frac{647.15 - T}{647.15} \right) \]

\( x_{NaOH} \) is the mole fraction of NaOH in the aqueous phase. The estimation error for the surface tension of NaOH solution was reported to be 1.3%. The surface tension of water \( \sigma_w \) was obtained from Vargaftik et al. [212]. For the surface tension of the organic phase, surface tensions of nBAL and 2EHEL were obtained from Yaws [210] and were averaged with their mass fractions as described by Eq. E.3.

\[ \sigma_{org} = \sum_i w_i \sigma_i ; \sigma_{nBAL} = 69.216 \left( 1 - \frac{T}{525} \right)^{1.222} ; \sigma_{2EHEL} = 69.991 \left( 1 - \frac{T}{630} \right)^{1.222} \quad (E.3) \]
APPENDIX F.  3D-CFD STUDY FOR P/V AND THE DISTURBANCE BY THE BORESCOPE

A three-dimensional computational fluid dynamics (CFD) technique using ANSYS FLUENT 14.5 [213] was employed to check the disturbance by the borescope and to calculate the P/V (power per volume) of the 300ml stirred tank for calculation of the continuous phase mass transfer coefficient.

The flow disturbance by the borescope was checked through calculating the change of P/V by inserting the borescope inside the reactor. For this purpose, two meshes for the 300 ml stirred tank were prepared for the cases with or without the borescope using GAMBIT 2.4.6. Several possible cases could be evaluated because the borescope was located at the top and bottom positions and two different sized sight tubes (1/8” and 3/16”) were also used. Among these, the bottom position with the larger 3/16” sight tube was meshed (Figure F.1) because it was expected to provide the largest effect on the reactor hydrodynamics.
Figure F.1: The mesh inside the 300ml stirred tank reactor: the borescope (1), the agitator (2), the sample tube (3), injection nozzle (4), thermo couple (5) and the cooling tube (6).

The mesh was primarily hexahedral but tetrahedral mesh was also used around some parts of the cooling tube. The total mesh number was 306,160/306,856 with or without the borescope, respectively. The models used for the CFD study were RANS (Reynolds Averaged Navier Stokes), standard k-ε with standard wall function for turbulence, MRF (multiple reference frame) for impeller, Euler-Eulerian for multiphase model and continuity equation [213]. The calculations were made for the same values as in the experimental study; organic phase: 60 ml, aqueous phase: 140 ml.

Using the CFD technique, the momentum around the agitator was evaluated, and then the power was calculated (Table F.1).
Table F.1: Comparison of P/V (W/m³) with or without the borescope

<table>
<thead>
<tr>
<th>rpm (min⁻¹)</th>
<th>with borescope</th>
<th>without borescope</th>
<th>difference</th>
</tr>
</thead>
<tbody>
<tr>
<td>600</td>
<td>279</td>
<td>276</td>
<td>0.8%</td>
</tr>
<tr>
<td>1000</td>
<td>1289</td>
<td>1278</td>
<td>0.9%</td>
</tr>
</tbody>
</table>

The agitation power with the borescope was slightly increased but, the difference with cases without the borescope was < 1%. This implies that the effect of the borescope on the hydrodynamics in the reactor is negligible. This may be because there were various tubes in the reactor for sampling, injection, cooling, etc., which already served as baffles. Therefore, the borescope, with similar diameter as the other tubes, did not significantly affect the hydrodynamics.

For the calculation of the continuous phase mass transfer coefficient, P/V as a function of rpm was calculated by the 3D-CFD model as shown in Table F.2.

Table F.2: Conversion of agitation speed to P/V

<table>
<thead>
<tr>
<th>rpm (min⁻¹)</th>
<th>P/V (W/m³)</th>
</tr>
</thead>
<tbody>
<tr>
<td>600</td>
<td>279</td>
</tr>
<tr>
<td>700</td>
<td>431</td>
</tr>
<tr>
<td>800</td>
<td>638</td>
</tr>
<tr>
<td>900</td>
<td>901</td>
</tr>
<tr>
<td>1000</td>
<td>1289</td>
</tr>
</tbody>
</table>

Through the simulations, the power number, Po was about 1.22 under all operating conditions for the agitator used in the stirred tank.
APPENDIX G. NBAL SOLUBILITY IN THE REACTION TEMPERATURE 
AND THE ATMOSPHERIC TEMPERATURE

