Hydrodynamic effects of air sparging on hollow fiber membrane filtration in a bubble column reactor

Xia, Lijun

2014

Xia, L. (2014). Hydrodynamic effects of air sparging on hollow fiber membrane filtration in a bubble column reactor. Doctoral thesis, Nanyang Technological University, Singapore.

<https://hdl.handle.net/10356/59976>

<https://doi.org/10.32657/10356/59976>

**NANYANG
TECHNOLOGICAL
UNIVERSITY**

**Hydrodynamic Effects of Air Sparging on Hollow
Fiber Membrane Filtration in a Bubble Column
Reactor**

XIA LIJUN

School of Civil and Environmental Engineering

2013

**HYDRODYNAMIC EFFECTS OF AIR SPARGING ON
HOLLOW FIBER MEMBRANE FILTRATION IN A
BUBBLE COLUMN REACTOR**

XIA LIJUN

School of Civil and Environmental Engineering

A thesis submitted to
the Nanyang Technological University
in fulfilment of the requirements for the degree of
Doctor of Philosophy

2014

ACKNOWLEDGEMENTS

Working on the Ph.D. has been a marvellous and unforgettable experience. I would like to take this opportunity to express my profound gratitude to many people for sharing their precious experience in the pursuit of my Ph.D.

First and foremost, I am deeply grateful to my supervisor Assoc Prof. Adrian Wing-Keung Law. It is a great pleasure to be able to work with you. Despite such a busy working schedule, you have always been patient and encouraging in times of new ideas and difficulties. Your passion and professionalism are very much admired. Your continuous support and care also made me feel a friend, which I really appreciate from the bottom of my heart.

Also, I would like to express my special thanks to Prof. Tony Fane. You have always listened to me and your unique perspective on the research has made this project productive and stimulating. Your earnest working attitude and high scientific standard are also deeply impressed.

Furthermore, I would like to thank Dr. Adrian Yeo for the guidance and support at the preparing stage of this project. Thanks to your help, I was able to build up the experimental setup within a relatively short time.

I also very much cherished the time collaborating with many great people: Research Associates Ma Cong and Lin Long, and the Final Year Project students, Hsu Wen-Long and Lou Pang-Boon, Thomas and Michelle, Ong Huikiang and Ong Yishuang. They have contributed to the lab experiments and data collection, which were indispensable to this project.

The continuous supports from my friends: Mr. Zhao Bing, Ms Li Tian, Ms Song Jie, Mr. Jiang Baoxin and Ms Li Min are very heart-warming. They shares in my sorrows as well as in my joys.

I would like to express my great gratitude for the continuous support from the Singapore Membrane Technology Centre (SMTTC) which provided both the Research Scholarship together with the School of Civil and Environmental Engineering under NTU as well as facility and space support that were keys to the completion of this study.

Last but not least, I would like to thank my family members for their continuous care and understanding. They always believe in me and give me wings to soar. I really love you all!!!

This research was financially supported by the National Research Foundation (NRF) under its Environmental & Water Technologies Strategic Research Programme (EWI RFP 0901-IRIS-03-01) and administrated by the Environmental & Water Industry Programme Office (EWI) of PUB Singapore.

TABLE OF CONTENTS

ACKNOWLEDGEMENTS	I
TABLE OF CONTENTS	III
SUMMARY	VIII
LIST OF TABLES	XI
LIST OF FIGURES	XII
LIST OF SYMBOLS	XVII
LIST OF ABBREVIATIONS	XX
RESEARCH OUTPUT	XXII
1. INTRODUCTION	24
1.1 Background	24
1.2 Aims and Objectives	26
1.3 Thesis Layout	26
2. MULTIPHASE REACTORS	28
2.1 Introduction	28
2.2 Fluidized Bed Reactor (FBR)	28
2.3 Bubble Column Reactor (BCR)	31
2.3.1 Flow Regimes in BCRs	31

2.3.2	Factors Affecting System Performance	33
2.3.2.1	Superficial Gas Velocity	33
2.3.2.2	Liquid Phase Properties	35
2.3.2.3	Operating Conditions	35
2.3.2.4	Gas Sparger	36
2.3.2.5	Column Configuration	37
2.3.2.6	Solid Concentration	37
2.4	Membrane Reactors (MBRs)	38
2.4.1	History of MBR Development	38
2.4.2	Fouling in MBRs	40
2.4.2.1	Hollow Fiber Membranes	40
2.4.2.2	Membrane Fouling	42
2.4.3	Factors Affecting Fouling Formation	44
2.4.3.1	Feed and Biomass Characteristics	44
2.4.3.2	Membrane Characteristics	46
2.4.3.3	Operating Conditions	49
3.	MEMBRANE FILTRATION TESTS	53
3.1	Introduction	53
3.2	Air Sparging	53
3.2.1	Bubble Characteristics	54
3.2.2	Bubbling for Fouling Mitigation	56
3.2.2.1	Fouling Alleviation Mechanism	56
3.2.2.2	Flow Dynamics	57
3.2.3	Air Sparger Configurations	58
3.3	Experimental Methodologies	62
3.3.1	Experimental Setup	62
3.3.1.1	Apparatus	62
3.3.1.2	Membrane Specifications	64
3.3.1.3	Feed Solutions	65

3.3.2	Operating Conditions	69
3.4	Results and Discussions	73
3.4.1	Measurements and Calculations	73
3.4.2	Continuous Air Sparging	73
3.4.2.1	The Effect of Flow Regime	73
3.4.2.2	The Effect of Air Flow Rate	74
3.4.3	Alternate Air Sparging	75
3.4.3.1	The Effect of Flow Regime	75
3.4.3.2	The Effect of Air Flow Rate	77
3.4.3.3	The Effect of Switching Interval	78
3.4.3.4	The Effect of Membrane Module Orientation	79
3.4.4	The Effect of Feed Solution	80
3.5	Conclusions	84
4.	DYNAMIC PRESSURE SENSING TECHNIQUE	85
4.1	Introduction	85
4.2	Hydrodynamics Measurement Techniques	86
4.2.1	Point Measurement Probes	86
4.2.2	Imaging and Velocimetric Techniques	91
4.2.3	Constraints of Different Techniques	96
4.3	Experiments	97
4.3.1	Apparatus	97
4.3.2	Experimental Procedures	98
4.3.3	Submersible Pressure Sensor	99
4.3.4	Noise Reduction	99
4.4	Results and Discussions	102
4.4.1	Measurements and Calculations	102
4.4.2	Time-Series Data Analysis	102
4.4.2.1	Statistical Analysis	102

4.4.2.2	Spectral Analysis	111
4.4.2.3	Hurst Analysis	119
4.4.2.4	Chaos Analysis	123
4.5	Conclusions	126
5.	COMPUTATIONAL FLUID DYNAMICS (CFD)	127
5.1	Introduction	127
5.2	Literature Review	128
5.2.1	Multiphase Models	128
5.2.2	Simulation of MBRs	135
5.3	Conservation Laws of Fluid Motion	136
5.3.1	Governing Equations	136
5.3.2	Interfacial Forces	139
5.4	Turbulence Modelling	140
5.4.1	k - ε Turbulence Closures for RANS Models	140
5.4.1.1	Standard k - ε closure	140
5.4.1.2	Realizable k - ε closure	142
5.4.2	Large Eddy Simulation (LES)	143
5.5	Model of BCR	146
5.5.1	Model Geometry	146
5.5.2	Meshing	147
5.5.3	Physical Properties and Boundary Conditions	148
5.5.4	Results and Discussions	151
5.5.4.1	Grid Independence Test	151
5.5.4.2	Liquid Velocity	152
5.5.4.3	Gas Volume Fraction	157
5.6	Model MBR with Submerged Membrane Module	160
5.6.1	Membrane Module Specifications	160

5.6.1.1	Block Membrane Module	160
5.6.1.2	Hollow Fiber Membrane Module	162
5.6.2	Results and Discussions	163
5.6.2.1	Air Axial Velocity	163
5.6.2.2	Air Volume Fraction	170
5.6.2.3	Liquid Axial Velocity	175
5.6.2.4	Turbulent Kinetic Energy (TKE)	179
5.7	Data Validation using PIV	184
5.8	Comparisons among Different Approaches	188
5.9	Conclusions	193
6.	CONCLUSIONS AND RECOMMENDATIONS	194
6.1	Conclusions	194
6.2	Recommendations	195
	REFERENCES	197

SUMMARY

Submerged membrane filtration systems have many advantages over conventional water and wastewater treatment systems including better effluent quality, smaller footprint, higher organic loading and lower surplus sludge production. However, membrane fouling remains a challenge. Air sparging is now a standard approach to reduce concentration polarization and fouling of membrane modules in membrane bioreactors (MBRs). The hydrodynamic shear stresses, bubble-induced turbulence and cross flows induced by air sparging all help alleviate the deposition of foulants onto the membrane surface. However, despite its popularity, the detailed quantitative knowledge on the effect of air sparging on membrane fouling remains lacking in the literature due to the complex hydrodynamics generated by the gas-liquid flows around the membranes. To date, there is no valid model that quantifies the relationship between the membrane fouling performance and the flow hydrodynamics induced by air sparging as far as we are aware.

This study examines the relationship between the membrane filtration performance and the bubble-induced hydrodynamics around a submerged hollow fiber membrane module in a laboratory scale bubble column reactor (BCR). Membrane filtration tests were first carried out to investigate the relationship in a quantitative manner. The fouling performance was assessed under constant flux conditions with the monitoring of the trans-membrane pressure (TMP). Two axial locations were tested in order to differentiate the effect of flow regimes on the cake formation. Moreover, two aeration schemes of continuous and alternate air sparging were also studied to understand the associated flow dynamics and its influence on the membrane filtration performance. Additionally, factors such as the membrane module orientation, switching interval of alternate aeration as well as the nature of feed solutions were considered in the experiments.

Due to various constraints, most measuring techniques for hydrodynamic diagnostic are not applicable in multiphase reactors. In this study, a novel dynamic pressure

sensing technique was developed to characterize the hydrodynamics of the BCR with and without the submersion of the membrane module. The transient dynamic pressure fluctuations were recorded at different air flow rates, and special computational algorithms were applied to extract the hydrodynamic characteristics from the dynamic pressure signals. Finally, with the known hydrodynamic characteristics, the correlation with the measured membrane performance in the BCR was explored.

In addition to experimentation, computational fluid dynamics (CFD) modelling was also performed to study the time-averaged and transient flow behaviour inside the BCR. Three-dimensional numerical simulations were conducted using the Euler-Euler approach, with both the realizable $k-\varepsilon$ turbulence closure and Large Eddy Simulation (LES) to account for the turbulence effects. The BCR with and without the submerged membrane module were simulated respectively. Owing to the importance of the membrane configuration on the filtration efficiency, the membrane module geometry was simulated realistically as twenty solid tubes in this study to take account of the effect of individual fibers in order to improve the accuracy of the numerical predictions of the hydrodynamics around the membrane fibers. Overall, the computer-based method enabled the assessment of the fluid motions over the entire domain of the BCR, which was essential for system diagnostics and optimization.

In summary, the membrane filtration experiments confirmed the existence of an optimal air flow rate for the air sparging of a membrane module. Various factors such as the flow pattern, switching interval of alternate air sparging as well as membrane module orientation were all shown to have an influence on the membrane filtration performance. The dynamic pressure sensing measurement technique demonstrated its capability on the hydrodynamic characterization of the BCR. Nonlinear parameters of the Hurst exponent (H) and largest Lyapunov exponent (LLE) can be potentially used to delineate the fluid motions inside the BCR as well as to determine the range of the optimal air flow rate for a membrane filtration system. In addition, the implementation of CFD simulations with specific construction of membrane geometry provided detailed information on both the

transient and time-averaged flow behavior inside the BCR, especially the flow along and within the fiber bundle. Therefore, the integrated research of the three approaches enabled the full understanding of the hydrodynamics inside the multiphase reactor and its relationship with the membrane performance, which was rarely reported in earlier studies. The knowledge is useful for the optimization of the air sparger configuration and system operation.

LIST OF TABLES

Table 3.1: Summary of the liquid property (Tadaki and Maeda 1961)56

Table 5.1: The summary of RANS-based turbulence models134

Table 5.2 Physical properties of the air and water phases151

LIST OF FIGURES

Figure 2.1 Fluidization regimes with the increase of gas velocity in gas-liquid FBRs (Jakobsen 2008)	30
Figure 2.2 Schematic of possible flow regimes in BCRs (Kantarci <i>et al.</i> 2005).....	31
Figure 2.3 Flow pattern map of Hewitt and Robert (1969)	33
Figure 2.4 MBR systems (a) submerged configuration; (b) Side-stream configuration (Gupta <i>et al.</i> 2008).....	39
Figure 2.5 Characteristics of bubbles with: (a) axial and (b) transverse fiber array (Cui <i>et al.</i> 2003)	41
Figure 2.6 Mechanisms of cake depolarisation - bubbles outside fibers	41
Figure 2.7 Suction pressure profiles for tight and loose hollow fiber bundles (Chang <i>et al.</i> 2002)	42
Figure 2.8 Membrane fouling mechanisms (Le-Clech <i>et al.</i> 2006).....	43
Figure 2.9 Fouling Mechanisms for MBR operated at constant flux (Zhang <i>et al.</i> 2006)	51
Figure 3.1 Bubble shapes and wakes (Cui <i>et al.</i> 2003).....	55
Figure 3.2 Memjet module with integral jet aerator (Judd 2011).....	60
Figure 3.3 Microza EFM HF membranes aerated module schematic (Judd 2011)	61
Figure 3.4 Experimental setup for the membrane filtration tests.....	63
Figure 3.5 Picture of the bubbling flow with the air flow rate of 2.0 L/min at 0.6 m location.....	67
Figure 3.6 Hollow fiber membrane module used in the membrane filtration tests	68
Figure 3.7 Variations of dynamic viscosity for the clean tap water and three different concentrations of bentonite solutions of 0.65, 1.3 and 2.65 g/L	68
Figure 3.8 Sketch of the flow field with central aeration in the BCR	70
Figure 3.9 Transient flow condition induced by alternate aeration in the BCR	71
Figure 3.10 Membrane module orientations: (a) 0 degree and (b) 90 degree.....	72
Figure 3.11 Fouling rates for 0.65 g/L bentonite solution at the 0.3 and 0.6 m locations (Flux = 71.6 L/m ² hr).....	75

Figure 3.12 Variations of TMPs with alternate aeration at 0.5, 1 and 2 minutes switching intervals at the 0.3 and 0.6 m locations for a 0.65 g/L bentonite solution.....	76
Figure 3.13 Variations of TMPs at the two membrane orientations for a 0.65 g/L bentonite solution.....	79
Figure 3.14 Fouling rates for 0.65 g/L bentonite and yeast solutions at the 0.3 and 0.6 m locations	81
Figure 3.15 Variations of TMPs at (a) 0.3 m and (b) 0.6 m locations for a 0.65 g/L yeast solution (Flux = 71.6 L/m ² hr).....	83
Figure 4.1 Experimental setup for the dynamic pressure sensing measurements.....	100
Figure 4.2 Dynamic pressure fluctuations at the 0.6 m (left) and 0.3 m (right) locations at the air flow rates of 0.1, 0.5 1.0 and 2.0 L/min at a sampling rate of 0.01s	101
Figure 4.3 Average pressure at the (a) 0.3 and (b) 0.6 m locations for different air flow rates.....	104
Figure 4.4 Standard deviations of the dynamic pressure at 0.3 and 0.6 m locations under different air flow rates.....	105
Figure 4.5 Average dynamic pressure at the (a) 0.3 and (b) 0.6 m locations under different air flow rates for clean tap water, 0.65 g/L bentonite and 0.65 g/L yeast solutions.....	108
Figure 4.6 Standard deviation of the dynamic pressure at the (a) 0.3 and (b) 0.6 m locations under different air flow rates for clean tap water, 0.65g/L bentonite and 0.65 g/L yeast solutions.....	110
Figure 4.7 Power spectrum at the 0.3 m location without (left) and with (right) the submersion of the membrane module for the four air flow rates.....	116
Figure 4.8 Power spectrum at the 0.5 m location without (left) and with (right) the submersion of the membrane module for the four air flow rates.....	117
Figure 4.9 Power spectrum at the 0.7 m location without (left) and with (right) the submersion of the membrane module for the four air flow rates.....	118
Figure 4.10 Development of the Hurst exponent with the air flow rate at the 0.3 and 0.6 m locations in clean tap water.....	121

Figure 4.11 Development of the Hurst exponent with the air flow rate at the 0.3 and 0.6 m locations in clean tap water with a submerged hollow fiber membrane module.....	122
Figure 4.12 Development of the <i>LLE</i> with the air flow rate at the 0.3 and 0.6 m locations in clean tap water.....	124
Figure 4.13 Development of the <i>LLE</i> with the air flow rate at the 0.3 and 0.6 m locations in clean tap water with the submerged hollow fiber membrane module.....	125
Figure 5.1 Schematic of the BCR (all dimensions in mm) and measurement axial locations (reproduced from Gan et al., 2011).	147
Figure 5.2 Numerical meshes: (a) the BCR and the hemispheric inlet boundary for (b) top view and (c) side view.....	149
Figure 5.3 Estimated average bubble size for different air flow rates	150
Figure 5.4 Grid independence test: liquid axial velocity at the centreline.....	152
Figure 5.5 Contours of the liquid axial velocity magnitude (m/s) at a 0.13 L/min air flow rate: (a) <i>k-ε</i> transient, (b) <i>k-ε</i> time-average, (c) LES transient, and (d) LES time-average.....	154
Figure 5.6 Comparison between the CFD simulation results and the experimental data (Gan et al., 2011) of the liquid axial velocity at six axial elevations ($z=100, 130, 160, 200, 400$ and 600 mm) for three air flow rates of (a) 0.13, (b) 0.25 and (c) 0.38 L/min (r is the radial co-ordinate, and R is the radius of the BCR)	156
Figure 5.7 Contours of the air volume fraction at 0.13 L/min air flow rate: (a) <i>k-ε</i> transient, (b) <i>k-ε</i> time-average, (c) LES transient, and (d) LES time-average.....	157
Figure 5.8 Comparison between the CFD simulation results and the experimental data (Gan et al., 2011) for the air volume fraction at six axial elevations ($z=100, 130, 160, 200, 400$ and 600 mm) for the three air flow rates of (a) 0.13, (b) 0.25 and (c) 0.38 L/min.....	159
Figure 5.9 Geometry of the block membrane: (a) top view and (b) side view	161
Figure 5.10 Numerical meshes for the BCR with the membrane module located at (a) 0.3 and (b) 0.6 m from the bottom of the reactor.....	161

Figure 5.11 Geometry of the fiber bundle constructed of solid tubes: (a) top view and (b) side view	162
Figure 5.12 Detailed dimensions for the fiber model	162
Figure 5.13 Top view of the BCR showing specific data coordinates ($\pm 0.03, 0$)	163
Figure 5.14 Contours of the air axial velocity magnitude (m/s): (a) 0.3 m block, (b) 0.3 m fiber, (c) 0.6 m block and (d) 0.6 m fiber	164
Figure 5.15 Air axial velocity against the air flow rate for the block and fiber models	165
Figure 5.16 Air axial velocity for the block and fiber models at the 0.3 m membrane position at a 1 L/min air flow rate along the x-axis at (a) 0.9 m, (b) 0.6 m and (c) 0.3 m column heights	167
Figure 5.17 Air axial velocity for both models at the 0.6 m membrane position at 1 L/min air flow rate along the x-axis at (a) 0.9 m, (b) 0.6 m and (c) 0.3 m column heights	169
Figure 5.18 Contours of the air volume fraction for the two models: (a) 0.3 m block, (b) 0.3 m fiber, (c) 0.6 m block and (d) 0.6 m fiber.....	170
Figure 5.19 Air volume fraction against the air flow rate.....	171
Figure 5.20 Air volume fraction for the two models at the 0.3 m membrane position at a 1 L/min air flow rate along the x-axis at (a) 0.9 m, (b) 0.6 m and (c) 0.3 m column heights	173
Figure 5.21 Air volume fraction for the two models at the 0.6 m membrane position at a 1 L/min air flow rate along the x-axis at (a) 0.9 m (b) 0.6 m and (c) 0.3 m column heights	174
Figure 5.22 Liquid axial velocity versus air flow rate	175
Figure 5.23 Liquid axial velocity predicted by the models at the 0.3 m membrane position at 1 L/min air flow rate along the x-axis at (a) 0.9 m (b) 0.6 m and (c) 0.3 m column heights	177
Figure 5.24 Liquid axial velocity predicted by the models at 0.6 m membrane position at a 1 L/min air flow rate along the x-axis at (a) 0.9 m (b) 0.6 m and (c) 0.3 m column heights	178
Figure 5.25 TKE against the air flow rate for the block and fiber models	179

Figure 5.26 The TKE for the two models at the 0.3 m membrane position at a 1 L/min air flow rate along the x-axis at (a) 0.9 m, (b) 0.6 m and (b) 0.3 m column heights	182
Figure 5.27 The TKE for the two models at the 0.6 m membrane position at a 1 L/min air flow rate along the x-axis at (a) 0.9 m, (b) 0.6 m and (c) 0.3 m column heights	183
Figure 5.28 Air axial velocity profile for block, fiber and PIV data with the membrane submerged at the 0.3 m location at (a) 0.2 and (b) 0.5 L/min air flow rates.....	186
Figure 5.29 Air axial velocity profile for block, fiber and PIV data with the membrane submerged at the 0.6 m location at (a) 0.2 and (b) 0.5 L/min air flow rates.....	187
Figure 5.30 Development of the Hurst exponent and the fouling rate as a function of the air flow rate at the 0.3 m location	189
Figure 5.31 Development of the <i>LLE</i> and the fouling rate as a function of the air flow rate at the 0.3 m location	190
Figure 5.32 Development of the air axial velocity for the fiber model and the fouling rate as a function of the air flow rate at the 0.3 m location	191
Figure 5.33 Development of the TKE for the fiber model and the fouling rate against the air flow rate at the 0.3 m location	192

LIST OF SYMBOLS

A	Cross-sectional area of the BCR
C_D	Drag coefficient
C_s	Smagorinsky constant
d	Distance to the closest wall (m)
d / a	Bubble deformation factor
D	Fluid domain
f_s	Sample frequency (Hz)
$\overline{F}$	Body force (N)
F_D	Drag force (N)
F_x	Discrete Fourier Transform
F_x^*	Complex conjugate of F_x
g	Acceleration due to gravity (m/s ²)
g_i	Gravitational vector in the i th direction (m/s ²)
G	Filter function
G_b	Generation of turbulent kinetic energy due to buoyancy
G_k	Generation of turbulent kinetic energy due to mean velocity gradient
H	Hurst exponent
k	Turbulent kinetic energy (m ² /s ²)
L	Distance between two points
L_s	Mixing length (m)
$\dot{m}_{pq}$	Mass transfer from phase p to q
$\dot{m}_{qp}$	Mass transfer from phase q to p
M	Morton number
M_t	Turbulent Mach number
n	Number of phases
N	The number of pressure signals
N_f	Resistance number
N'_s	Shear number

P_f	Feed pressure (Pa)
P_p	Permeate pressure (Pa)
P_r	Retentate pressure (Pa)
Pr_t	Turbulent Prandtl number
P_s	Suction pressure (Pa)
P_{TMP}	TMP (Pa)
P_x	Time-series of pressure signals at position x
Q	Air flow rate (m ³ /s)
Re	Reynolds number
R / S	Rescaled range of the time series
S	Module of the mean rate-of-strain tensor
$\overline{S_{ij}}$	Rate-of-strain sensor
S_K, S_ε	User-defined source terms
S_k	Standard deviation of the time-series
t	Time (s)
T	Sub-period length
$\vec{T}$	Strain stress (N/m ²)
U_b	Bubble velocity (m/s)
V_a	Average superficial air velocity (m/s)
x_i	Magnitude of the pressure signals (kPa)
$x(t)$	Discrete time series
Y_M	Fluctuation dilatation
α_k	Volume fraction for phase k
β	Thermal expansion coefficient
ε	Turbulent dissipation rate (m ² /s ²)
η	Scalar measure of the deformation sensor
κ	Von Karman constant
λ	Lyapunov exponent (1/s)
μ	Mean pressure (kPa)
μ_m	Mixture viscosity (kg/m·s)
μ_L	Liquid viscosity (kg/m·s)

μ_q	Dynamic viscosity of phase q (kg/m·s)
μ_t	Turbulent (eddy) viscosity of phase q (kg/m·s)
ν	Turbulent kinematic viscosity (m ² /s)
$\vec{v}_{d,k}$	Drift velocity for phase k (m/s)
$\vec{v}_{d,p}$	Drift velocity for phase p (m/s)
$\vec{v}_m$	Mass-averaged mixture velocity (m/s)
$\vec{v}_{k,q}$	Slip velocity of phase k (m/s)
$\vec{v}_{p,q}$	Slip velocity of phase p (m/s)
ρ_L	Liquid density (kg/m ³)
ρ_m	Mixture density (kg/m ³)
ρ_p	Density of phase p (kg/m ³)
σ	Surface tension (kg/s ²)
σ_k	Turbulent Prandtl number for turbulent kinetic energy
σ_ε	Turbulent Prandtl number for dissipation rate
σ_{ij}	Stress tensor
σ_{STD}	Standard deviation of pressure signals (kPa)
τ_{ij}	Sub-grid scale stress
φ_{xx}	PSDF
$\overline{\Omega_{ij}}$	Mean rate-of-rotation tensor
ω_k	Angular velocity (1/s)
Δ	Local grid scale

LIST OF ABBREVIATIONS

BCR	Bubble Column Reactor
CARPT	Computer Automated Radioactive Particle Tracking
CFD	Computational Fluid Dynamics
CFV	Cross Flow Velocity
CT	Computerized Tomography
CTA	Constant Temperature Anemometry
DAQ	Data Acquisition system
DFT	Discrete Fourier Transform
DIA	Digital Image Analysis
DNS	Direct Numerical Simulation
DO	Dissolved Oxygen
DSP	Digital Signal Processing
ECT	Electrical Capacitance Tomography
EPS	Extracellular Polymer Substance
ERT	Electrical Resistance Tomography
FBR	Fluidized Bed Reactor
F/M	Food to Microorganism ratio
HRT	Hydraulic Retention Time
KE	Kolmogorov-Sinai Entropy
LDA	Laser Doppler Anemometry
LES	Large Eddy Simulation

LLE	Largest Lyapunov Exponent
MBR	Membrane Bioreactor
MF	Micro-Filtration
MLSS	Mixed Liquor Suspended Solid
PFS	Pressure Fluctuation Signals
PISO	Pressure Implicit with Split Operator
PIV	Particle Image Velocimetry
PLC	Programmable Logic Controller
PSDF	Power Spectral Density Function
PVC	Polyvinyl Chloride
PVDF	Polyvinylidene Fluoride
QM	Quality of Mixedness
RANS	Reynolds Averaged Navier-Stokes
SIMPLE	Semi-Implicit Method for Pressure-Linked Equations
SRT	Solid Retention Time
TMP	Trans-Membrane Pressure
TVD	Total Variation Diminishing
UF	Ultra-Filtration
WPT	Wavelet Packet Transform

RESEARCH OUTPUT

- **Publications**

Xia, L., A. W. K. Law and A. G. Fane (2013), “Hydrodynamic effects of air sparging on hollow fiber membranes in a bubble column reactor” Water Research, **47** (11): 3762-3772.

Lin, L., **Xia L.**, A. W. K. Law, “Numerical simulations of a bubble column with a bottom-mounted air source”, Chemical Engineering Science, Submitted

- **Patent Application**

Xia, L., A. W. K. Law and A. G. Fane, “Pressure Sensing System for Membrane Filtration Reactors”, in progress

- **Conferences**

Xia, L., A. W. K. Law and A. G. Fane, “Air sparging on hollow fiber membranes in a bubble column reactor”, Membrane Science and Technology Workshop, Singapore, 24 August 2011, Oral Presentation

Xia, L., A. W. K. Law and A. G. Fane, “Air sparging on hollow fiber membranes in a bubble column reactor”, Membrane Technology Conference & Exposition, Glendale, Arizona, U.S.A. 27 Feb- 01Mar, 2012, Oral Presentation

Xia, L. T. Li, A. W. K. Law and A. G. Fane, “Dynamic shear enhancement for fouling control in hollow fiber membrane filtration”, Singapore International Water Week, 1-4 July, 2012, Poster Presentation

Lin, L. **Xia, L.**, A. W. K. Law and A. G. Fane, “CFD simulations of a bubble column reactor with a submerged hollow fiber membrane module: relevance to MBRs”, MBR Asia, 21-22 January, 2013 , Oral Presentation

1. Introduction

1.1 Background

Due to the rapid growth of the world's population, the demand for water is rising tremendously. New technologies are needed to increase the drinkable water supply, while reducing the related energy consumption for the production at the same time. Among them, the membrane bioreactor (MBR) technology is the most important new technology due to its many advantages such as smaller footprint, higher volumetric loading, less sludge produced as well as better effluent quality compared to conventional activated sludge treatment processes (Chang 2002; Kang et al. 2008; Santos et al. 2011). However, membrane performance can be significantly reduced by fouling, which causes flux decline or higher trans-membrane pressure (TMP) and modifies the retention properties of the membranes. Hence, numerous methods both physical and chemical, have been developed and employed in the industry to alleviate MBR fouling (Le-Clech et al. 2006; Le-Clech 2010).

For submerged MBRs, aeration is now a standard approach for fouling mitigation. The function of aeration can be divided into three major aspects: (a) providing oxygen to the biomass; (b) reducing fouling by inducing turbulence; and (c) keeping the activated sludge suspended. It was reported that the control of fouling depends significantly on the bubble-induced hydrodynamics around the membrane surfaces (Böhm *et al.* 2012). Thus it is critical to quantify the hydrodynamic characteristics in MBRs due to their importance in influencing the processes inside the reactor. Furthermore, the energy cost of air blowers constitutes 30- 40% of the total energy consumption of MBRs (Judd 2011). Under this circumstance, improved design of the air diffusers is desirable for energy reduction and system optimization.

Until recently, the focus on membrane fouling alleviation by air sparging is on the mean flow circulation in the reactor (e.g. Liu et al. 2000; Kang et al. 2008; Khalili-Garakani et al. 2011). However, there is now an increasing recognition that turbulence fluctuations induced by bubbling also play an important role in membrane fouling control (e.g. Xu et al. 2002; Yeo et al. 2006; Chan et al. 2011). Overall, the effect of air sparging on membrane filtration is still not well understood due to the complexity caused by the phase-phase interactions. In terms of the hydrodynamics, the highly irregular and unsteady flow behaviour inside the reactor also makes it challenging to quantify the velocity distributions. When expanded to pilot or plant scale, the aggressive environment inside the reactor makes the application of the majority of the measurement techniques even more difficult. Till now, there is no valid quantitative model that can be used to correlate the membrane fouling performance with the flow hydrodynamics as far as we are aware. The system optimization only relies empirically on the manufacturer's recommendations and operator's experience (Böhm *et al.* 2012).

In the present study, in order to characterize the hydrodynamics of the MBR, a novel method of dynamic pressure sensing is developed by analysing the temporal signatures of parameters that are functions of the prevailing hydrodynamics. The technique is low cost and robust, and thus offers an attractive alternative to study the hydrodynamics of MBRs and other kinds of multiphase reactors, especially those with harsh reactor environments. Besides, state-of-the-art computational fluid dynamics (CFD) simulations of MBRs are also attempted in this study. The realistic simulation of the hollow fiber membrane module as a bundle of solid tubes enables the assessment of the flow through and around the module, which is critical for fouling control analysis. The membrane filtration experiments are also conducted as a fundamental approach to quantify the membrane performance under various operating conditions. The experimental results can be used as references to verify the accuracy of the measurements and numerical solutions derived from the dynamic pressure sensing and CFD.

1.2 Aims and Objectives

The focus of this research is to quantify the hydrodynamic effect of air sparging on a hollow fiber membrane module in a laboratory scale BCR.

The objectives of the present study are to:

- Monitor the TMP changes with time under various air flow rates in a laboratory scale BCR to investigate the effects of air sparging on the membrane filtration performance.
- Collect and analyse the dynamic pressure fluctuation signals (PFSs) to correlate the hydrodynamic characteristics with the membrane filtration performance.
- Establish a numerical model using the Reynolds-Averaged models with realizable $k-\varepsilon$ turbulent closures and Large Eddy Simulation (LES) to predict the hydrodynamic conditions of the BCR with and without the submersion of a membrane module under various transient and time-averaged operating schemes, and finally
- Integrate the experimental analysis and numerical results to attempt a method for quick determination of the optimized aeration condition for a specific membrane filtration system.

1.3 Thesis Organization

The report is arranged as follows:

Chapter One gives a brief introduction and background to the problems studied.

Chapter Two summarizes three types of multiphase reactors that are commonly used in industry, namely, the fluidized bed reactor (FBR), BCR and MBR. Basic

working principles as well as some important features that affect the system performance are discussed.

Chapter Three introduces the membrane filtration experiments that are conducted to investigate the impact of air induced-hydrodynamics on the membrane fouling mitigation. Background studies on air sparging are reviewed, followed by the description of the experimental apparatus and methodologies. The results are discussed to illustrate the influence of several factors on the membrane filtration performance. At the end of this chapter, conclusions are given to summarize the major experimental findings.

Chapter Four describes the novel dynamic pressure sensing approach that is used to quantify the hydrodynamics of the BCR with and without the existence of a membrane module. A number of measuring techniques are first introduced with a review of advantages and disadvantages. Subsequently, the experimental procedures are elaborated followed by the discussions of the results. Finally, nonlinear parameters that can be used for hydrodynamic diagnostic of the BCR are proposed.

Chapter Five introduces the concept of the CFD and elaborates the theoretical background of the method such as the governing equations and turbulent closures. The numerical modelling of the BCR with and without the membrane module is discussed respectively. The results are then compared with Particle Image Velocimetry (PIV) data for validation. In the end, conclusions are drawn to summarize the work that has been done.

Chapter Six concludes the major findings in this study and proposes some recommendations for future work.

2. Multiphase Reactors

2.1 Introduction

Multiphase reactors are widely used in the chemical, petrochemical, biochemical, metallurgical and environmental industries (Roy *et al.* 2004). For single-phase reactors, the only process involved is the mixing of the reacting components which influences the efficiency of the chemical conversion. However, the reaction processes in multiphase reactors are much more complex since not only the reacting components must be efficiently mixed, but also conditions have to be controlled to transport those components efficiently across the interfaces between the phases. Due to the complexity of the flow pattern and phase-phase interactions inside the reactor, the detailed hydrodynamics of the multiphase reactors are not completely understood yet (Lin *et al.* 2001; Al-Masry *et al.* 2005). In order to investigate the flow conditions in solid-gas-liquid three-phase membrane bioreactors (MBRs), the review on the basic characteristics of the solid-gas fluidized bed reactors (FBRs) and gas-liquid bubble column reactors (BCRs) can provide useful background information. Hence, in this chapter, three kinds of multiphase reactors, namely, the BCR, FBR and MBR, are reviewed with some important features discussed in the following.

2.2 Fluidized Bed Reactor (FBR)

A FBR is a type of reactor that can be used to carry out a variety of multiphase chemical reactions. When the fluid velocity is low, the particles are packed together, whereas they are fluidized when the fluid velocity exceeds a certain limiting value

(Werther 2000; Michaelides 2013). The circulation inside the reactor enhances the flow instability such as vortices and turbulence, which results in high level of mixing for both the particles and the carried fluid. The mixing processes involve momentum exchange, mass exchange as well as energy exchange (Michaelides 2013).

Some of the inherent advantages of the (gas-solid) FBR as a reaction system include (Werther 2000):

1. Easy handling and transport of solids due to liquid like behaviour of the FBR
2. Uniform temperature distribution due to intensive solids mixing
3. Larger solid-gas exchange area by the virtue of small solids grain size
4. High heat-transfer coefficients between the bed and immersed heating or cooling surfaces
5. Uniform (solid) product in the batchwise process because of intensive solid mixing

Against these advantages are the following disadvantages (Werther 2000):

1. Expensive solid separation or gas purification equipment required because of solids entrainment by fluidized gas
2. The high solids mixing rate causes a non-uniform residence time of solids, back-mixing of gas, and resulting lower conversion
3. In catalytic reaction, undesired bypass or broadening of the residence-time distribution for the reactor gas due to bubble development
4. Erosion of internals and attrition of solids (especially significant with catalysts), resulting from high solid velocities
5. Possibility of de-fluidization due to agglomeration of solids
6. Gas-solid counter-current motion possible only in multistage equipment
7. Difficulties in scaling-up

Fuel conversion in a FBR was first introduced by Winkler who patented a gasifier in 1922. The first large scale use of FBRs, the Winkler gas generator, was

established for the process of gasification of coal in 1926 (Froment and Bischoff 1990; Kunii and Levenspiel 1991; Jakobsen 2008). Since then, FBRs have been applied in many industrial processes. When gas is passed through a bed of solid particles, various flow regimes from fixed bed to pneumatic conveyance are observed with an increase in the gas velocity (see Figure 2.1) (Bi and Grace 1995; Jakobsen 2008). The prevailing flow regime and the quality of fluidization depend on several factors including the solid flow rate, gas flow rate and system configuration (Vivek 2002). Although transitions between the hydrodynamic regimes are expected, the quantitative values delineating the higher velocity regimes of fluidization and solid transport are poorly characterized in the literature. Since different flow regimes have different hydrodynamic characteristics which could result in different mixing and mass transfer characteristics, thus it is essential to determine the desired or prevailing flow regime in the FBRs at the design stage (Vivek 2002).