The different nBAL solubility in the reaction temperature and the atmospheric temperature can cause discrepancy between the actual conversion in the reactor and the apparent conversion from the GC. This is possible because the sample taken from the reactor at 80-140 °C is cooled to the atmospheric temperature and separated to the organic and aqueous phase. At this time, some of the organics exist in the aqueous phase according to their mutual solubility. The organic phase is composed of nBAL and 2EHEL and their solubility is significantly different (5.5% nBAL in water at 30 °C; 0.11% 2EHEL in water at 30 °C; data from Stephenson [66]). In addition, nBAL solubility increases with the decrease of the temperature. At atmospheric temperature, more nBAL is dissolved in water than 2EHEL. As a result, the nBAL ratio of the organic phase can decrease from the actual nBAL ratio. Therefore, the nBAL conversion from the GC characterization using the organic sample may be different from the actual conversion. To confirm the discrepancy quantitatively, the UNIQUAC (Universal Quasi Chemical Activity Coefficient) [65] activity model with the mutual solubility data for nBAL/water and 2EHEL/water by Stephenson [66] was used. The detail of this modeling and the experimental verification is in Appendix A. Considering the nBAL solubility in water for the liquid-liquid biphasic system of the nBAL/2EHEL/water using this solubility model, the apparent nBAL conversions from the organic sample were calculated according to the
actual conversions from 0 to 100%. 0.1 and 0.9 of the organic phase ratio cases were also calculated to compare with 0.3 of the organic phase ratio which was applied in this study as shown in Figure G.1.

![Figure G.1: Discrepancy between the apparent and actual conversions according to the organic phase fraction (Legend).](image)

From the results, the conversion discrepancies increased from 0% to 50% of the actual nBAL conversion while the discrepancies from 50% to 100% were decreased. This is because 2EHEL converted from nBAL by the reaction suppress nBAL to dissolve into water. In the case of 0.1 of the organic phase ratio, the discrepancy may be very serious. The nBAL component ratio in the organic phase is more easily changed than the other cases because the organic phase volume is smaller while the dissolved nBAL in water is larger due to a higher water volume. On the other hand, the maximum discrepancy in 0.3
of the organic phase ratio was 3.3% at 50% of the actual conversion. Therefore, through this preliminary study, it was revealed that the apparent conversion by the GC with the organic sample can be used as the actual nBAL conversion in case of 0.3 of the organic phase ratio.
APPENDIX H. COMPARISON AMONG STATISTICAL CORRELATIONS FROM DOES AND THE ANALYTICAL MODEL USED IN THE STIRRED TANK REACTOR STUDY

Multivariable linear regression techniques, as a statistical correlation, can be useful to estimate reactor performance with less effort, despite a decrease in understanding of the reactor behavior relative to analytical modeling [214]. The multivariable linear regression basically used in this section is:

\[ k_{overall} = \beta_0 + \beta_1 A + \beta_2 B + \beta_3 C + \beta_4 A^2 + \beta_5 B^2 + \beta_6 C^2 + \beta_7 AB + \beta_8 AC + \beta_9 BC (H.1) \]

\[ \{A: Temperature (°C); B: C_{NaOH}; C: rpm; \beta_i: empirical parameters\} \]

As 2\textsuperscript{nd} order partial least square regression, this regression can consider nonlinearity and interaction of multivariable, limitedly. Because each DOE set provides a statistical correlation, two statistical correlations using the regression were obtained from DOE 1 and 2, using MINITAB 16, which were the inscribed central composite and the Box-Behnken with central points, as types of the response surface design [103]. The empirical parameters for the correlations were determined to maximize R\textsuperscript{2} under statistically reliable conditions where P values for variables are below 0.05 and the P value for the residual error is above 0.05. The R\textsuperscript{2} values for the correlations fitted by DOE 1 (CDOE 1) and 2 (CDOE 2) were 86.5 and 97.2 %, respectively. Each correlation was applied to the
different DOE set and compared with the analytical model used in this study as shown in Table H.1 and Figure H.1.