Figure 2.1 Fluidization regimes with an increase in the gas velocity in gas-liquid FBRs (Jakobsen 2008)

2.3 Bubble Column Reactor (BCR)

BCRs are multiphase reactors in which a discontinuous gas phase in the form of bubbles is brought into contact and reacts with a continuous liquid phase (Vial *et al.* 2000). Just as FBRs, BCRs are also used in a wide range of industries (Degaleesan *et al.* 2001). Owing to a number of advantages including high mass and heat transfer characteristics, low maintenance costs and long duration, BCRs have drawn considerable attentions in the past two decades (Kantarci *et al.* 2005).

2.3.1 Flow Regimes in BCRs

Generally, three flow regimes are commonly observed in BCRs: the bubbly flow regime, the churn-turbulent regime and the slug flow regime (see Figure 2.2) (Kantarci *et al.* 2005).

Figure 2.2 Schematic of possible flow regimes in BCRs (Kantarci *et al.* 2005)

Bubble flow occurs when the bubble sizes are significantly less than the tube or channel size. In bubbly flows, bubbles are easily observable as the gas is dispersed in the form of discrete bubbles in the continuous liquid phase. They vary widely in shape but are typically nearly spherical. Since the applied gas flow rate is usually low, the bubble flow is one of the typical flow regimes found in the system (Cui *et al.* 2003).

Churn flow occurs when the air flow rate increases, with the flow structure becoming unstable and the fluid travelling up and down in an oscillatory fashion but with a net upward flow. The instability is the result of the relative parity of the gravity and shear forces acting in the opposing directions. Churn flow is typically a flow regime to be avoided in a two-phase transfer since the mass of the slugs may have a destructive consequence on the piping system.

The slug flow mainly occurs with increasing gas void fractions. The proximity of the bubbles is very close such that they collide and coalesce to form larger bubbles, which are similar in dimension to the tube diameter. These bubbles are commonly referred to as Taylor bubbles and they have a characteristic shape similar to a bullet with a hemispherical nose and a blunt tail end. They are separated from one another by a slug of fluid which may include small bubbles, and are surrounded by a thin liquid film between them and the tube wall which may flow downward due to the force of gravity, even though the net flow of fluid is upward.

If the gas stream is introduced continuously with a diffuser type of sparger, the flow pattern will depend on the gas and liquid flow rates and the properties of the liquid. In this case, the flow pattern can be identified using the Hewitt and Robert map as a rough guide (see Figure 2.3) (Cui *et al.* 2003).

Figure 2.3 Flow pattern map of Hewitt and Robert (1969)

2.3.2 Factors Affecting System Performance

In this section, six primary factors affecting the BCR system performance are discussed. For each factor, four features are studied, namely, the gas holdup, the bubble size, and the heat and mass transfer coefficients. For mass transfer, the volumetric mass transfer coefficient $k_L a$ is normally used to characterize the transport efficiency.

2.3.2.1 Superficial Gas Velocity

The superficial gas velocity, which is a very important measure of the system performance, is the average velocity of the gas that is sparged into the column. It is

equal to the volumetric flow rate divided by the cross-sectional flow area of the column:

$$V_a = \frac{Q}{A} \quad (2.1)$$

where V_a is the average superficial air velocity, Q is the air flow rate and A is the cross-sectional area of the BCR.

- The superficial gas velocity controls the changes in the gas holdup, which is a key parameter for the design and characterization of transport phenomena of BCRs (Shah et al. 1982; Luo et al. 1999). It was found that the gas holdup increased with increasing superficial gas velocity and the increase was proportional to the magnitude of the superficial gas velocity with low void fractions (Deckwer et al. 1980; Schumpe and Grund 1986; Saxena et al. 1990; Daly et al. 1992; Pino et al. 1992; Hyndman et al. 1997; Krishna et al. 1997; Li and Prakash 2000; Prakash et al. 2001).
- In terms of bubble size, it was reported that it increased with increasing superficial gas velocity and at a certain gas velocity, the maximum bubble size was attained (Fukuma et al. 1987; Saxena et al. 1990). However, Akita and Yoshida (1974) claimed that the bubble size decreased with the gas flow rate.
- The superficial gas velocity also has a great influence on the heat and mass transfer coefficients in both two- and three-phase BCRs. It was found by many researchers that the heat transfer coefficient increased with the increase in superficial gas velocity irrespective of the solid or liquid phase properties (Deckwer et al. 1980; Saxena et al. 1990; Li and Prakash 1997; Li and Prakash 2000; Prakash et al. 2001).

- As for the mass transfer, Krishna and van Batern (2003) pointed out that the volumetric mass transfer coefficient $k_L a$ increased with the superficial gas velocity with the same trend as the gas holdup increased with the superficial gas velocity.

2.3.2.2 *Liquid Phase Properties*

- The liquid phase properties such as surface tension and viscosity impact the bubble formation and coalescing, and hence are important factors affecting the gas holdup. Generally, the gas holdup increased with increasing liquid viscosity (Li and Prakash 1997). Besides, the addition of surface active reagents and impurities also increased the gas holdup (Hikita et al. 1980; Sada et al. 1984; öztürk et al. 1987).
- The average bubble size was found to decrease with decreasing surface tension of the liquid and increasing foam agent concentrations while increased with increasing liquid viscosity (Akita and Yoshida 1974; Li and Prakash 1997; Veera et al. 2004).
- Concerning the volumetric mass transfer coefficient, the $k_L a$ value increased with the gas density and the presence of surfactants (öztürk et al. 1987; Muller and Davidson 1995).
- The heat transfer coefficient was found to decrease as a function of liquid viscosity (Kantarci *et al.* 2005).

2.3.2.3 *Operating Conditions*

The effects of operating pressure and temperature were also widely investigated.

- It was found that the gas holdup increased with an increase in the operating pressure. Concerning the temperature, it was commonly accepted that its influence was not so significant although opposing opinions still exist

especially for small dimension reactors (Deckwer et al. 1980; Reilley et al. 1986).

- It was observed that an increase in temperature or pressure resulted in reduced bubble sizes (Luo *et al.* 1999).
- Although it was reported that the pressure had no significant effect on the mass transfer coefficient, both the k_La values and the interfacial area were found to increase with increasing pressure (Dewes et al. 1995; Behkish et al. 2002).
- Regarding the heat transfer coefficient, it was reported that with increasing temperature, the heat transfer coefficient also increased (Saxena *et al.* 1990). In addition, increasing pressure also resulted in an increased heat transfer coefficient (Chen *et al.* 2003).

2.3.2.4 Gas Sparger

- It was found that the gas holdup was strongly affected by the type of sparger used in BCRs (Luo *et al.* 1999). The number as well as the diameter of the orifice holes had an effect on the gas holdup especially when the gas velocity was low (Schumpe and Grund 1986; Luo *et al.* 1999; Bouaifi *et al.* 2001). Moreover, it was observed that small orifice gas spargers actually led to higher gas holdup values (Bouaifi *et al.* 2001).
- The bubble sizes observed in BCRs were determined by the sparger configuration such that smaller orifice diameter plates enabled the formation of smaller size bubbles (Kantarci *et al.* 2005).
- In terms of the volumetric mass transfer coefficient, high values of k_La could be obtained with the spargers for which the gas holdup values were also higher (Kantarci *et al.* 2005).

2.3.2.5 *Column Configuration*

- The gas holdup was not much affected by the column dimension when the height to diameter ratio (aspect ratio) was larger than 5 (Shah *et al.* 1982; Luo *et al.* 1999).
- With the increase of the column diameter, the average bubble size and the rise velocity of large bubbles were found to be higher (Koide *et al.* 1979; Li and Prakash 2000).
- For the mass transfer, it was found that the $k_L a$ values decreased with the column diameter and were independent of the initial bed height (Krishna and van Baten 2003; Verma and Rai 2003).
- Saxena *et al.* (1990) reported that the heat transfer coefficient was higher with a larger column size due to a higher mixing rate.

2.3.2.6 *Solid Concentration*

- Generally, adding solids to a two-phase system decreased the holdup (Saxena *et al.* 1990; Pino *et al.* 1992; Krishna *et al.* 1997; Li and Prakash 1997; Li and Prakash 2000).
- Addition of solids and increasing the solid concentration increased the bubble coalescence, leading to the formation of larger size bubbles with higher rise velocities.
- As a matter of fact, the addition of solids caused larger bubble sizes with induced velocities and thus higher heat transfer rates (Deckwer *et al.* 1980; Li and Prakash 1997; Luo *et al.* 1999; Li and Prakash 2000). However, a decrease in the heat transfer coefficient with increasing solid concentration was reported as well (Li and Prakash 1997).

- In addition, the k_1a values were found to decrease with the addition of solids and increasing solid concentration (Schumpe and Grund 1986)

Despite the many studies in the literature directed towards the quantification of the effect of different operating conditions, BCRs are still not well understood. The phase-phase interactions should be studied intensively due to their importance in the system design and operation. Better mathematical and numerical models need to be developed to describe the phenomena inside the BCRs.

2.4 Membrane Bioreactors (MBRs)

2.4.1 History of MBR Development

The history of MBRs can be traced back to the 1960s, when the MBR process was first proposed by Dorr-Olivier Inc. which combined the use of a cross-flow membrane filtration loop and the activated sludge bioreactor (Kraume and Drews 2010). Although the idea of replacing the settling tank of the conventional activated sludge process was attractive, it was difficult to justify the use of such a process because of the high cost of membranes, low economic value of the product (tertiary effluent) and the rapid loss of filtration performance due to fouling. The breakthrough for MBRs came in 1989 with the idea of Yamamoto and co-workers to submerge the membranes in the bioreactor. The submerged MBRs worked with low suction pressure and made use of aeration to reduce fouling. This concept paved the way towards a significant reduction of capital and operating costs due to the reduction and simplification of equipment, and the abandonment of the energy demanding sludge recirculation loop (Judd 2011).

Nowadays, two MBR configurations are commonly applied in domestic and industrial wastewater treatment processes, namely, the submerged and side-stream configurations (Gupta *et al.* 2008). A submerged MBR is a reactor tank where the

filtration element is submerged inside (See Figure 2.4 (a)). The filtration system can consist of flat sheet, hollow fiber or tubular membranes, and can also incorporate an online backwash system which reduces membrane surface fouling by pumping membrane permeate back through the membrane. If the filtration element is installed external to the reactor and the treatment process relies on high TMP to maintain filtration, then it becomes the side-stream configuration as shown in Figure 2.4 (b). Submerged MBR systems are usually preferred compared to side-stream systems, especially for domestic wastewater treatment, since the energy demand of the submerged system can be up to two orders-of-magnitude lower than that of the side-stream systems. The submerged hollow fiber MBRs are attractive because the membranes are self-supporting and easy to be backwashed. Moreover, the fabrication of submerged modules is relatively simple and economical (Cui *et al.* 2003).

Figure 2.4 MBR systems (a) submerged configuration; (b) Side-stream configuration (Gupta *et al.* 2008)

The first application of the MBR technology to wastewater treatment was believed to be Yamamoto as well (Cui *et al.* 2003). In order to improve the membrane performance, Yamamoto and co-workers added a jet aerating device to provide intermittent vigorous aeration. To date, bubbling is considered as one of the major

mechanisms affecting membrane and system performances both hydraulically and biologically (Cui et al. 2003; Judd 2011).

All the parameters involved in the design and operation of MBR processes have an influence on the membrane fouling (Le-Clech *et al.* 2006). Generally four aspects are listed: (1) feed characteristics; (2) biomass characteristics; (3) membrane and module characteristics, such as the membrane pore size, materials as well as configurations and finally, (4) operating conditions, including the solid retention time (SRT), air injection etc. The detailed explanations can be found in Section 2.4.3.

2.4.2 Fouling in MBRs

In this section, the basic features of hollow fiber membrane are introduced first, followed by the description of primary fouling mechanisms.

2.4.2.1 Hollow Fiber Membranes

Membranes can be broadly classified into four categories in terms of membrane configurations: flat sheet, spiral-wound, tubular and hollow fiber. In this study, hollow fiber membranes were selected to study the flow conditions around the fibers as they are extensively used in water treatment processes to remove turbidity and in MBRs for wastewater treatment. The beneficial features include modest energy requirement, larger surface area per unit volume, flexible and low operation cost (Çulfaz *et al.* 2011). Although having a lot of excellent features, hollow fibers have the disadvantage of a high propensity to fouling due to poor hydrodynamics associated with the modules (Yeo *et al.* 2006). To ensure that hollow fiber membranes are used efficiently, it is important to design the module and its hydrodynamics in such a way that the fibers are fouled evenly (Mulder 1996).

Submerged hollow fibers can be aligned horizontally or vertically. The horizontal orientation has the disadvantage that bubbles can get trapped in gas pockets between groups of fibers (see Figure 2.5 (b)) whereas the vertical orientation favors the various mechanisms induced by rising bubbles, as illustrated in Figure 2.5 (a) and 2.6. Thus, in the MBR industry, hollow fibers are commonly aligned vertically (Cui *et al.* 2003).

Figure 2.5 Characteristics of bubbles with: (a) vertical and (b) horizontal fiber array (Cui *et al.* 2003)

Figure 2.6 Mechanisms of cake depolarisation - bubbles outside fibers (Cui *et al.* 2003)

It is believed that smaller fibers have a significantly better performance with bubbling (Wicaksana *et al.* 2005). However, it should be noted that there is a practical constraint on using very small diameter fibers because of the increased pressure drop on the lumen-side that causes a significant flux decrease along the fiber and added energy cost. Actually, it had been proven that there exists an optimum inner diameter for the hollow fibers for productivity maximization per unit energy, and a certain degree of fiber looseness and flexibility are very important (Chang *et al.* 2002). The advantage of loose fibers is shown in Figure 2.7. It is obvious that the suction pressure used for filtration is significantly lower for loose fibers compared with tight ones.

Figure 2.7 Suction pressure profiles for tight and loose hollow fiber bundles
(Chang *et al.* 2002)

2.4.2.2 Membrane Fouling

Le-Clech *et al.* (2006) reviewed and analysed more than 300 publications to provide a better understanding of the complex fouling mechanisms and propensities occurring in MBR processes. Essentially, fouling is the accumulation or adsorption

of solutes or particles onto the surface or into the pores of the membrane. Fouling causes flux declines or higher trans-membrane pressure, and modifies the retention properties of the membranes. It differs from concentration polarisation, which is a reversible accumulation of solutes or particles in the boundary layer. However, concentration polarisation can worsen fouling.

Generally there are three fouling mechanisms: (1) pore closure; (2) pore plugging; and (3) cake formation, as shown in Figure 2.8.

Figure 2.8 Membrane fouling mechanisms (Le-Clech *et al.* 2006)

Fouling in MBRs can be physical or biological, inorganic and organic (Bella *et al.* 2007). There can be small particles and colloids which close pores and/or deposit as a cake layer, or bacteria deposits which develop biofilms on the membrane surface. There can also be macromolecules such as proteins, natural organic matters and effluent organic matter. Besides, inorganic salts with low solubility precipitate on membrane surface if they are retained. A good understanding of the foulant structures in mixed species systems allows better control measures to prevent foulant build-up or to disengage the foulant layer. The lessons learnt from such studies are important for the understanding of fouling in MBRs (Le-Clech *et al.* 2006). In the present study, the cake layer formed on the surface of the membrane was studied, as the effect of air sparging is the most distinct on fouling mitigation for this type of fouling mechanism.

2.4.3 Factors Affecting Fouling Formation

All parameters involved in the design and operation of MBR processes have an influence on membrane fouling. Here, three categories of factors are defined: (1) feed and biomass parameters; (2) membrane features; and (3) operating conditions (Zhang *et al.* 2006). While some of these parameters have a direct influence on MBR fouling, many others result in subsequent effects on phenomena exacerbating the fouling propensity. The complex interactions among these parameters contribute towards MBR fouling, thus it is crucial to obtain an in-depth understanding of the biological, chemical as well as physical behaviour in MBRs in order to assess the fouling propensity of the membranes (Le-Clech *et al.* 2006).

2.4.3.1 Feed and Biomass Characteristics

- Nature of feed and concentration

Fouling in MBRs is mostly affected by the interactions between the membranes and the biological suspensions in the wastewater (Choi *et al.* 2005). The most striking effect of the wastewater nature is on the physicochemical change in the biological suspensions (Le-Clech *et al.* 2003; Jefferson 2004). For these reasons, the fouling propensity of the wastewater is indirectly taken into consideration during the characterization of the biomass (Le-Clech *et al.* 2006).

- Mixed Liquor Suspension Solids (MLSS) concentration

MLSS has a complex interaction with MBR membranes. Controversial findings about its effect on membrane filtration have been reported. If other biomass characteristics are not accounted for, the increase in MLSS concentration seems to have the most negative effect on the MBR hydraulic performance (Chang and Kim 2005). However, some authors also reported a positive impact (Defrance and Jaffrin 1999) or an insignificant effect (Hong *et al.* 2002). Thus, at present there is no clear correlation between MLSS concentration and any other foulant characteristics such as permeate flux and cake resistance. This might imply that the MLSS

concentration alone is a poor indicator of the biomass fouling propensity (Brookes 2003; Jefferson 2004). These authors recommended the use of fundamental operating parameters like the hydraulic retention time (HRT) and solid retention time (SRT) for prediction of foulant production (Le-Clech *et al.* 2006).

- Viscosity

Biomass viscosity is closely related to its concentration in MBRs (Yeom 2004). Increased viscosity reduces the efficiency of the mass transfer of oxygen and affect dissolved oxygen (DO) (Germain 2005). MLSS viscosity is important since it can modify the bubble size and dampen the movement of hollow fibers in submerged bundles (Wicaksana *et al.* 2006). The net result of increased viscosity is a greater rate of fouling.

- Temperature

Temperature affects the membrane filtration through its influence on the permeate fluid viscosity (Mulder 2000). The common approach to compare hydraulic performance obtained at different temperatures is to normalize the operating flux at a reference temperature (generally 25°C). On average, a greater deposition of fouling would occur on the membrane surface at lower temperatures (Rosenberger *et al.* 2006).

- Dissolved oxygen (DO)

The average level of DO in a bioreactor is controlled by the aeration rate which not only provides oxygen to the biomass but also limits the fouling formation on the membrane surface. The effects of DO on the MBR fouling therefore are multiple and include changes in biofilm structure, floc size distribution, etc. (Le-Clech *et al.* 2006). As a general trend, higher DO tends to lead to better filterability and a lower fouling rate. This can be explained by the lower specific cake resistance of the fouling layer which has larger pore sizes and greater porosity (Kang *et al.* 2003; Kim *et al.* 2006).

- Floc characteristics

With the significant changes in biomass characteristics from one plant to another, different biomass parameters affect MBR fouling with various propensities. These disparities are also partly due to the different analytical methods and instruments used in the reported studies. A large number of recent publications indicated that the biomass supernatant and its carbohydrate fraction are the main parameters affecting MBR fouling. However, the more detailed characterization of the supernatant and the fouling layer also revealed the significant role played by the protein fraction (Le-Clech *et al.* 2006).

2.4.3.2 Membrane Characteristics

- Pore size and distribution

Smaller pore membranes reject a wider range of materials, and the resulting cake layer features a higher resistance compared to large pore membranes. However, this type of fouling is more reversible and can be easily removed during the maintenance cleaning than fouling caused by internal pore clogging that occurs in larger pore membrane systems (Le-Clech *et al.* 2006).

The complex and changing nature of the biological suspensions present in MBR systems and the large pore size distributions of the membranes generally used in MBRs are the main reasons for the undefined general dependency of the flux propensity on pore size (Chang 2002; Le-Clech *et al.* 2003). There should be a dynamic layer of greater overall resistance created for more selective membranes at low flux conditions. In addition, increasing the pore size might decrease the deposition onto the membrane at the expense of internal adsorption. Long term trials confirmed this theory as progressive internal deposition eventually lead to catastrophic increase in resistance (Ognier *et al.* 2002; Le-Clech *et al.* 2003).

- Porosity/roughness

Fang and Shi (2005) proposed that membrane microstructure, material and pore size distribution all affected MBR fouling significantly. They conducted tests to investigate pore fouling of four different microfiltration membranes for the filtration of activated sludge. Although the four membranes were operated under the same constant pressure, the track-etched membrane, with its dense structure and small but uniform cylindrical pores, featured the lowest resistance due to pore fouling. In contrast, the other three membranes with interwoven sponge-like microstructures were more prone to pore fouling due to their highly porous network.

Kim *et al.* (2004) compared the performance of two micro-porous membranes prepared by the stretching method, and revealed the influence of the pore aspect ratio (mean major axis length/mean minor axis length) on fouling in a MBR. With both membranes, the average pore size and pure water flux were identical, but less fouling was observed with the membrane having the higher pore aspect ratio (elliptical pore) rather than with the circular pore membrane.

It was not possible to clearly determine the effect of roughness on MBR fouling. However, the large “filling-in points” present on rougher membranes were more prone to fouling than the fewer and smaller “crevices” observed on smoother membranes (He *et al.* 2005). Detailed discussion about the effects of membrane surface properties on cell attachment could be found in Kang *et al.* (2003).

- Membrane configuration

The current trend in MBR design for domestic wastewater treatment tends to favour submerged over side-stream configurations. In submerged MBR processes, the membrane can be configured as vertical flat sheet plates, vertical or horizontal hollow fibers or more rarely as tubes. Hollow fiber modules are generally cheaper to manufacture, allow high membrane density and can tolerate vigorous backwashing. At the same time, hollow fibers may require more frequent washing and cleaning while the fluid dynamics are easier to control for flat sheet and tubular membranes (Cui *et al.* 2003).

The packing density is an important parameter for submerged hollow fiber membranes. As summarized by Yeo *et al.* (2005 and 2006), the distance between adjacent membranes could directly impact the shear stresses and air demand. Besides, increasing the packing density could lead to severe clogging by gross solids and slower rise of bubbles, limiting their effects on fouling limitation. Experiments performed by Yeo *et al.* (2006) with a model bundle of nine fibers revealed that the overall module performance was much worse than that of an individual fiber. It was recommended that the packing density should be lower than 30% in order for the bundle to perform similarly to a single fiber.

- Hydrophobicity

Studies have shown that due to the hydrophobic interactions occurring between solutes, microbial cells and membrane materials, membrane fouling was expected to be more severe with hydrophobic rather than hydrophilic membrane (Chang and Lee 1998; Madaeni *et al.* 1999; Yu *et al.* 2005). However, it is now widely recognized that this parameter plays only a minor role during extended filtration periods since once fouled, the membranes' chemical characteristics become secondary to those of the sludge materials covering the membrane surface (Le-Clech *et al.* 2006).

- Materials

The occurrence of irreversible fouling is dependent on the membrane material, as some fractions of the organic matter present in the biomass express a higher affinity for certain polymeric materials. Despite having superior chemical, thermal and hydraulic resistances, ceramic membranes are not the preferred option for MBR applications due to their high cost (Le-Clech *et al.* 2006). Novel stainless steel membranes have shown good hydraulic performance and fouling recovery for anaerobic MBR wastewater treatment (Zhang *et al.* 2005). However, the large majority of the membranes used in MBRs are polymeric based (Le-Clech *et al.* 2006). Yamato *et al.* (2006) compared the fouling performance of polyethylene (PE)

and PVDF membranes, and indicated that PVDF materials had a better prevention of irreversible fouling and that PE membranes fouled more quickly.

2.4.3.3 *Operating Conditions*

- Aeration and cross flow velocity (CFV)

Since the origin of submerged MBRs, bubbling has been used as the strategy of choice to induce flow circulation and shear stress on the membrane surface. The use of bubbling to enhance membrane performance was extensively investigated and reviewed, and the effect of bubbling on submerged hollow fiber module was also assessed (Cui *et al.* 2003).

Bubbling alleviates issues related to the high packing density in hollow fiber bundles. However, it is still a challenge task for MBR designers to achieve an effective aeration throughout the population of fibers in a bundle (Yeo 2005). In the literature, it has been suggested conceptually that an optimum aeration rate exists beyond which a further increase will have no significant effect on fouling suppression (Ueda *et al.* 1997; Liu *et al.* 2003; Psoch and Schiewer 2005). Studies have also revealed that, for specific conditions, the filtration mode had more effect on MBR fouling than changes in aeration rates. An intensive aeration rate might also damage the floc structures, reducing their sizes, and releasing EPS in the bioreactor (Park *et al.* 2005; Park *et al.* 2005; Ji and Zhou 2006).

In a small cross flow cell, fouling decreased linearly with increasing CFV (up to 4.5m/s) and no optimum CFV was observed (Choi *et al.* 2005). A high CFV was more effective in reducing fouling in the MF rather than in the UF-MBR system, which was shown by analysing the mass, thickness and density of the fouling layer deposited in/on the membrane under different operating conditions (Le-Clech *et al.* 2006).

The determination of the CFV on the membrane surface induced by aeration can be difficult. Techniques employing the electromagnetic flow velocity meter, Particle Image Velocimetry (PIV), and Constant Temperature Anemometry (CTA), have

been used for liquid velocity estimation in submerged MBR reactors (Sofia *et al.* 2004; Yeo 2004; Le-Clech *et al.* 2006).

- Solid retention time (SRT)

The SRT (and consequently the F/M ratio) which ultimately controls the biomass characteristics is probably the most important operating parameter affecting the fouling propensity in MBRs. Operating a MBR at higher SRT leads inevitably to an increase in the MLSS concentration, but this in itself may not lead to greater fouling (JINSONG *et al.* 2006).

Extended SRT operation does not necessarily offer lower fouling. Other operating conditions such as the fluxes and aeration rates also have a major influence on the fouling propensity. The other difficulty with very high SRT is the raised viscosity that could attenuate the effect of bubbling (J.C. Orantes 2004; Han *et al.* 2005). As a general rule, the F/M ratio was recommended to be maintained below 0.5 (Ng and Hermanowicz 2005). The reasons suggested for the increased fouling rate at a very low SRT included the increased levels of EPS (Le-Clech *et al.* 2006).

It is believed that there is an optimal SRT, between the high fouling tendency of very low SRT operation and the high viscosity suspension prevalent for very long SRT. However, the difficulty relates to properly adjusting a pilot scale MBR to different SRTs and conducting a fair comparison. Criteria such as the designed rate of sludge and MLSS concentration recommended by the membrane supplier are more often used to define the working SRT.

- Fouling mechanisms in MBRs

The filtration test can be carried out under a constant flux or a constant TMP conditions. Very few studies have been reported for constant TMP operations, for which a rapid flux decline is expected to occur during the initial stage of the filtration. The rate of fouling then decreases before reaching a plateau. With a constant flux condition, however, the convection of foulant does not diminish. The fouling phenomenon self-accelerates and eventually creates a sharp increase with

TMP. Generally, fouling behaviour has been characterized by a three-step pattern: stage 1, conditioning fouling; stage 2, slow (steady) fouling; and stage 3, a TMP jump (see Figure 2.9).

Generally, in the first stage, colloid adsorption and initial pore blocking of new or cleaned membranes by organic substances is expected (Zhang *et al.* 2006). The intensity of this effect depends on the membrane pore size distribution and surface chemistry (and especially hydrophobicity) (Ognier *et al.* 2002). In stage 2, a small increase in the TMP is expected. For trials carried out over extended time periods, a noticeable change in the rate of TMP increase then arises after some critical period. This then leads to stage 3, for which a significant flux decrease is expected at some specific locations. Pollice *et al.* (2005) introduced two parameters to represent the operating conditions: the critical time (t_{crit}) and the fouling rate ($dTMP/dt$). In this study, the second parameter, i.e., $dTMP/dt$ is adopted to indicate the degree of fouling accumulation in the reactor.

Figure 2.9 Fouling Mechanisms for an MBR operated at a constant flux condition (Zhang *et al.* 2006)

Overall, the design of a BCR shall be conducted with caution. A proper range of air flow rate is essential for the air flow patterns produced inside the BCR. Besides, the air sparger configuration is also another important consideration. When expanding to the design of a three-phase MBR, factors including the viscosity and the nature of the feed should be taken into account as the flow is usually opaque.

In the present study BCRs with a submerged hollow fiber membrane module was mainly investigated. Three different approaches, namely, the membrane filtration test, submersible pressure sensing measurement and the computational fluid dynamics (CFD) modelling, were employed to explore the hydrodynamic characteristics inside the BCR. In Chapters 3, 4 and 5, the three approaches were illustrated with the literature review, methodology, results and discussion given in detail, respectively.

3. Membrane Filtration Tests*

3.1 Introduction

In the air sparged membrane system shear stress is the critical parameter governing the foulant removal and enhancing the permeate flux. Shear stress is related to the viscosity and the velocity gradient at the membrane surface, which can be controlled by the hydrodynamic conditions of the reactor (Chan *et al.* 2007). However, the conditions are very complex due to the gas-liquid two-phase interactions, and they are also influenced by the membrane module configuration, gas sparger configuration as well as the reactor geometry (Meng *et al.* 2009).

In this chapter the hydrodynamic effect of air sparging on the membrane performance in a laboratory scale cylindrical BCR is investigated. The relationship between the hydrodynamics induced by air sparging and the membrane fouling rate is explored quantitatively. Factors including the air flow rate, bubble-induced turbulent flow, air sparger geometry, membrane module orientation as well as the nature of feed solutions are examined.

3.2 Air Sparging

* This chapter has been published in the Journal of Water Research (Xia, L., A.W.-K. Law, and A.G. Fane, *Hydrodynamic effects of air sparging on hollow fiber membranes in a bubble column reactor*. Water Research, 2013. **47**(11): p. 3762-3772.)

The concept ‘air sparging’ was first introduced by Hitachi in electro dialysis systems in 1986. In order to prevent the formation of particle deposits or concentration polarization, gas is injected in the concentrate during filtration or cleaning processes (Cabassud *et al.* 2001). Aeration used in MBR systems has three major roles: (1) providing oxygen to the biomass; (2) maintaining the activated sludge in suspension; and (3) mitigating fouling by constant scouring of the membrane surface (Le-Clech *et al.* 2006). In the following the bubble characteristics in BCRs are discussed first and different flow regimes with increasing air flow rate are described subsequently. Next, the use of bubbling to mitigate fouling is elaborated and finally, recent developments on the air sparger configuration designs are summarized.

3.2.1 Bubble Characteristics

When gas is injected into a stationary liquid, bubbles are formed and move upward driven by buoyancy. The bubble shape changes from spherical to ellipsoidal and then to a spherical cap as the bubble size increases.

Figure 3.1 depicts these bubble shapes and the associated streamlines and wakes (Cui *et al.* 2003). Small spherical bubbles are typically less than 1 mm in diameter and obey Stokes’ Law. As such, they do not induce any disturbance to the flow around them. Ellipsoidal bubbles range from 1.5 to 15 mm. They appear to rock back and forth, and leave helical vortices in their wakes. Spherical cap bubbles are larger than 15 mm. The primary wakes induced by these bubbles are about 4.5 times the volume of the bubbles. A strong secondary flow effect can be created by these bubbles.

Figure 3.1 Bubble shapes and wakes (Cui *et al.* 2003)

Tadaki and Maeda (1961) proposed empirical correlations for the bubble deformation factor, d/a , for the three bubble shapes, respectively:

$$d/a = 1, \quad \text{Re } M^{0.23} < 2 \text{ (Spherical)} \quad (3.1)$$

$$d/a = 1.14(\text{Re } M^{0.23})^{-0.176}, \quad 2 \leq \text{Re } M^{0.23} < 6 \quad (3.2)$$

$$d/a = 1.36(\text{Re } M^{0.23})^{-0.28}, \quad 6 \leq \text{Re } M^{0.23} < 16.5 \quad (3.3)$$

$$d/a = 0.62, \quad 16.5 \leq \text{Re } M^{0.23} \text{ (Spherical cap)} \quad (3.4)$$

where Re is the Reynolds number and $M (= g \mu_L^4 / \rho_L \sigma^3)$ is the Morton number. Here g denotes the acceleration due to gravity and σ is the surface tension (Miyahara *et al.* 1988). The summary of the liquid properties is listed in the following table:

Table 3.1: Summary of the liquid property (Tadaki and Maeda 1961)

Liquid	Temp. [°C]	σ [dyne/cm]	μ [g/cm·sec]	ρ [g/cm ³]	M ($=\eta\mu^4/\rho\sigma^3$)
7% sodium chloride and water	10.6	75.9	0.014	1.07	8.04×10^{-11}
3.5% sodium chloride and water	12.5	75.0	0.0127	1.04	5.89×10^{-11}
Water	12.7	73.8	0.0122	1.00	5.41×10^{-11}
75% glycerine and water	15.7	58.7	0.50	1.20	2.52×10^{-4}
55% glycerine and water	12.5	61.0	0.10	1.10	3.75×10^{-7}
8% ethanol and water	12.8	52.8	0.018	0.970	7.08×10^{-10}
Nitrobenzene	14.3	44.6	0.0234	1.082	3.39×10^{-9}
0.5% EPAN-450 [†] and water	12.4	41.2	0.0120	1.00	2.91×10^{-10}
30% ethanol and water	12.7	33.8	0.0260	0.960	4.30×10^{-8}
80% acetic acid and water	11.7	32.1	0.0350	1.07	4.16×10^{-8}
Isoamyl alcohol	12.5	24.5	0.0560	0.813	8.03×10^{-7}
Ethanol	12.5	23.1	0.0141	0.796	3.88×10^{-9}
Toluene	11.8	29.5	0.0065	0.912	9.52×10^{-11}
Ethyl acetate	13.0	25.0	0.00489	0.909	3.94×10^{-11}

[†] Additional polymer of ethylene oxide with poly propylene oxide, nonionic surfactant

3.2.2 Bubbling for Fouling Mitigation

3.2.2.1 Fouling Alleviation Mechanism

In ultra-filtration (UF) the mechanisms of fouling control by bubbles have been identified as follows (Cui *et al.* 2003):

1. Bubble induced secondary flow

Secondary flows and wakes are generated by bubbles passing. The moving bubbles can disrupt the mass transfer boundary layer and promote local mixing near the membrane surface. The effect of shear was examined by Vera *et al.* (2000), who reported a linear relationship between two dimensionless parameters, the shear number (N_s') and the resistance number (N_f), which were plotted for gas sparging filtration of a dextran solution, ferric hydroxide suspensions, and secondary effluent using an inorganic tubular membrane. Their results showed that gas sparging could completely eliminate the resistance for filtration of a hydroxide suspension and dextran solution in the slug flow region but only partly eliminated the filtration resistance caused by secondary treated wastewater.

2. Physical displacement of the concentration polarization layer

Small gas bubbles can penetrate the concentration polarization layer and detach loose particles from membrane surfaces. Bellara *et al.* (1996) conducted experiments to investigate the effect of gas-liquid two-phase flow on controlling the concentration polarization in a UF hollow fiber system while using albumin and dextran as the test media. The test results showed that the liquid film thickness between the membrane wall and the slug in small diameter tubes was less than the computed mass transfer layer thickness in single-phase liquid flow at the same liquid flow rate.

3. Pressure pulsing caused by passing bubbles

The injection of gas bubbles increases the mean pressure in a membrane system. The bubble movement also causes pressure fluctuations in the liquid. It is believed that both factors contribute to a flux increase (Cui *et al.* 2003).

4. Increase in superficial cross-flow velocity

High gas flow rate injection can increase the cross flow velocity, which will result in a flux increase. However, the improvement in membrane performance is only significant at high gas flow rates, which is energy intensive. There exists an optimum aeration rate beyond which a further increase has no significant effect on fouling suppression (Ueda *et al.* 1997). However, there were not sufficient quantitative data in his study that clearly illustrated this phenomenon.

3.2.2.2 *Flow Dynamics*

Until recently, the focus on membrane fouling alleviation by air sparging was placed on the mean flow circulation in the reactor induced by bubbling. For example, Liu *et al.* (2000) studied the hydraulic characteristics in a submerged MBR system by measuring the average CFV distribution along the membrane surface. They concluded that a critical CFV of 0.3 m/s existed, below which the TMP increased sharply. Kang *et al.* (2008) developed an integrated numerical approach to quantify the hydrodynamics in both prototype and pilot MBR systems, with simulations targeting the mean flow circulation inside the MBRs. Khalili-

Garakani et al. (2011) investigated the mechanisms that led to flux enhancement and fouling reduction using both experimental approaches and numerical simulations. For turbulent flows, the time-averaged equations-of-motion introduces the ‘Reynolds stress’ term. In order to solve the turbulence flow, the Reynolds stress term should be determined using some methods. In this case, they computed numerically the mean gas and liquid velocities using the k - ϵ turbulence closure, which is one of the common ways to solve the turbulence, and suggested that the shear stresses on the membrane surface had a high correlation with the filtration resistance. These studies implied that the mean shear stress on the membrane is the key towards the reduction of fouling on the membrane surface.

In addition to the mean flow circulation, there is now an increasing recognition that turbulence fluctuations induced by bubbling also play an important role in membrane fouling control. Yeo et al. (2006) examined the filtration performance of submerged hollow fibers with and without bubbling, and quantified the relationship between the fouling performance and the unsteady flow conditions. His results clearly illustrated the importance of turbulent shear stresses on membrane fouling mitigation. Chan et al. (2011) found that the fouling rate for membranes subject to transient shear conditions was lower than for membranes subject to constant shear conditions. Various approaches have been proposed to enhance the fluid instability including turbulence promoters, pulsation, rotating membrane filter as well as secondary flows (Xu *et al.* 2002). In terms of aeration schemes, various design configurations have been developed for which the major attempt has been to maximize the ability of the hydrodynamic characteristics in the reactor for effective membrane fouling control. The detailed descriptions of these recent air spargers are given in the following section.