Figure H.1: Parity plots of the two correlations for DOE 1, 2; a) CDOE 1; b) CDOE 2.
Table H.1: $R^2$ values of the two statistical correlations and the analytical model for DOE 1 and 2

<table>
<thead>
<tr>
<th>Evaluation</th>
<th>Statistical CDOE 1</th>
<th>Statistical CDOE 2</th>
<th>Analytical</th>
</tr>
</thead>
<tbody>
<tr>
<td>DOE 1</td>
<td>86.5%</td>
<td>37.6%</td>
<td>91.2%</td>
</tr>
<tr>
<td>DOE 2</td>
<td>55.6%</td>
<td>97.2%</td>
<td>85.1%</td>
</tr>
</tbody>
</table>

The lower $R^2$ of CDOE 1 than CDOE 2 is because CDOE 1 was fitted from the larger operation range which means larger $k_{overall}$ range as shown in a) of Figure H.1. However, the $R^2$ of CDOE 1 for DOE 2 can be higher than the $R^2$ of CDOE 2 for DOE 1 because the experimental range of DOE 2 is within DOE 1. In the case of CDOE 2, the $R^2$ is the highest at 97% due to a narrower experimental range and smaller number of experiments. But, the lowest $R^2$ of CDOE 2 for DOE 1 means that 97% is the result of overfitting by many empirical parameters and limited number of experiments as shown in b) of Figure H.1. Comparing with the large difference between $R^2$ values of the statistical correlations for DOE 1, 2, the $R^2$ values of the analytical modeling for DOE 1, 2 were not much different between each other although the intrinsic reaction kinetics were obtained in a part of the whole industrial operating condition. This demonstrates that the analytical model can be more robust and accurate in wider operation range than statistical correlations because the analytical model is based on the better theoretical understanding of the reactor behavior.
APPENDIX I. SIMULATION AND ANALYSIS OF THE OVERALL REACTION RATE FOR THE STIRRED TANK

For the intermediate reaction regime $0.6 < Ha < 1.4$, the reactor model of this study considered the reaction combined with the mass transfer in the aqueous phase film, the reaction in the bulk region of the aqueous phase and the film resistance of the organic phase. In order to confirm the necessity of the complex model and better understand the characteristics of the biphasic reaction in the intermediate reaction rate, the simulation with temperature at $1.9 \ M \ C_{NaOH}$ and 1000 rpm by the model was carried out and compared with the model for other regimes, as shown in Figure I.1. The temperature range was set from $25 \ ^\circ C$ for the slow reaction regime to $180 \ ^\circ C$ for the fast reaction regime. In this simulation, the rates of bulk reaction ($R_{bulk}$), mass transfer without reaction ($R_{aq,kL}$) and film reaction using penetration model ($R_{penetration}$) were calculated as the maximum rate.

$$R_{bulk} = \varepsilon_{aq}V_t k^* C^*; \quad R_{aq,kL} = aV_t k_{L,aq} C^*$$

$$R_{penetration} = aV_t C^* \sqrt{Dk}; \quad R_{org,kL} = aV_t k_{L,org} C_{org}$$

(I.1)
Figure I.1: Comparison of the film model for intermediate reaction with bulk reaction for slow reaction, mass transfer without reaction, film reaction by penetration model for fast reaction

For 25 °C, the film reaction can be ignored because the Hatta number was 0.17 and the bulk reaction rate is 10 times lower than the mass transfer rate without reaction. Thus, the overall reaction rate is dominated by the bulk reaction rate in the aqueous phase. The film model used in this work was followed to the bulk reaction rate in this regime. If the temperature is increased to around 60 °C, the bulk reaction rate is similar with the mass transfer rate without reaction so that the reaction in the bulk region is limited by the mass transfer rate. In addition, because $C_{\text{bulk}}$ is 60% of $C^*$, the film reaction cannot be ignored and the actual bulk reaction rate with 60% of $C^*$ becomes significantly decreased with
the actual bulk region decreased by the subtraction of the film volume. Thus, the overall reaction rate by the model is closed to the film reaction rate by penetration model as the maximum case. At around 110 °C, the film reaction rate is same with the mass transfer rate without reaction, with $Ha = 1$. However, as this is the intermediate reaction regime, the film reaction cannot dominate the overall reaction rate due to the mass transfer rate which affects the overall reaction rate, significantly for $Ha < 3$. The bulk reaction, though its effect keeps decreasing, has to be considered because $C_{\text{bulk}}$ is still 20-40 % of $C^*$. According to increasing the reaction rate in the aqueous phase, the film resistance in the organic phase has to be considered especially for low rpm cases. Thus, as described in Eq. 3.12, the model for intermediate reaction regime has to consider these points to estimate accurately the reaction rate as shown in Figure I.1. If the temperature is over 140 °C, the overall reaction rate is gradually dominated by the reaction rate in the film with 1.5 – 3 of the Hatta number. With decreasing rpm, the reaction rate in the film can dominate the overall reaction rate in lower temperature because the mass transfer rate strongly depends on rpm. Also, the film resistance of the organic phase ($R_{\text{org,kl}}$) starts to affect the overall reaction rate, although, in this case, its effect was much small due to around 10 times larger than the reaction rate in the aqueous phase.
APPENDIX J.  AN IMPROVED INTERFACIAL AREA ESTIMATION
METHOD: IAC WITH CFD