3.2.3 Air Sparger Configurations

While fundamental studies and mathematical models have been applied to well-defined membrane configurations, the effect of bubbles on submerged hollow fiber

modules is still not well understood. The addition of a biomass mixture (non-Newtonian fluid containing solutes, colloids and particulates) makes the analysis more difficult and challenging (Le-Clech *et al.* 2006). Regarding the energy cost, aeration is an important consideration for the MBR operation (Buer and Cumin 2010). Since both coarse and fine bubbles are used for scouring the membrane surface, the cost of aeration can be up to 90% of the total costs (Braak *et al.* 2011). Therefore, there has been intense interest to optimize the aeration scheme in order to minimize the energy consumption and operating costs. However, the issue is intricate; for example, deriving the optimal bubble size for a particular reactor is not straightforward. Altering the flow regimes at the membrane surface towards a more turbulent flow pattern was beneficial for foulant removal (Flemming 1995). In terms of aeration scheme, continuous injection is typically adopted in the beginning of MBR development. It achieved good fouling control but the energy consumption was excessive. Since then, many innovative schemes have been developed with good results and a summary can be found in Judd (2011).

First is the Zenon's system (Jeffery 2005) which makes use of cyclic aeration. In this aeration mode, the blowers are operated continuously at a fixed speed and cyclic aeration is achieved by switching the airflow from one air header to the other using pneumatically actuated valves. This is preferred over continuous aeration primarily for technical and commercial reasons. Technically, continuous aeration often results in air channelling through the module and can only thoroughly clean portions of the module. Introducing shear and airflow instability into the process through cyclic aeration eliminates channelling (Wu and He 2012). Continuously operating aerators also have a tendency to plug, but effectively flushing the aerators through the cyclic aeration action significantly alleviates the problem.

In Memcor's aerator units (Judd 2011), the air bubbles entrained in the mixed liquor are introduced into the module using a patented two-phase mixing system (see Figure 3.2). This creates high shear conditions, thereby minimizing solids concentration polarization at the membrane surface near the jet. Higher membrane packing densities are permitted as a result.

Figure 3.2 Memjet module with integral jet aerator (Judd 2011)

This is followed by Koch's PURON's units (Judd 2011) in which scouring air is injected between the filaments intermittently by means of a central air nozzle at the module base so as to limit the degree of clogging there. Later, Asahi Kasei Chemical Corporation launched another MBR membrane module (Judd 2011) in which air is introduced via a separate aerator and directed into the element by a skirt fitted to the element base (see Figure 3.3).

Figure 3.3 Microza EFM HF membranes aerated module schematic (Judd 2011)

To prevent rapid membrane fouling, especially in a high suspended solids environment where a continuous fluid transfer is critical, Siemens provided fluid transfer in the form of mixed liquor agitation and air scour energy through their innovative MemPulse MBR (Siemens 2009). This technology introduces air continuously to the base of the module with irregular pulses that enter the membrane fiber bundles. The mixed liquor (from a separate biological system) is drawn into the bottom of the membrane modules through an airlift effect. The air bubbles blend with the mixed liquor and rise between the individual membrane fibers, providing scouring energy at the membrane surface as well as fluidizing the membrane surface to prevent solids accumulation.

Generally, all the air sparger configurations tend to maximize the turbulence inside the reactor since the flow fluctuation plays an important role on fouling alleviation (Braak *et al.* 2011). Beside the air sparger design, other factors like the reactor geometry, air flow rate, bubble size as well as the membrane module configuration all have decisive effects on the filtration performance (Drews 2010). In this study, some of the factors are investigated to contribute to a better understanding of the

bubble-induced hydrodynamics. In the following sections the experimental setup and methodologies are described before discussing the results.

3.3 Experimental Methodologies

3.3.1 Experimental Setup

3.3.1.1 Apparatus

Figure 3.4 (a) shows the schematic diagram of the experimental setup for the membrane filtration tests with air sparging. The setup consisted of a cylindrical BCR, a hollow fiber membrane module and a data logging system. The BCR had an inner diameter of 140 mm and an outer diameter of 152 mm. The feed solution was typically filled to a depth of 1 m. A hollow fiber membrane module was placed at both 0.3 and 0.6 m above the bottom of the reactor to examine the effect of the hydrodynamic conditions at these two locations. Constant flux conditions were maintained by carefully varying the rotational speed of a peristaltic pump (Spectral-Teknik (S) Pte Ltd) which was controlled by a programmable logic controller (PLC) system. The Tygon tubing used to extract the permeate should be replaced regularly as the Tygon tubing tends to lose its flexibility such that it could cause the reduction of permeate flux gradually. Under the constant flux condition, the TMP changes of the membrane module were then monitored for one hour during which time raw data were logged and stored in a computer system for post-processing. During the experiment, a sensor was suspended at the top of the reactor with its tip touching the feed solution surface to regulate the feed level. It would send a signal to the PLC system to activate a feed pump if the feed level dropped too low. The air flow rate was determined by an air flow meter manufactured by Dwyer Company. A pulse dampener was installed to eliminate the pulsation of the pumping system.

Figure 3.4 Experimental setup for the membrane filtration tests

Figure 3.4 (b) and (c) show the central and alternate aeration schemes used in the membrane filtration tests, respectively. The ball-shaped corundum air stones used in the current study is made from sintering. They are great anti-oxidation, rigid and easy positioning through submersion, which can provide 360 degrees air supply. For the central aeration scheme, a perforated air stone with a diameter of 30 mm was mounted at the central bottom of the reactor for air bubble generation. This type of point air source is commonly used in the environmental industry for wastewater treatment due to its simplicity, high air flux and easy control (Liu and Tay 2004; Gan *et al.* 2011). For the alternate aeration scheme, two air stones of 20 mm diameter each were placed at the opposite end of the diameter of the BCR. Particular attention was paid towards the switching interval and its effect on membrane fouling. The curtain shape membrane module was aligned with as well as perpendicular to the alignment of the two air stones. For both locations, three alternate aeration patterns were tested with a switching interval of 0.5, 1 and 2 minutes. When the switching interval was specified, one of the air stones would

release air for the period first, after which it would then be turned off and the other air stone was activated. The switching would repeat itself for one hour. Thus, although the air sparging remained continuous, the hydrodynamics was very much different from the central aeration scheme. Figure 3.5 shows a picture of bubbly flow taken at the 0.6 m location with the central aeration scheme and the air flow rate of 2.0 L/min. It could be observed that the bubbles were dispersed in the continuous liquid phase. They varied widely in size but were typically near spherical. Note that only the homogenous bubbly regime was considered in this study; the experimental range did not include heterogeneous regimes with breakup or coalescence of the bubbles.

3.3.1.2 *Membrane Specifications*

Figure 3.6 shows the hollow fiber membrane module used in the membrane filtration tests. The hollow fibers were PVDF UF membranes with 1.0 mm inner diameter and 2.0 mm outer diameter, and of approximately 0.1 μm pore size (MemStar Technology Ltd). The module consisted of two custom-made polyvinyl chloride (PVC) membrane holders and a detachable stainless steel membrane stand. Each fiber bundle was composed of twenty hollow fiber membranes of 10 cm effective length each. The hollow fibers were attached to a membrane support frame with the base end sealed with epoxy. High quality permeate was then extracted from the upper part of the module using a peristaltic pump.

When multiple fiber bundles were used, a series of permeability tests were conducted to select the appropriate bundles for the experiments. The major objective of these permeability tests was to increase the flux stepwise, and record the steady state value of the suction pressure which was normally 30 minutes after starting the measurements. The suction pressure versus flux curve was plotted, and the best fit line was obtained. The gradient of each line was then the permeability of the membrane module. The permeability values of different membrane modules were compared to make sure that the initial TMPs were close to each other. Those that showed large deviations from the initial TMP values were eliminated for use.

Beyond this, the measurement of the water permeability determined after backwashing whether the hollow fiber membranes could still be re-used or not.

When the filtration process ended, the hollow fiber membrane bundle was backwashed with permeate by reversing the rotating direction of the peristaltic pump for about 10 minutes. Subsequently, 1% of Terg-A-Zyme enzymatic detergent solution provided by Sigma-Aldrich was used to wash the module for another 10 minutes. After that, the module was rinsed with Milli-Q water to wash off the chemicals attached to the membrane surface, and finally the module was immersed in deionised water for further use. The measurement of water permeability after cleaning was necessary since the hollow fiber membrane would not be suitable for further use if its permeability could not be fully recovered.

It had been proven that there exists an optimum inner diameter for the hollow fibers for productivity maximization per unit energy; in addition a certain degree of fiber looseness and flexibility is very important (Cui *et al.* 2003). However, since the major focus in the present study was to correlate the hydrodynamics induced by air sparging with the membrane fouling performance, the fibers were thus glued tightly, i.e., 0% looseness, to eliminate the turbulent effect induced by the fiber movement.

3.3.1.3 Feed Solutions

Two kinds of feed solutions, i.e., the bentonite (Sigma-Aldrich) and yeast (Bake King's brand) solutions were tested since both of them are widely used as synthetic feed media for laboratory testing. Bentonite is an absorbent aluminum phyllosilicate. It is essentially impure clay consisting mostly of montmorillonite, which is a dioctahedral smectite mineral with a layered crystal structure with negative surface electrical charges arising from isomorphic substitutions of magnesium for aluminum in the crystal sheets. The clay swells in water by taking up inter-layer water, eventually leading to complete dispersion. Thus the size of montmorillonite particles is not very well-defined, but they may be regarded as flat plate-like particles of lateral dimensions on the order of 1 μm and thickness on the order of 10 nm (Ní Mhurchú and Foley 2008). Bentonite clay has been proven to have complex

rheological characteristics; the non-Newtonian nature of bentonite suspension can be expected to have a pronounced effect on the filtration characteristics (Benna et al. 1999; Güngör 2000; Mahto and Sharma 2004). Many studies have been conducted to investigate the fouling mechanisms using either dead-end or cross-flow filtration with the bentonite suspension as the feed solution (e.g., Hamachi and Mietton-Peuchot 1999; Benna *et al.* 2001).

As for yeast, it is a eukaryotic microorganism classified in the category of fungi. Yeast size is typically 3-4 μm in diameter but can vary greatly depending on the species. A notable feature of yeast is the large proportion of nitrogenous substances which it contains. The quantity varies a good deal according to the conditions of nutrition under which the yeast has been grown, but in general, more than one-half of the dry substances consist of proteins and other nitrogenous bodies. The proximate constituents other than nitrogen compounds include glycogen, gum, mucilage, fat, resinous matters and cellulose, together with a good proportion of mineral ingredients (Simmonds 1919). Membrane filtration tests using yeast are of practical importance since yeast is one of the most important hosts for genetic modification for bio-product manufacturing (Sur and Cui 2005).

In the experiments the same mass concentration of 0.65 g/L was used for both feeds. The feed solution was prepared every other day with 10 grams of bentonite or yeast powders weighted and mixed with water in a beaker. A stirring machine was used for mixing to assure a uniform dispersion of the particles. After that, the concentrated solution was poured into the reactor, and clean water was added to fill the liquid height to 1 m. Since the bentonite particles swelled once mixed with water, the inorganic solution was usually prepared a day ahead to make sure that majority of the solid particles no longer increased in size.

The dynamic viscosity of the bentonite solution was also tested before the filtration experiments. Generally, the dynamic viscosity can be computed as a ratio of the shear stress over the shear rate. Figure 3.7 shows the variations of dynamic viscosity for clean tap water and three different concentrations of bentonite

solutions: 0.65, 1.3 and 2.6 g/L. The dynamic viscosities were found to be 8.83×10^{-4} , 9.27×10^{-4} , 9.33×10^{-4} , 9.80×10^{-4} Pa·s correspondingly. Essentially the values were similar and the difference in viscosities between the clean water and the 0.65 g/L bentonite solution was only 2.57%. Thus it could be expected that the flow conditions would also be similar in clean water and these low concentration bentonite solutions when subjected to air sparging.

Owing to the cylindrical geometry of the BCR and the non-transparent nature of the feeds being tested in this study, the direct measurements of the velocity distribution in the reactor with the feed solutions were difficult. However, since a relatively low concentration of feeds was used, the flow pattern with the feed solutions can be assumed to be similar to the clean water. Thus the bubble swarm could be studied through direct visual observation of the bubble movement in clear water. The extent of fouling control due to different flow conditions was mainly represented by the variation of membrane fouling rates.

Figure 3.5 Picture of the bubbling flow with the air flow rate of 2.0 L/min at 0.6 m location

Figure 3.6 Hollow fiber membrane module used in the membrane filtration tests

Figure 3.7 Variations of dynamic viscosity for the clean tap water and three different concentrations of bentonite solutions of 0.65, 1.3 and 2.65 g/L

3.3.2 Operating Conditions

- Flow regimes in the BCR

As mentioned earlier, two locations of 0.3 and 0.6 m from the bottom of the reactor were investigated. These two locations were chosen due to the differences in hydrodynamic conditions. It had been observed that with the central aeration scheme, the hydrodynamics would vary along the height of the BCR as illustrated by the instantaneous and time-averaged flow field sketched in Figure 3.8. Gan *et al.* (2011) measured the two-phase hydrodynamics using the Phase Doppler Anemometry (PDA) technique using the same bubble column geometry with clean water as the feed solution. They found that in the region from the tip of the air stone $z = 50$ mm to about $z = 300$ mm, the injected bubble swarm rose from the air sparger forming a bubble plume. Meandering flows were dominant and the instantaneous circulation cells were not symmetrical about the centreline. Towards the middle portion of the reactor, the flow became more orderly. Near the surface, the bubbly flow was more or less uniform, and bubbles were well distributed across the entire cross-section (see Figure 3.8 (a)). For the time-average behaviour of the flow pattern as shown in Figure 3.8 (b), the bubble plume was developing along the z -axis and in the middle portion, the bubbles were concentrated in the central core and the flow became relatively uniform. Near the air sparger, large scale circulation existed due to the wandering of the bubble plume and hence fewer bubbles were concentrated except for some tiny ones. In this regard the two locations were picked with 0.3 m located in the meandering region and 0.6 m in the developed region. Thus, due to the difference in the hydrodynamic behaviour at these two locations, the effects of the flow regime on the membrane filtration performance can be characterized and optimized for fouling minimization.

- Aeration schemes

In terms of aeration schemes, as mentioned earlier, continuous injection is typically adopted in the beginning of MBR development. It achieves good fouling control, but the energy consumption can be high. Since then, many innovative schemes have

been developed with good results. These include the alternate aeration scheme in which bubbling is switched alternately between the two sides of the membrane module. It was proven that alternating the injection of air was highly beneficial for the overall filtration system performance (Guibert *et al.* 2002). Hence, in the present study, in addition to deploying a single air stone to provide continuous bubbling, a two air stone configuration was also employed to generate the alternate bubbling pattern. Both the central and alternate aeration schemes are commercially applied in the industry at present, and the two air stone configuration was adopted with the intent to mimic the basic switching principle.

Figure 3.8 Sketch of the flow field with central aeration in the BCR
(From Gan *et al.* 2011)

Figure 3.9 depicts the transient flow conditions in the BCR when the alternate aeration scheme was applied. At the 0.3 m location, the bubble plume was developing spatially but the plume adhered to the side of the membrane module due to the near wall placement of the two air stones. Large flow circulation could be observed near the air sparger region. Moving up to the 0.6 m location, the bubble plume became developed with bubbles dispersed throughout the cross-section of the reactor. This would continue for the switching interval before the bubbling air stone was changed to the opposite one. The same testing locations, namely, the 0.3 and 0.6 m to the bottom of the reactor were chosen since 0.3 m was located in a region with large flow circulation while the 0.6 m location was located where the flow behaved more regular.

Figure 3.9 Transient flow condition induced by alternate aeration in the BCR

- Permeate flux

The type of fouling considered in the filtration experiments was primarily surface fouling. Hence, our main interest was on the hydrodynamic conditions in the vicinity of the membrane surface caused by bubble movement, and the detailed deposition mechanisms in the membrane surface were not considered. Continuous filtration was carried out within a relatively short operating time (~ 1.0 hour), and a permeate flux of $71.6 \text{ L/m}^2\text{hr}$ was adopted in the experiments to produce the cake layer. During the filtration process, the highest TMP did not exceed 60 kpa, which was within the maximum operating pressure recommended by the manufacturer for the hollow fiber membranes.

- Membrane module orientation

When the alternate aeration scheme was utilized in the membrane filtration experiments, two different module orientations were tested: alignment with the two air stones (0 degree) and perpendicular to the alignment (90 degree) (see Figure 3.10). The objective was to investigate the effect of the direction of cross-flow to the membrane curtain on the fouling performance. In this case the 1.0 minute switching interval was examined.

Figure 3.10 Membrane module orientations: (a) 0 degree and (b) 90 degree

3.4 Results and Discussion

3.4.1 Measurements and Calculations

- TMP and fouling rate

Generally, the filtration flux and TMP are the best indicators of membrane fouling. The TMP is calculated as follows:

$$P_{TMP} = \frac{(P_f + P_r)}{2} - P_p \quad (3.5)$$

where P_{TMP} is the TMP, P_f is the feed pressure, P_r is the retentate pressure and P_p is the permeate pressure. In the present study the TMP values were measured under constant flux conditions and the temporal rise in TMP was used to assess the fouling rate of the system as follows:

$$\frac{dTMP}{dt} = \frac{dP_s}{dt} \quad (3.6)$$

where P_s is the suction pressure. In the data analysis the fouling rates were compared rather than the absolute values of TMP themselves.

3.4.2 Continuous Air Sparging

3.4.2.1 *The Effect of Flow Regime*

We first examined the effect of air flow rate on the membrane filtration performance with a 0.65 g/L concentration of bentonite solution using the central aeration scheme. Figure 3.11 illustrates the development of the fouling rate at a

constant flux of 71.6 L/m²hr with various air flow rates at the 0.3 and 0.6 m locations. Firstly, we compared the membrane performance at the two locations with the increase in air flow rate. Initially, the fouling rate at the 0.3 m location (~15.84 kpa/hr) was higher than that at the 0.6 m location (~12.24 kpa/hr) when the air flow rate was around 0.1 L/min. Along with the increase of aeration supply, a lower fouling rate was found at the 0.3 m location compared with the 0.6 m location. The quantitative results suggested that the meandering flow regime dominant at the 0.3 m location was more beneficial to fouling alleviation. Compared to a relatively steady circulation pattern at the 0.6 m location, the unsteady flow condition at the 0.3 m location was more effective in eroding the deposition layer. Hence, for a specific MBR system, the hydrodynamics should be characterized and the effect maximized such as by immersing the membrane modules in the highly turbulent region to suppress fouling.

3.4.2.2 *The Effect of Air Flow Rate*

Next, we studied the variations of fouling rates with different air flow rates. It could be observed from Figure 3.11 that with the increase of the air flow rate, the effectiveness of bubbling towards membrane fouling alleviation could be categorized into three stages. When the air flow rate was small and below 0.6 L/min, there was a fast reduction stage in which the fouling rate decreased substantially when the air flow rate just increased slightly. This was then followed by a slow mitigation stage from 0.6 to 1.2 L/min, whereby the fouling rate continued to decrease but at a much slower pace compared to the first stage. When the air flow rate was further increased beyond 1.2 L/min, however, a limiting stage was reached such that no further reduction of the membrane fouling rate could be observed. The above quantitative results suggested that introducing a small amount of air into the membrane filtration system could improve the membrane performance to a great extent, as illustrated in the first stage (see Figure 3.11). This is also well recognized in the industry, and explains the wide popularity of using air sparging in MBRs for membrane fouling control. Additionally, the results confirmed the existence of an optimal air flow rate for the air sparging of a membrane system beyond which further increases have no benefit on fouling mitigation. Many earlier researches had

also pointed to the existence of this optimal aeration value (e.g., Ueda *et al.* 1997; Sofia *et al.* 2004; Le-Clech *et al.* 2006). Clearly, the determination of this optimal air flow rate for a specific configuration of a membrane filtration system is of great importance in order that the energy cost for the air sparging can be optimised.

Figure 3.11 Fouling rates for 0.65 g/L bentonite solution at the 0.3 and 0.6 m locations (Flux = 71.6 L/m²hr)

3.4.3 Alternate Air Sparging

3.4.3.1 The Effect of Flow Regime

Figure 3.12 shows the variations of fouling rate at the 0.3 and 0.6 m locations when the three switching intervals of $T = 0.5$, 1.0 and 2.0 minutes were employed for a

0.65 g/L bentonite solution. First of all, the effects of flow regime at the two locations were compared. With $T = 0.5$ min, the 0.6 m location showed a better filtration performance when the air flow rate was higher than 0.6 L/min. When T was increased to 1.0 min, the variations of fouling rate for the two locations became more or less similar. A local minimum was found for both locations at which the magnitudes were about the same as the asymptotic minimum fouling rate. When the switching interval was further extended to $T = 2.0$ minutes, the two locations showed substantial fouling reduction with the increase of air flow rate and a lower fouling rate was found at the 0.6 m location.

Figure 3.12 Variations of TMPs with alternate aeration at 0.5, 1 and 2 minutes switching intervals at the 0.3 and 0.6 m locations for a 0.65 g/L bentonite solution

The quantitative results suggested that for either a shorter (0.5 min) or longer (2.0 mins) switching interval, 0.6 m was a more desirable location to place the membrane module due to lower fouling rates. However, the occurrence of a local minimum for both locations when 1.0 min was employed implied that a very small air flow rate was able to achieve an effective fouling control. Under this

circumstance, where the membrane module was placed became less critical since either location could provide a good hydrodynamic condition for fouling control. In this regard, the aeration demand can be significantly reduced. The current results also demonstrated in quantitative details the complexity of the two-phase hydrodynamics of the BCR.

3.4.3.2 *The Effect of Air Flow Rate*

The membrane filtration performance was also studied with the increase of the air flow rate. From Figure 3.12, it is obvious that three stages can be distinguished for the fouling performance with the increase of air flow rate at the 0.3 m location, namely, fast reduction, slow mitigation and limiting stages. Generally, 0.5 and 1.0 minute alternate aerations showed similar decreasing trend as did the central aeration in terms of fouling rate, whereas the fouling rates were much higher for the 2.0 minutes alternate aeration when a small air flow rate (<0.6 L/min) was applied. Between 0.6 and 1.5 L/min, the fouling rate for the three switching intervals decreased with the increase of air flow rate but at a relatively slower pace. Beyond 1.5 L/min, all three switching intervals exhibited a similar fouling rate trend, i.e., towards an asymptotic minimum fouling rate. The experimental results again confirmed that an optimal air flow rate existed for the alternate aeration scheme similar to the central aeration scheme. The optimal air flow rate was roughly about the same, namely, 1.5 L/min for all the three switching intervals.

For the 0.6 m location, generally, all the three switching intervals showed relatively lower fouling rates with the increase of air flow rate compared with the 0.3 m location. Among them, the 2.0 minutes switching interval had lower fouling rates compared with the 0.3 m location when a small air flow rate was applied (< 0.5 L/min). Moreover, the 0.5 minute switching interval achieved a local minimum fouling rate between 0.8 and 1.5 L/min for both locations. The results suggested that besides the magnitude of the air flow rate which was identical for the two locations, the local hydrodynamics due to the alternate aeration also played a significant role on the membrane fouling performance. The occurrence of the local minimum for both locations again demonstrated the complexity of the hydrodynamic conditions inside the bubble column driven by alternate aeration.

3.4.3.3 *The Effect of Switching Interval*

From Figure 3.12, the fouling rates were much higher for the 2.0 minutes alternate aeration when a small air flow rate (< 0.6 L/min) was applied. The large fouling rate obtained for the 2.0 minutes alternate aeration showed that the switching interval cannot be chosen randomly and must be carefully determined. In general, the curves for 0.5 and 1.0 min showed similar trends but the magnitude of the local minimum was smaller for 1.0 min switching interval. Hence at the 0.3 m location, a 1.0 min switching interval is more effective in eroding the cake layer compared with the other two switching intervals.

For the 0.6 m location, all the three switching intervals showed relatively lower fouling rates with the increase of air flow rate compared with the 0.3 m location. The superiority of 1.0 min switching interval no longer existed but the local minimum at 1.0 min still served as the desirable operating condition for better fouling control due to low fouling rates and small aeration demand.

From Figure 3.12, it is clear that the flow regime, air flow rate as well as the switching interval all play a role in influencing the hydrodynamics of the BCR when the alternate aeration scheme was employed. Although the flow behaviour became much more complex with alternate aeration, it was found that an optimal air flow rate existed for both aeration schemes. The magnitude of this optimal air flow rate might change when the operating conditions change, however, the three-stage membrane filtration performance with the increase of air flow rate is always applicable for a membrane filtration system.

3.4.3.4 The Effect of Membrane Module Orientation

Figure 3.13 illustrates the variations of the membrane fouling rates for the two module orientations with the air flow rate changing from 0.1 to 1.0 L/min. From the figure, it is obvious that the fouling rate decreased for both orientations with the increase of air flow rate. However, the fouling rate was always the highest for the 90 degree orientation at the 0.3 m location. This was followed by the 90 degree orientation at the 0.6 m location for which at first the fouling rate was also high but then quickly dropped to a relatively low level. The results suggested that the 0 degree orientation was more beneficial compared with the 90 degree orientation due to lower fouling rates. This might be attributed to the fact that only one side of the membrane was scoured at any one time for the 90 degree orientation, during which time the other side could get fouled quickly. For the 0 degree orientation, however, both sides of the module could still be cleaned at the same time despite the fact that the scouring action would be stronger for the part of the membrane modules closer to the active air stone.

Figure 3.13 Variations of TMPs at the two membrane orientations for a 0.65 g/L bentonite solution

From Figure 3.13, it can also be observed that for the 90 degree orientation, the 0.6 m location showed a better membrane performance compared with the 0.3 m location. This phenomenon can be correlated to the development of the bubble plume. Due to the immersion of the membrane module, the mixing inside the BCR was reduced. At 0.3 m the alternate aeration was only able to reach part of the membrane surface whereas moving up to the 0.6 m location, the bubbling plume was more developed and could expand to a larger area, which as a result promoted the mixing and helped control fouling. Beyond 0.7 L/min, the membrane performance at the 0.6 m location for the 90 degree orientation was almost identical to the 0 degree orientation. Due to the general better membrane performance achieved by the 0 degree orientation, the 0 degree orientation was chosen to study the effect of switching interval.

3.4.4 The Effect of Feed Solution

Two types of feeds, namely, bentonite and yeast solution were examined in the current study. Figure 3.14 illustrates the variations of fouling rates against the air flow rate for both the bentonite and yeast solutions with the same mass concentration of 0.65 g/L. The operating conditions with the yeast solutions were identical to those presented in Figure 3.11 for the bentonite solutions, except that the constant flux was reduced from 71.6 L/m²hr to 28.6 L/m²hr. The reduction of permeate flux was necessary as otherwise the foulant deposited too fast with the yeast solution. Further explanation of using lower flux is provided in the following.

From the figure, it could be observed that initially the fouling rate for the yeast solution was much higher compared with the bentonite solution. With the increase of air flow rate, the pattern of the three stages described above, namely, fast reduction, slow mitigation and limiting stages, appeared. When the 0.6 L/min air flow rate was reached, the fouling rates no longer decreased for both the 0.3 and 0.6 m locations, and a minimum fouling rate was achieved. Hence, the optimal air flow rate would be 0.6 L/min for the yeast solution in this case. The experimental results illustrated that the action of hydrodynamics on membrane fouling performance was similar for both feeds, in terms of the emergence of the three-stage fouling

performance. The smaller magnitude of the minimum fouling rate obtained with the yeast solution actually implied that air sparging might be practically more favorable to remove bio-particles rather than inorganic feeds.

Figure 3.14 Fouling rates for 0.65 g/L bentonite and yeast solutions at the 0.3 and 0.6 m locations

Regarding the filtration flux, we originally conducted the membrane filtration tests at 71.6 L/m²hr with the yeast solution. The experimental results are shown in Figure 3.15 (a) and (b). In the two figures the TMP values shot up to 100 kPa for both locations as soon as the filtration was initiated. Thus, the fouling rate was much higher than the bentonite solution shown in Figure 3.11 with the same mass concentration. This could be partially attributed to the cake formation mechanisms which were considerably different between bentonite and yeast solutions. Another

reason might be due to the presence of supernatant in the yeast solution which contained cell debris and extracellular polymer substances (EPS), with EPS already being proven to be the major factor that causes rapid foulant accumulation (Sur and Cui 2005). In addition, the difference in fouling rate could be contributed to the difference of particle numbers in the two feeds. The particle size distributions were measured for both solutions using a Mastersizer 2000. It was determined that the bentonite particles had an average diameter of 5.83 μm , while for the yeast particles the average diameter was about 4.95 μm . Although the diameters were similar, the densities were vastly different for the two solutions. After calculation, it was found out that the particles in the yeast solution were ~ 3.5 times the number of particles in the bentonite solution. Last but not least, the presence of very small size particles in the yeast solution might also have led to potential pore blockage in the membranes. Hence, under the same operating conditions, the membrane was more prone to fouling in the yeast solution due to the existence of supernatant, the much larger number of particles as well as the presence of extremely small size particles. As a result, the membrane fouling behavior could only be monitored if the trans-membrane flux was reduced. After some trial and error, the flux in the yeast solution was finalized to 28.6 $\text{L}/\text{m}^2\text{hr}$.

(a)

(b)

Figure 3.15 Variations of TMPs at (a) 0.3 m and (b) 0.6 m locations for a 0.65 g/L yeast solution (Flux = 71.6 L/m²hr)

3.5 Conclusions

The membrane filtration tests confirmed quantitatively the existence of an optimal air flow rate for the air sparging of a membrane module. The optimal air flow rate would serve as the upper limit for the assessment of the air flow rate associated with the sustainable flux. The experimental results also showed that the meandering flow regime which exhibits higher flow instability is more effective towards fouling alleviation.

Two different air sparger configurations, namely, the central and alternate aeration schemes, were tested. Various factors such as the air flow rate, flow regime, switching interval and the membrane module orientation were investigated. Compared with the central aeration scheme, the alternate aeration produced by the two air stones with 1.0 min switching interval can greatly reduce the aeration demand by applying the air flow rate at the local minimum with the 0 degree membrane module orientation. In addition, the application of inorganic and organic feeds for membrane filtration demonstrated that the particle composition as well as particle size distribution all contribute to the cake formation on the membrane surface.

4. Dynamic Pressure Sensing Technique

4.1 Introduction

In Chapter 3 the effect of air sparging on the membrane filtration performance was investigated using an experimental approach. However, the detailed hydrodynamic characteristics of the laboratory scale BCR remained unknown. The physical understanding of the hydrodynamics in multiphase reactors including the BCR is important for the design and operation, control as well as for the system scale up purposes. Although a number of techniques have been developed and applied for the hydrodynamics measurements, the majority of them are not suitable for pilot or prototype scale reactors due to the sensitive equipment involved.

This chapter focuses on the hydrodynamic diagnostics of the flow conditions inside the BCR using a novel dynamic pressure sensing approach. The BCR with the same dimensions as the one used for the membrane filtration tests is utilized here. The transient dynamic pressure fluctuations are recorded at different air flow rates with special computational algorithms developed to extract the hydrodynamic information from the time-series data. In the following a review of the techniques that are usually employed for hydrodynamics measurements is given before discussing the current experimental setup and methodologies. In addition, various time-series analysis methods are introduced and applied in this study for hydrodynamics quantification.

4.2 Hydrodynamics Measurement Techniques

Basically the techniques for hydrodynamic diagnostics in multiphase reactors can be classified into two categories. The first one is the various types of intrusive probes that are utilized to study the dynamic fluctuations of a signal related to the flow pattern (Vial *et al.* 2001). The second is the imaging and velocimetric techniques (Shaikh and Al-Dahhan 2007). We shall first review the literature that involves the use of intrusive probes, and then follow by the description of the application of non-invasive imaging and velocimetric techniques.

4.2.1 Point Measurement Probes

- Pressure probes

Nishikawa *et al.* (1969) used pressure probes to capture the static pressure pulsations in upward air-water flows. After computing the standard deviation, frequency function as well as the spectral density of the static pressure pulsations, they classified the features of the properties that each flow pattern presented. However, this method was not preferable since the static pressure signals included pressure fluctuations not only at the measured point but also places outside the measurement section.

Matsui (1984) estimated the flow regimes of a nitrogen gas-water mixture in a vertical pipe based on the classification of features of differential pressure fluctuations. Four static pressure signals were collected at both short and long intervals to eliminate the influence of pressure fluctuations occurring outside the measurement section. Statistical properties including the probability density function, cross-correlation, mean and variance were computed to demonstrate the predominant features for each flow pattern.

Drahoš *et al.* (1991) determined the three basic flow patterns, namely, the homogenous, transition and turbulent regimes based on the statistical characteristics of pressure fluctuations in a BCR. Various operating parameters including the superficial gas velocity and liquid height were investigated both axially and radially. In addition, spectral analysis was done to distinguish different sources of the pressure signals.

Letzel *et al.* (1997) used piezoelectric sensors with a tube mounted on its tip to measure the pressure fluctuations in a BCR. The method of chaos analysis was employed to characterize the flow regimes and regime transitions. The Kolmogorov entropy (KE) as a function of the gas velocity indicated a sharp transition from the homogeneous to the churn-turbulent flow regime. The regime transition was less clearly identified when other methods such as the computation of variance and gas holdup were utilized. Hence, they argued that the method of chaos analysis might be a powerful tool to study the influence of pressure on the hydrodynamics in BCRs.

Trnka *et al.* (2000) proposed a novel model to determine an actual operation regime based upon on-line evaluation of amplitude spectra of the pressure fluctuations in a FBR. By assessing the E value, which was defined as square root of the spectral power of the pressure fluctuations in the chosen frequency band, it was possible to explore different states of fluidization inside the FBR. They claimed that this instrumentation technique was more sensitive to the changes in gas flow rate compared to just measuring the static pressure.

Park and Kim (2003) collected differential pressure signals in a BCR and applied the Fast Fourier Transform (FFT) and wavelet packet transform (WPT) techniques to analyse the fluctuation signals. It was found that WPT was able to capture the features of signals that changed over time while FFT could not obtain the time-varying frequencies. By resorting to the WPT technique, the bubbles in bubbly flow were found to have finer scales and frequencies than those in the churn-turbulent flow regime.

Shou and Leu (2005) measured the absolute pressure fluctuations in a FBR both above and below the air distributor. The flow regimes and transition velocities were

characterized by means of multiple analysis including the standard deviation, Power Spectral Density Function (PSDF), the energy of PSDF and wavelet analysis. They proved that the results derived from the normalized energy of PSDF and wavelet analysis were in good agreement with the standard deviation analysis.

Nedeltchev *et al.* (2011) extracted different entropies (KE and maximum information entropy) derived from the gauge pressure fluctuations to identify the boundaries of different flow regimes in three two-phase reactors involving a BCR, FBR and spouted bed reactor. They concluded that, in general, the KE was large in the case of a high degree of turbulence and small for more regular flow behaviour.

- Optical fiber probes

Bakshi *et al.* (1995) studied the local structure and regime transition in a gas-liquid BCR via multi-resolution analysis of the local gas holdup signals using optical fiber probes. They showed that the flow behaviour could be characterized by the energy content of the measured gas holdup signals in the frequency domain and the intermittence of the gas holdup signals, i.e., the multi-fractal nature. Hence, the signals collected were decomposed using wavelet-based techniques to characterize the energy content as well as the fractal nature of the gas holdup signals. They suggested that the wavelet analysis provided a rigorous framework for unifying statistical, spectral and fractal analysis methods.

Farag *et al.* (1997) used optic and capacitance probes simultaneously to investigate the local flow properties in pilot plant scale FBRs. They found that the optical probe tended to underestimate the bubble fraction while the capacitance probe overestimated its value. Moreover, the difference between the optical and capacitance probe measurements increased with temperature.

Vázquez *et al.* (2007) applied optical fibre probes to monitor bubble presence at different radial positions in a FBR. The method of using two probes located at different axial heights to measure the bubble size and velocity was also developed. It was found that the optical fibre probes were more capable of detecting smaller bubbles comparing with pressure sensor measurements.

Heinrich *et al.* (2007) measured the velocity of a particle moving perpendicular to the detecting optical fiber rows in a spouted FBR. The probe was inserted at ten different positions at the circumference of the FBR to measure the porosity and vertical particle velocity components at different radial and axial locations. By evaluating the time difference between the two signals and the distance between the two detecting rows, the velocity could be obtained. The experimental results illustrated that the porosity increased at higher distances from the air distributor and the air flow rate had a strong effect on the porosity and particle velocity.

- Capacitance probes

In 1951, Morris and Ballou reported on the use of a probe to measure the capacitance to quantitatively determine ‘fluidization uniformity’. Since then, the application of capacitance probes became popular and especially widely used in fluidized bed research. However, it was realized that the use of probes involved problems of calibration and interpretation.

Geldart and Kelsey (1972) calibrated a capacitance probe in a two-dimensional bed by making simultaneously with the probe trace, cine film records of the bubbles passing the probe head. The calibration and standardisation proposed for 2D beds corresponded well with point bubble frequencies and results derived from the capacitance probes. For 3D systems, the performance was satisfactory only for high gas flow rates and when the bubble sizes were larger than 7 cm. Hence, new calibration criteria should be developed separately for 3D beds.

van Lare *et al.* (1997) used a capacitance probe consisting two measuring tips with a distance of 10 mm to study the bubble behaviour in a FBR with four different size quartz sand powders. A statistical method was applied to determine the bubbling rising and contact times. They found that the stable bubble height seemed to increase with the average powder size.

Wiesendorf and Werther (2000) measured the local solid volume concentrations and velocities in industrial FBRs with rough conditions using an improved capacitance probe measurement technique. They proposed a new empirical calibration relationship which was applicable for both liquid-solid and gas-solid systems. They

demonstrated its capability for flow structure characterization in FBRs under harsh conditions.

- Acoustic probes

Glassgow *et al.* (1992) conducted experiments to study the interfacial phenomena using the macrovideography and acoustic signals recorded at both the sparger and free surfaces. Time-series data collected from the microphone were processed to attain the mean and root-mean-square values. Fourier transforms were also computed to identify dominant signal energies. A sequence of experiments was also conducted to correlate interfacial phenomena to the frequencies of the noise produced.