The requirement of an improved interfacial area estimation method was suggested to increase the estimation accuracy of the nBAL aldol condensation in the biphasic stirred tank reactor, described in section 3.4.3. The interfacial area concentration (IAC) with computational fluid dynamics (CFD) can be a good option for the purpose, considering the 3D effect on drop distribution in the reactor with the averaged drop diameter in a unit volume, as introduced in section 1.1.3. Among total four IAC models [215], the model by Hibiki and Ishii [216] was selected because it considered the Weber number in both breakup and coalescence efficiencies, without the wake entrainment term which is not needed for a biphasic stirred tank. Because the model provided reverse trend for temperature and salt effects, similar to the semi-empirical correlations in section 3.3.5, the model was modified by adding the viscosity correction term for the Weber number. The IAC model used in the present work is:

\[
\frac{\partial a_i}{\partial t} + \nabla \cdot (a_i \vec{v}_i) = S_{RC} + S_{TL}
\]

\[
S_{RC} = - \frac{C_{RC}}{6^{\frac{11}{3}}(\alpha_{max} - \alpha)} \frac{1}{\alpha^{\frac{5}{3}}} \exp \left( \frac{1}{\alpha} \frac{\rho_f}{\sigma} \frac{1}{\alpha^{0.5}} \frac{1}{\alpha} \frac{\mu_c}{\mu_d} \right) \exp \left( \frac{5}{6} \frac{\alpha}{\alpha} \frac{1}{\alpha} \frac{1}{\alpha} \frac{\mu_c}{\mu_d} \right)
\]

\[
S_{TL} = \frac{C_{TL}}{6^{\frac{11}{3}}(\alpha_{max} - \alpha)} \alpha^{\frac{2}{3}} \exp \left( \frac{5}{6} \frac{\alpha}{\alpha} \frac{1}{\alpha} \frac{1}{\alpha} \frac{\mu_c}{\mu_d} \right)
\]
The five adjustable parameters in the model were obtained from measurement data in section 3.3 using MATLAB without the convection flow term, $\nabla \cdot (a_i \vec{v}_i)$, comparing with the parameters from the literature obtained from adiabatic air-water bubbly flows for a nuclear reactor [216].

Table J.1: Comparison of adjustable parameters between literature and this work

<table>
<thead>
<tr>
<th></th>
<th>$C_{TI}$</th>
<th>$k_B$</th>
<th>$C_{RC}$</th>
<th>$k_C$</th>
<th>$C_{mu}$</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hibiki (2000)</td>
<td>0.264</td>
<td>1.37</td>
<td>0.188</td>
<td>1.29</td>
<td>-</td>
</tr>
<tr>
<td>This work</td>
<td>0.8382</td>
<td>0.78</td>
<td>1.27E-05</td>
<td>1.61</td>
<td>1.54</td>
</tr>
</tbody>
</table>

Table J.1 shows that the breakage and coalescence efficiencies in this work are smaller and larger than the results of Hibiki and Ishii [216], respectively. This may be because the salt effect in this work makes coalescence easy and breakage difficult. On the other hand, adjustable variables for breakage and coalescence are larger and smaller, respectively because drop diameters for stirred tanks are smaller than the diameters for vertical air-water bubbly flows without agitation. The modified IAC model in the present work explained the temperature and salt effects well, which is similar to section 3.3.5 for the semi-empirical correlation as shown in Figure J.1.
Figure J.1: Estimation results of the modified IAC model for the temperature and salt effects at 800 rpm agitation speed (symbols: measurements; lines: estimations)