Höller *et al.* (2003) carried out a series of experiments in a three-phase BCR staged with structured fibrous catalysts. They used a coupled high speed camera system and passive acoustic measurements to study the flow behaviour inside the reactor. With the application of the acoustic probe technique, the building frequency of bubble swarm formation was found to be a linear function of the superficial gas velocity.

Al-Masry *et al.* (2005) utilized acoustic hydrophones to determine the bubble size distribution in a BCR. Statistical tools and auto-correlation functions were exploited to quantify the bubble characteristics. They recommended acoustic hydrophones for use as a strong potential sensor for on-line control of BCR at moderate conditions in which classical sensors were not applicable due to foaming fouls.

In addition to those mentioned above, other types of probes were also applied by various researchers to study the hydrodynamics of multiphase reactors. Latifi *et al.* (1989) used micro-electrodes to analyze the liquid-solid mass transfer fluctuations at the wall of a packed bed reactor to identify the flow regimes. The rate of fluctuation of the mass transfer coefficient as a function of the liquid flow rate was proven to be able to distinguish the transition between the inertial flow regime and turbulent flow regime.

Thimmapuram *et al.* (1992) employed a heat transfer probe to diagnosis the different hydrodynamic regimes of a BCR. They concluded that the temperature-fluctuation history data could be used to discriminate the flow regime. A statistical function such as the variance and the average frequency of the data could be used to delineate the flow regimes.

Zhang *et al.* (1997) proposed a new set of criteria to determine the flow regime transition for two-phase flow and three-phase FBRs using conductivity probes. It was shown that the churn flow, bridging flow and annular flow occurring at high gas velocities in three-phase FBRs also existed in gas-liquid two-phase flows. Besides, complete flow regime maps for concurrent three-phase FBRs were obtained and empirical correlations were developed to predict the threshold of the flow regime in three-phase systems.

4.2.2 Imaging and Velocimetric Techniques

Due to the intrusive nature, the presence of probes could interfere with the flow field and result in an inaccurate measurement of the flow quantities (Gan *et al.* 2011). As a consequence, the application of non-intrusive techniques for hydrodynamics measurements is gaining popularity. With the development of measuring techniques, various imaging and velocimetric techniques have been used to study the flow dynamics in multiphase reactors.

- Particle Image Velocimetry (PIV)

Chen *et al.* (1994) studied the macroscopic flow structure of three-dimensional BCR and FBR systems under various operating conditions using PIV. They successfully identified three flow regimes and further proposed a conceptual four-region structure for the vertical-spiral flow regime. In addition, similarities between 2D and 3D flows were pointed out.

Lin *et al.* (1996) conducted experiments to quantify the instantaneous flow behaviour and regime transitions in a 2D BCR based on flow visualization and PIV. They divided the flow regimes into dispersed and coalesced bubble regimes characterized by two flow conditions (4- and 3-region flow) with coherent flow structures. In addition, numerical simulations of dispersed bubbly flow were performed. The predictions were found to lie in a good agreement with the experimental results obtained by PIV and from flow visualization.

Laverman *et al.* (2007) studied the hydrodynamics of a free bubbling, gas-solid FBR experimentally using coupled PIV and Digital Image Analysis (DIA). The emulsion phase circulation patterns, bubble holdup, bubble size and velocity distributions were determined simultaneously. The combination of these two techniques allowed correcting for the influence of particle raining through the roof of the bubbles on the time-averaged emulsion phase velocity profiles.

Julián *et al.* (2012) also coupled the PIV and DIA techniques to analyse the bubble and particle fluidization properties in a two-zone FBR. The PIV analysis allowed the study of channelling phenomena which appeared to be influential at low relative gas velocities. Besides, it was able to provide useful information about the solid circulation pattern in the transition section.

- Laser Doppler Anemometry (LDA)

Vial *et al.* (2001) measured the time-averaged liquid mean and root-mean-square velocity profiles using LDA in a BCR equipped with three different air spargers. A simple model based on momentum balance equations was used to evaluate the Reynolds shear stress. In addition, a detailed elaboration on the local hydrodynamics of the liquid phase was obtained.

Olmos *et al.* (2003) studied the non-linear dynamics and flow regime transitions in BCRs using LDV by considering the system as a 2D reactor at a higher void fraction than in 3D systems. Various signal processing methods including frequency analysis, fractal analysis and deterministic chaos analysis were applied to the LDA measurements. Based on these analysing methods, two transition regimes were highlighted with detailed information on the respective physical mechanisms.

Bhole *et al.* (2006) conducted measurements on the flow pattern in a BCR using LDA in a forward scatter mode. The hydrodynamics induced by two different gas spargers, namely, the porous plate and perforated plate, were investigated. With the porous plate sparger, smaller bubbles and higher gas holdup were found comparing with those obtained with the perforated plate sparger.

- Computer Automated Radioactive Particle Tracking (CARPT)

Cassanello *et al.* (2001) carried out experiments to study the liquid displacement in BCRs via CARPT. The liquid-phase dynamics were examined using Lagrangian and qualitative dynamics tools in BCRs of different sizes at different pressures. They concluded that the liquid motion was chaotic and the information loss rate was strongly affected by pressure but weakly sensitive to vessel diameter.

Al-Dahhan (2002) studied the flow dynamics and visualized the flow field in various multiphase reactors using both CARPT and γ -ray computational tomography (CT). The data were obtained in a non-invasive manner to validate the results from CFD. They believed that the coupled CT and CARPT technique could be used to advance the design, scale-up and modelling of the system.

Nedeltshev *et al.* (2003) adopted two new quantities, i.e., KE and the quality of mixedness (QM) derived from CARPT data to identify the flow regimes in a BCR. They pointed out that the KE was proportional to both the bubble frequency and bubble impact. In addition, the QM in the lower and upper parts of the column was found to be the same at the transition velocity.

- Electrical Resistance Tomography (ERT)

Dong *et al.* (2003) analyzed the measured data of an ERT system to identify the flow regimes and void fraction. A novel method was provided to identify the flow regimes online based on the feature extraction of the ERT boundary voltage signals. A method was also developed to estimate the void fraction based on the physical analogue. Later, the same team conducted two-phase flow measurements with the continuous phase of a conductive liquid using the dual-plane ERT combined with the eigenvalue correlation measurement technique. By applying this method, they

claimed that the transient flow rate of the disperse phase going through the measured cross-section could be derived (Dong *et al.* 2005).

Vijayan *et al.* (2007) studied the effect of sparger geometry in a BCR using ERT measurements. The effect of column pressure and superficial air velocity for different air sparger geometries and for different flow patterns was explored. It was found that the gas holdup derived from the ERT measurements fitted well with the results obtained by pressure sensors. The spectral analysis of the ERT data showed that the void fraction wave characteristics were an important parameter for flow regime discrimination.

Kim *et al.* (2013) utilized ERT to visualize multiphase flows. With respect to the non-linearity and ill-posedness of the inverse problem, an adaptive multi-threshold method was applied to improve the reconstruction accuracy and decrease the ill-posedness. In addition, numerical simulations were carried out to evaluate the reconstruction performance of the proposed method.

- Electrical Capacitance Tomography (ECT)

Bennett *et al.* (1999) employed an ECT sensor to investigate the potential of this technique for bubble column visualization and control. Various methods involving the computation of variances and gas holdup were proposed to distinguish between homogeneous bubble flow and churn flow in a BCR. Factors such as the air sparger configuration distance between the sparger and sensor as well as the frother concentration were considered to test the viability of the method.

Warsito and Fan (2001) conducted measurements of the real-time flow structures in gas-liquid and gas-liquid-solid systems using the ECT technique. They developed an image reconstruction technique which was a key factor in producing an accurate image of the complex flow systems. It was found that this approach could be applied for low gas holdup systems.

Warsito and Fan (2005) also developed a ‘dynamic’ 3D image reconstruction technique for ECT imaging based on a neural network multi-criteria optimization. The technique enabled a real time, 3D imaging of a moving object to be conducted.

Furthermore, it also allowed a total investigation of a whole volume within the sensor domain to be implemented.

- X-ray detectors

Rowe and Partridge (1997) used X-ray cine photography to observe bubble characteristics in gas-fluidized beds. It was found that stable bubbles had an essentially spherical form with a particle-wake region occupying the lower part of the sphere. Also, both the size of the particle and the bubble velocity depended upon the size and shape of the fluidized particles.

Yates *et al.* (2002) summarized some examples of utilization of the X-ray technique to study the flow characteristics inside multiphase systems. The application of this technique in the analysis of the hydrodynamic features of bubbling and circulating FBRs was reviewed. Also, measurements of jet penetration into bubbling beds and a study of bubble slurry columns were also studied. Overall, they proposed that the technique was useful for real-time observation and analysis of the many hydrodynamic features of FBRs.

- γ -ray Computational Tomography (CT)

Kumar *et al.* (1995) measured the void fraction and its distribution in two-phase systems such as FBRs and BCRs. The automatic scanner was proven to be able to image flows in tested sections. Among various reconstruction algorithms, the Estimation-Maximization (E-M) algorithm demonstrated its superiority for image reconstruction. Overall, this approach was believed to be able to provide both local and global information over the entire flow domain.

Shaikh and Al-Dahhan (2005) carried out experiments to characterize the flow regimes using single source γ -ray CT in a high pressure BCR. The regime transition velocities obtained from CT measurements were analysed to demarcate the hydrodynamic flow regimes. It was found that with an increase in operating pressure, the regime transition decayed over a range of superficial air velocity.

Roy *et al.* (2005) implemented this technique to study the liquid flow distribution in a column packed with a corrugated structure using counter current air-water flow.

The liquid holdup and its radial distribution at various gas and liquid velocities were investigated. They found that the liquid distribution was fairly uniform and better at the bottom of the bed compared with the upper section.

4.2.3 Constraints of Different Measuring Techniques

Although various types of measuring techniques are available for flow dynamics characterization in multiphase reactors as discussed in Sections 4.2.1. and 4.2.2., most are not feasible for industrial pilot or prototype scale applications due to different constraints. The capacitance probes can be used to determine solid volume concentration and velocities (Wiesendorf and Werther 2000). However, it must be calibrated for every fluid-solid system and operating conditions (particularly temperature and pressure) and thus is very time-consuming. Probes based on sensing local electrical properties are difficult to use when the reactor operates over a range of temperatures and particle compositions. X-ray photography, PIV as well as DIA are expensive, and limited to regimes with a low concentration of bubbles. When expanded to large scale, not only the temperature and the chemically aggressive environment have to be considered, but also the mechanical stresses due to the vigorous motion of the solid particles and an erosive attack caused by sandblasting effects make the application of the majority of the techniques difficult (Werther 1999).

Among all the approaches, the pressure sensor offers the simplest and least expensive detection device. Despite being an intrusive method, the pressure sensing measurements can be conducted inside BCRs, FBRs and even for industrial chambers of very large size (Wiesendorf and Werther 2000). This approach thus offers an attractive alternative to study the hydrodynamics in multiphase reactors, particularly those with harsh reactor environments.

Despite having many advantages, the use of dynamic pressure sensing for hydrodynamic diagnostics in MBRs has not been well studied and reported in the past. In the literature, rather than dynamic pressure, absolute or static pressure fluctuation signals were mostly collected and analyzed. The primary application for signal analysis was to discriminate different flow regimes in multiphase reactors. However, the dynamic pressure fluctuations induced by the passing of individual bubbles and macro-scale circulation of the bubble plume are worth studying since they are closely related to the flow instability, which is essential for foulant removal in MBRs. Furthermore, not only the threshold of each flow regime affects the reaction process, the dynamics of the flow behavior for each flow regime for different operating conditions was also important in terms of air sparger design and system optimization. Hence, the development of diagnostic parameters derived from dynamic pressure signals is necessary for the characterization of the flow hydrodynamics of MBRs. In this study, a novel method was developed to measure the transient dynamic pressure fluctuations at various axial locations inside a BCR with and without the submersion of a membrane module. Special algorithms were developed to unlock the complex hydrodynamic pattern of the laboratory scale BCR. The detailed hydrodynamics were quantified under different air flow rates to correlate the flow behavior with the membrane filtration performance discussed in Chapter 3.

4.3 Experiments

4.3.1 Apparatus

Figure 4.1 shows the schematic diagram of the experimental setup for the dynamic pressure sensing measurements. The setup consisted of a BCR, a submersible pressure sensor and a data logging system. The BCR had the same geometry as the one used in the membrane filtration tests. Three kinds of feed solutions, namely,

clean tap water, a 0.65 g/L bentonite suspension and a 0.65 g/L yeast solution were used, and the solution was filled to one meter height. Nine holes with a diameter of 27 mm and at 0.1 m interval were drilled into the wall of the BCR. The distance between the center of the lowest hole to the bottom of the reactor was 0.1 m. Two locations, 0.3 and 0.6 m to the bottom of the reactor, were measured to study the local hydrodynamic characteristics and further correlate with the membrane filtration performance. The air flow rate changed from 0.1 to 2.0 L/min at an interval of 0.1 L/min with the instantaneous dynamic pressure signals recorded. The submersible pressure sensor (a Kistler Instrument's Product) was connected to a power supply box (Kistler Instrument Pte Ltd), which served the function of signal activation. The power supply box was further linked to a data acquisition system (DAQ) (a National Instrument's product). The data logging system was able to export data at a speed of 100 kS/s and the resolution per channel was 16 bits. The PFSs collected were logged and stored in a computer system for post-processing. The experiments were carried out with and without the submersion of a hollow fiber membrane module to compare the hydrodynamic characteristics.

4.3.2 Experimental Procedures

During the experiments the submersible pressure sensor was placed in the BCR through one of the holes. The distance from the tip of the sensor to the wall boundary of the reactor was maintained at 4 cm. This finite measuring distance was set to eliminate the near wall effect as the focus of this study was the free stream flow rather than the flow behaviour near the wall. When the sensor was properly positioned, the air compressor was switched on to provide air continuously into the reactor. Once steady state was reached, the fluctuating voltage-time signals which corresponded to the fluctuating pressure-time signals were collected at a sampling rate of 0.01s for three minutes. In this regard a total of 18000 data points were recorded. Subsequently, the data collection was carried out for a short period of 30 seconds with a 0.001 s^{-1} sampling rate. Figure 4.2 shows a time series of PFSs

collected at the 0.3 and 0.6 m locations at the air flow rates of 0.1, 0.5, 1.0 and 2.0 L/min and a sampling rate of 0.01 s respectively. It can be seen that with the increase of air flow rate, the magnitude of dynamic pressure also increased for both locations. However, when the dynamic pressure signals were compared for the two axial positions under the same air flow rate, the inherent propensities of the flow could not be revealed directly from the time-series data. Thus post-processing was needed to extract the hydrodynamic information from the dynamic pressure signals. The detailed analysis methods are discussed in Section 4.4.2.

4.3.3 Submersible Pressure Sensor

The diameter of the submersible pressure sensor was 26.9 mm and the length was around 127 mm. The measuring pressure range was from 0 to 0.2 bars, which corresponded to the voltage output ranging from 0 to 10 V. Before initiating the experiment, different models of submersible pressure sensors were tried out to test their response to the transient dynamic pressure changes. It was found that most of the sensors commercially available were not feasible due to a high inertia and a lack of sensitivity. After much effort, it was finally identified that the sensor produced by Kistler Instrument was suitable for handling this test. It has a fast response time of 5×10^{-6} s and with calibrated data-logging capability. Preliminary tests showed that it was capable of reacting to the transient dynamic pressure changes caused by the turbulent flow conditions.

4.3.4 Noise Reduction

Due to the small amplitude of the dynamic pressure signals, they can be easily contaminated by the background noise. In order to minimize the influence of outside noise, the transient dynamic pressure signals were collected simultaneously

in another BCR with identical dimensions as the one used for the measurements but with no aeration. The signals acquired were subtracted from those obtained in the original BCR so that only the desired local dynamic pressure related to the turbulent fluid motion could be retrieved. In addition to the simple subtraction scheme, low-pass filtering of 100 Hz was conducted to remove high frequency redundant signals mainly composed of noise. Prior to data analysis, the filtered signals were subtracted from their mean to account for the fluctuation part only.

Figure 4.1 Experimental setup for the dynamic pressure sensing measurements

(a)

(e)

(b)

(f)

(c)

(g)

(d)

(h)

Figure 4.2 Dynamic pressure fluctuations at the 0.6 m (left) and 0.3 m (right) locations at the air flow rates of 0.1, 0.5 1.0 and 2.0 L/min at a sampling rate of 0.01s

4.4 Results and Discussion

4.4.1 Measurements and Calculations

Gas volume fraction is also normally called gas holdup, which is a dimensionless key parameter for design purposes that characterizes the transport phenomena of BCRs. It is defined as the volume fraction that is occupied by gas bubbles. Gas volume fraction plays an important role in the design and analysis of the BCR, hence, it is employed to examine the hydrodynamic characteristics of BCRs (Kantarci *et al.* 2005).

The mean gas volume fraction, α , for a BCR can be determined by measuring the increase in the liquid level when gas is introduced to the reactor, i.e.:

$$\alpha = (1 + \frac{H_0}{\Delta H})^{-1} \quad (4.1)$$

where H_0 is the initial liquid level, which is equal to 1 m in this case, and ΔH is the increase of the water level when subjected to air sparging in the reactor (Gan *et al.* 2011).

4.4.2 Time-Series Data Analysis

4.4.2.1 Statistical Analysis

For discrete pressure signals the mean (μ) which represents the average value of a series of signals computed as follows, determines the centre of the data distribution (Smith 1997):

$$\mu = \frac{\sum_{i=1}^N x_i}{N} \quad (4.2)$$

where N is the number of signals and x_i is the magnitude of each pressure signal. Besides μ , the standard deviation, σ_{STD} , is also commonly adopted to indicate the degree of dispersion of the signals around the mean, which is calculated as follow (Smith 1997):

$$\sigma_{STD} = \sqrt{\frac{1}{N} \sum_{i=1}^N (x_i - \mu)^2} \quad (4.3)$$

In the following the PFSs were first collected at a sampling rate of 0.01s and the statistical analysis was carried out to quantify the hydrodynamics at the 0.3 and 0.6 m locations. Two factors, the effect of flow regime and the nature of feed solutions, were considered with the detailed results discussed.

Figure 4.3 (a) and (b) show the mean pressure changes versus the air flow rate at the 0.3 and 0.6 m in the clean tap water. It can be seen that the mean pressure value increased slightly with the air flow rate for both locations. Earlier studies (Deckwer *et al.* 1980; Schumpe and Grund 1986; Saxena *et al.* 1990; Daly *et al.* 1992; Pino *et al.* 1992; Hyndman *et al.* 1997; Krishna *et al.* 1997; Li and Prakash 2000; Prakash *et al.* 2001) reported that in BCRs, the mean gas volume fraction generally increases with the air flow rate, regardless of the reactor geometries. Since the mean gas volume fraction correlates with the liquid level as described in Section 4.4.1, the increase of the gas fraction thus causes the increase of the mean water level, which leads to the increase in the mean hydrostatic pressure.