To consider 3D effect with the convection flow term, it is necessary to combine the IAC model with continuity and momentum transport equations for biphasic system. The Reynolds averaged Navier-Stokes equations (RANS) with the standard $k$-$\varepsilon$ equations were used for turbulent flow with the Euler-Eulerian approach for multiphase flows and the multiple reference frame for agitation [201]. Schiller-Naumann and Troshko-Hassan models were used for the phase interactions for momentum transfer and $k$-$\varepsilon$ transport equations, respectively [201, 217]. The CFD models including the IAC model were numerically solved with 307,000 meshes for the reactor geometry using FLUENT 16.1. The IAC model provides interfacial area data averaged drop diameters, not drop size distribution. The IAC model with CFD, however, provides interfacial area distribution due to difference between unit volumes inside a tank. Although it is not inherent drop distribution in a unit volume, it provides the effect of the convective flow in a tank on drop diameters as shown in Figure J.2.
Figure J.2: Comparisons between drop distributions of measurements and estimations by IAC with CFD: a) 110 °C, 1.9 M C$_{\text{NaOH}}$ and 1000 rpm; b) 110 °C, 1.9 M C$_{\text{NaOH}}$ and 600 rpm

The figure shows that drop distributions at 1000 rpm can be explained by IAC with CFD model while the 600 rpm case is not satisfactory by the estimations. The drop distribution at 600 rpm became broad distribution by weak agitation and convection not sufficient to
be explained only by the 3D effect, requiring inherent drop distribution modeling using population balance equation (PBE). It is, however, clear that the 3D effect is a part of drop distribution in a reactor, which cannot be ignored to estimate drop distributions in stirred tank reactors.

The estimations by the IAC with CFD model were applied for the reactor model developed in this work, comparing with previous results in section 3.4.3 (Figure J.3).

![Figure J.3: Comparison of the estimation accuracy between the reactor model using $d_{32}$ obtained by measurements, the correlation and the IAC with CFD; the dash lines represent ±8% errors.](image)

The $R^2$ value of the model with the IAC/CFD was 98.1%, which was higher than 97.3% with $d_{32}$ by the semi-empirical correlation, and 98.5% from the model with $d_{32}$ measurements.
REFERENCES
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Shinbeom Lee received his bachelor’s degree in chemical engineering from Yonsei University in South Korea. Subsequently, he worked for mobile programming and marketing in the J.C.Hyun system for 39 months as substitution of military duty. Then, he received his master degree under Prof. En Sup Yoon’s direction in chemical and biological engineering from Seoul National University. During this period, he conducted a modeling study of a fixed bed reactor for dimethylether synthesis from syngas with collaboration of KOGAS, which was published in KIChE Journal and KIGAS Journal. After this, he worked in the LCD part of the Samsung Electronics for 4 months and at the R&D center of the Hanwha Chemical Company from Winter 2005 to Summer 2011.

With financial support of Hanwha Chemical Company, he joined the chemical engineering program at Purdue University in fall of 2011, where he joined the research group of Prof. Arvind Varma. There, Shinbeom began work on a biphasic stirred tank reactor for n-butyraldehyde aldol condensation. His work on the biphasic stirred tank reactor has led to two publications in Chemical Engineering Science and AIChE Journal, and another manuscript in preparation. He also carried out study on a trickle bed reactor for acetophenone hydrogenation with collaboration of the Eli-Lilly & Company, reporting two manuscripts submitted in Chemical Engineering Journal. In addition to his publications, Shinbeom has given oral presentations of his work at AIChE Annual
Meetings every year from 2013-2015, receiving an award for the best presentation in his session in 2014. He also presented his work in the I&EC graduate award invited session at the ACS meeting 2015, and received 2015 Doh Wonsuk Memorial Award from KICHE and three travel awards from Eastman, Purdue University and AIChE.

Outside his own research, Shinbeom has been involved in many aspects of the local community. These include serving as Chair of Korean Chemical Engineers at Purdue for the 2013-2014 academic year, Leader of 1st family group and Leading teacher of Youth group at the Purdue Korean Presbyterian Church from 2012 to 2015.

Shinbeom will graduate with a PhD degree in Chemical Engineering in December 2015, and will return to Hanwha Chemical Company in Korea.