(a)

(b)

Figure 4.3 Average pressure at the (a) 0.3 and (b) 0.6 m locations for different air flow rates

Figure 4.4 is the computational results for the standard deviation of PFSs under different air flow rates at the 0.3 and 0.6 m locations in the clean tap water. For both

locations, when the air flow rate was smaller than 1.1 L/min, the standard deviation increased with the air flow rate. Between 1.1 to 1.4 L/min, the standard deviation flattened out and a maximum point appeared at an air flow rate of 1.4 L/min. Beyond 1.4 L/min, the standard deviation decreased with increasing air flow rate and with a faster decrease at the 0.6 m location. The observed trend might be attributed to the two-phase turbulence fluctuations reaching a saturation maximum inside the BCR.

Figure 4.4 Standard deviations of the dynamic pressure at 0.3 and 0.6 m locations under different air flow rates

From Figure 4.4, the maximum saturation point for the 0.3 and 0.6 m locations was similar regardless of the different flow regimes in which the sensor was immersed. Thus, the attainment of turbulent fluctuation saturation was independent of the flow regime, and was controlled primarily by the air flow rate when water was used as the feed solution. It should be noted that the determination of the maximum saturation point is of great importance to the hydrodynamics optimization to enhance the mixing condition inside the BCR as the turbulent flow fluctuations are maximized under this condition.

Generally, the level of turbulence in the BCR would increase with higher mean gas volume fractions along with higher air flow rate. However, Lance and Bataille (1991) proposed that there are two distinct regimes in bubbly flow. The first one is the “pseudo turbulence” regime corresponding to lower values of the mean gas volume fraction, within which the hydrodynamic interactions between bubbles are negligible. Within this regime the bubble formation occurs in “bubble to bubble” mode whereby the motion of the bubbles is relatively independent of each other. The second one is the “wake turbulence” regime corresponding to higher values of the mean gas volume fraction, within which the bubbles mutually interact strongly due to their closer spacing. In Lance and Bataille’s paper they proposed a critical void fraction of the order of 1% as the demarcation between the two regimes. Magaud *et al.* (2001) studied the influence of the local void fraction on the turbulence intensity of the liquid flow in a vertical rectangular channel. They concluded that for lower void fractions ($\leq 5\%$), the measured turbulence intensity increased with the void fraction but for higher values ($\geq 5\%$), the turbulence intensity became nearly constant. In this case the critical void fraction was achieved at around 1.4 L/min with the turbulence fluctuations reaching the saturation maximum. Hence, further increase in the air flow rate would not induce additional turbulence fluctuations inside the BCR.

In addition to clean tap water a 0.65 g/L bentonite suspension and a 0.65 g/L yeast solution were also utilized to study the effect of the feed properties. Figure 4.5 shows the mean pressure under various air flow rates at the (a) 0.3 and (b) 0.6 m locations for the three solutions. It can be seen from Figure 4.5 (a) that the mean

value for the bentonite solution was close to that for the clean water, while it was higher for the yeast solution when the air flow rate was larger than 0.7 L/min. Furthermore, a rapid growth of the mean value for the yeast solution was observed at an air flow rate of 0.7 L/min. The overall higher mean values for the yeast solution might be due to higher gas volume fractions since the gas holdup was higher with the addition of organic particles compared with pure liquids (öztürk *et al.* 1987). Moreover, the presence of supernatant in the water also significantly increased the gas holdup values (Kantarci *et al.* 2005). Since the supernatant was a signature component of the yeast solution, it led to the higher gas holdup, which ultimately increased the mean pressure magnitude. The similar behavior of the water and bentonite solution could be attributed to their similar dynamic viscosities which was already mentioned in Section 3.3.1.3.

(a)

(b)

Figure 4.5 Average dynamic pressure at the (a) 0.3 and (b) 0.6 m locations under different air flow rates for clean tap water, 0.65 g/L bentonite and 0.65 g/L yeast solutions

In 4.5 (b), the mean values were the highest for the yeast solution. It indicated that the presence of the solid particles did have a considerable effect on the mean gas volume fraction values. Despite the difference in flow regimes developing from 0.3 to 0.6 m, the supernatant and organic particles in the yeast solution still played an important role in increasing the mean gas volume fraction. Figure 4.5 shows the standard derivations of the dynamic pressure fluctuations under various air flow rates at the (a) 0.3 and (b) 0.6 m locations for the three feed solutions accordingly. In Figure 4.5 (a), it can be seen that the developing trends were similar for all. Initially, the standard deviation values increased with increasing air flow rate. Subsequently, the three curves flattened out and a maximum saturation point was achieved. After that, the standard deviations decreased at high air flow rates. The maximum point for all three solutions again indicated the attainment of a maximum saturation for the turbulence fluctuations in the BCR at the optimal air flow rate. Further increase of the air flow rate would not promote additional turbulence in the BCR. In terms of magnitudes, the water and yeast solutions were roughly the same while for the bentonite solution, the overall standard deviation values were smaller and the air flow rate at the saturation point was also smaller. The difference in the standard deviation trend for the bentonite solution relative to the clean water indicated changes in the flow hydrodynamics due to the presence of inorganic particles. Particularly, the inorganic loading acted to dampen the turbulence level, which might reduce the mass and heat transfer efficiency in BCRs, which had been suggested by Gan *et al.* (2011). The similarity in the developing trend between water and the yeast solution implied that the particle size and density played significant roles and that the smaller yeast size and density did not have as much of an effect as the larger bentonite particles. Generally, it could be concluded that the presence of the solid phase and the solid particle properties were crucial for the turbulent flow fluctuations in the BCR. A small change in the particle properties could lead to a drastically different hydrodynamics behavior inside the BCR.

(a)

(b)

Figure 4.6 Standard deviation of the dynamic pressure at the (a) 0.3 and (b) 0.6 m locations under different air flow rates for clean tap water, 0.65g/L bentonite and 0.65 g/L yeast solutions

In Figure 4.6 (b), all three solutions again displayed a similar developing trend in terms of the magnitudes. Firstly, the standard deviations increased with the air flow rate. After attaining a maximum at around 1.3 L/min, the standard deviation started to decrease. Since the bubble plume was fully developed and stabilized at the 0.6 m, it indicated that in this flow regime, the influence of solid particle properties was not that dominant when low mass concentrations were applied.

Generally, various factors can affect the mixing process and complicate the hydrodynamics of the BCR. By comparing the average and standard deviation of the PFSs, the optimal air flow rate that leads to the maximum standard deviation at the saturated turbulence level can be determined. Hence, the turbulent flow conditions in the reactor can be optimized to improve the mass and heat transfer characteristics. This quantitative knowledge can also help determine the optimal air flow rate for air sparging to prevent excessive use of energy for aeration.

4.4.2.2 Spectral Analysis

The analysis of pressure fluctuations in the frequency domain is commonly used to quantify the fluidization regime and to understand the physical phenomena taking place in FBRs (Albasi *et al.* 2004; de Martín *et al.* 2011). Discrete signals can be converted from the time domain to the frequency domain using the Discrete Fourier Transform (DFT). The DFT, F_x , of a pressure-time series measured at a position x , P_x , is defined as:

$$F_x(f) = \frac{1}{N} \sum_{n=0}^{N-1} P_x(t_n) e^{-j2\pi f t_n / N} \quad (4.4)$$

In statistical signal processing, the spectrum of a time-series or signal is a positive real function of a frequency variable associated with a stationary stochastic process, or a deterministic function of time, which has dimensions of power per hertz (Hz), or energy per hertz. Intuitively, the spectrum decomposes the content of a stochastic process into different frequencies present in that process, and helps identify periodicities. The specific terms used is called the power spectral density function (PSDF).

The PSDF is calculated as follow:

$$\varphi_{xx}(f) = \frac{1}{f_s} F_x(f) F_x^*(f) \quad (4.5)$$

where F_x^* is the complex conjugate of F_x and f_s is the sample frequency.

The PSDF can be used to characterize the flow pattern inside the multiphase reactor and is also capable of discriminating between the respective flow regimes. However, the threshold for each flow regime cannot be determined exactly using the PSDF alone (Vial *et al.* 2001). Compared with the computation of the mean and standard deviation which are easier to determine, the quantification of the frequency spectrum is not that straightforward due to more stringent data acquisition requirements (Schouten and van den Bleek 1998).

It was found previously that from $z=50$ to 300 mm, the bubbles meandered within the column and from $z=300$ to 450 mm, the flow pattern was transitional. Beyond this region, the flow became relatively uniform (Gan *et al.* 2011). Hence, three locations from each respective regime, namely, 0.3, 0.5 and 0.7 m, were chosen with the power spectrum plotted accordingly. For the spectral, Hurst and chaos analysis to be discussed in the following, the dynamic pressure signals were collected at a 0.001s^{-1} sampling rate to prevent loss of information which might occur when the time interval between the two data points is long.

Figure 4.7 shows the power spectrum at the 0.3 m location without (left) and with (right) the submersion of the membrane module for the four air flow rates of 0.5, 1.0, 1.5 and 2.0 L/min, respectively. It could be seen that when no membrane was immersed as shown in Figure 4.7 (a-d), a dominant frequency of 0.2 Hz existed for the low frequency segment (0-10 Hz) and the amplitude generally increased with the increase of air flow rate. The existence of this local peak frequency could be due to the macro-scale circulation of the liquid flow induced by the bubble movement near the air sparger. Besides, a dominant frequency with slightly different peak values was also found in the high frequency segment (10-100 Hz), which dropped from 47 to 28 Hz with increasing air flow rate. Meanwhile, the amplitude of the peak frequency increased with the air flow rate. The appearance of this peak

frequency in the high frequency segment could be attributed to the meandering flow of the bubble plume which led to a large number of bubbles passing the sensor within a short time scale. The fairly high peak value indicated that the strength of the meandering flow was enhanced with increasing air flow rate. It also confirmed that the meandering movement of the bubble plume was the dominant flow pattern at the 0.3 m location. It was visualized that when the air flow rate was relatively small, the bubble plume was prone to moving upward instead of wandering back and forth due to the buoyancy effect. When the air flow rate was increased to a relatively high value, the meandering flow became distinct and dominated near the air sparger. Besides, the wide spectrum width resulted from the chaotic nature of the system which was also observed in Figure 4.2 as indicated by highly irregular pressure fluctuations over time.

With the presence of the membrane module (Figure 4.7 (e-h)), the dominant frequency of 0.2 Hz still existed and the amplitude also increased with the air flow rate. However, in comparison with Figure 4.7 (a-d) where no membrane module was submerged, the amplitude was significantly smaller for the four air flow rates. This indicated that the macro-scale liquid circulation near the air sparger were dampened by the presence of the membrane module. The total energy carried by the bubble plume was thus lowered. In terms of individual bubble movement, the dominant frequency that appeared in the high frequency segment was around the same value and with similar magnitudes for different air flow rates. This illustrated that individual bubble movement was not influenced significantly by the membrane module in comparison to the no membrane case.

Figure 4.8 shows the power spectrum at the 0.5 m location without (left) and with (right) the submersion of the membrane module for the four air flow rates, respectively. It can be seen that when no membrane module was submerged (Figure 4.8 (a-d)), the dominant frequency of 0.2 Hz located at the low frequency segment was less prominent when the location was shifted from 0.3 to 0.5 m. Besides, the dominant frequency of around 30 Hz located in the high frequency segment for the 0.3 m location was also found at the 0.5 m location. However, the amplitude was much smaller in comparison to those obtained at 0.3 m, especially for large air flow

rates (>1.5 L/min). It indicated that the meandering flow was no longer dominant at 0.5 m and the bubble plumes were more developed in comparison to the bottom portion of 0.3 m. Meanwhile, another dominant frequency at 85 Hz occurred when the air flow rate was larger than 1.0 L/min. The amplitude of the peak frequency increased with an increase in air flow rate, which indicated the increase in the energy level with increasing air flow rate. The emergence of two peak frequencies implied that 0.5 m was located in the transition regime (Vial *et al.* 2001). The appearance of the fairly high peak frequency of 85 Hz could be attributable to the highly unsteady flow condition taking place in this region.

With the submersion of the membrane module as shown in Figure 4.8 (e-h), the 0.2 Hz dominant frequency re-appeared. Since the membrane was immersed at the axial central position, the central space was occupied and thus the bubble plume had to develop from the sides of the column. In other words, with the submersion of the membrane module, the bubble plume was forced to swing back and forth rather than just moving upward due to buoyancy when the air flow rate was small. Similar to Figure 4.8 (a-d), two dominant frequencies of 30 and 85 Hz also occurred in the higher frequency segments when the air flow rate exceeded 1.0 L/min. This actually indicated that the presence of the membrane module did not change the flow dynamics significantly.

Figure 4.9 shows the power spectrum at the 0.7 m location without (left) and with (right) the submersion of the membrane module for the four air flow rates respectively. Compared with the other two locations, a wide frequency band was found at the air flow rate of 0.5 L/min. This indicated that at the 0.7 m location, the bubble plume was fully developed and the bubbles were dispersed across the whole cross-section of the BCR. Without the membrane module (see Figure 4.9 (a-d)), the dominant frequencies of 30 and 85 Hz were observed at the upper portion of the column as well. In terms of amplitudes, the 30 Hz dominant frequency had a smaller value compared with the 0.3 and 0.5 m locations while the 85 Hz peak frequency showed a substantially higher value than the other two locations. It implied that near the water surface, the flow dynamics was more complicated due to the influences of various sources such as the water level oscillations and variations

in bubble sizes. When the membrane module was submerged, the amplitude of the 85 Hz peak frequency was lowered. The smaller amplitude of the lower dominant frequency suggested that the bubble plume expanded across the whole column cross-section at the 0.7 m location. In addition, the energy carried by the bubbles was substantially less with the submersion of the membrane module in terms of smaller amplitudes over the entire frequency band. Hence, the existence of the membrane module was shown to suppress the bubble movement which was not beneficial for fouling control at this region.

Overall, the PSDF plot at the three locations with and without the submersion of the membrane module revealed the extremely intricate flow hydrodynamics of the BCR. The spectral analysis should not be used as the single signal processing tool for hydrodynamic diagnostics due to limited information it can provide. In the following, the Hurst and chaos analyses are developed to explore the flow dynamics in more details.

Figure 4.7 Power spectrum at the 0.3 m location without (left) and with (right) the submersion of the membrane module for the four air flow rates

Figure 4.8 Power spectrum at the 0.5 m location without (left) and with (right) the submersion of the membrane module for the four air flow rates

Figure 4.9 Power spectrum at the 0.7 m location without (left) and with (right) the submersion of the membrane module for the four air flow rates

4.4.2.3 Hurst Analysis

The Hurst exponent is a useful parameter to characterize the long-term correlation among fluctuations of a time-series signals (Shaikh and Al-Dahhan 2007; Luo *et al.* 2012). For a discrete time series $x(t)$ the Hurst exponent, H , is calculated from the Hurst rescaled range (R/S) of a time series. The steps involved are summarized as follows for direct reference (Shaikh and Al-Dahhan 2007):

$$\text{Mean} \quad \bar{x}_k = \frac{1}{\tau} \int_{t_k}^{t_k+\tau} x(t) dt \quad (4.6)$$

$$\text{Standard deviation} \quad S_k = \left\{ \frac{1}{\tau} \int_{t_k}^{t_k+\tau} [x(t) - \bar{x}_k]^2 dt \right\}^{0.5} \quad (4.7)$$

$$\text{Accumulated departure} \quad X_k(t) = \frac{1}{\tau} \int_{t_k}^{t_k+\tau} [x(t) - \bar{x}_k] du \quad (4.8)$$

$$\text{Range} \quad R_k = \max X_k(t) - \min X_k(t) \quad (4.9)$$

$$\text{Rescaled range} \quad (R/S)_k = \frac{R_k}{S_k} \quad (4.10)$$

The average of (R/S) can then be obtained for the N sub-periods as:

$$(R/S)_T = \frac{1}{N} \sum_{k=1}^N (R/S)_k \quad (4.11)$$

The evolution of $(\frac{R}{S})_T$ with different sub-period length T , which is normally equals to the power of a set integer value, can then be tracked as:

$$(R/S)_T = c \cdot T^H \quad (4.12)$$

The Hurst exponent, which is dimensionless, can then be computed as:

$$H = \frac{d[\ln(R/S)_T]}{d[\ln T]} \quad (4.13)$$

The Hurst exponent is equal to 0.5 for random signals which indicates that no correlation exists between past and future increments. An H value larger than 0.5 implies the persistent behavior of the time-series, i.e., an increase in the time series is likely to be followed by another increase and vice versa. On the other hand, an exponent less than 0.5 indicates anti-persistence in the data, that is, the trend will likely reverse itself (Briens *et al.* 1997).

Figure 4.10 shows the development of the Hurst exponent against the air flow rate at the 0.3 and 0.6 m locations in the clean tap water. It can be seen that all the H values were less than 0.5, which signified the negative correlation of the dynamic pressure signals. The small value of H also suggested that the flow inside the BCR was highly irregular and unsteady. Rather than long-term correlated, the PFSs were short-term correlated, which also confirmed the chaotic nature of the system. In addition, a similar trend in terms of the development of H was evidenced at both locations. When the air flow rate was smaller than 0.8 L/min, H increased substantially with an increase in the air flow rate. From 0.8 to 1.2 L/min, the Hurst exponent still increased but at a slower pace compared with the small aeration range. For large air flow rates (> 1.2 L/min), the magnitude of H became nearly the same at the two positions and approached a plateau. This implied that beyond 1.2 L/min, the level of chaotic instability was not further enhanced with an increase in the air flow rate. Hence, larger amounts of air flow would not be able to promote the turbulence further inside the BCR once the critical H was achieved at the optimal air flow rate, which was 1.2 L/min in this case.

Figure 4.10 Development of the Hurst exponent with the air flow rate at the 0.3 and 0.6 m locations in clean tap water.

Figure 4.11 shows the development of the Hurst exponent against the air flow rate at the two locations with a submerged hollow fiber membrane module. Similar to , all the H values were less than 0.5, which again confirmed the highly chaotic instability of the flow pattern inside the BCR. With the submersion of the membrane module, the overall development of H did not change significantly. For both locations, the H value exhibited a sharp increase till the air flow rate reached 0.7 L/min. From 0.7 to 1.2 L/min, the increase of H became much milder. Beyond 1.2 L/min, H became more or less constant for both locations. Thus it can be concluded that the level of chaos was not considerably enhanced inside the BCR when the air flow rate was larger than 1.2 L/min.

Figure 4.11 Development of the Hurst exponent with the air flow rate at the 0.3 and 0.6 m locations in clean tap water with a submerged hollow fiber membrane module.

It was also found that with the presence of the membrane module, the change of the gradient (from 0.8 to 0.7 L/min) at the two locations shifted slightly. The earlier appearance of the gradient change might indicate that the submersion of the membrane module affected the local hydrodynamics such that the same level of turbulence could be achieved with a smaller air flow rate. This finding is potentially useful for system energy reduction.

Again, the fact that the turbulence fluctuation reaches a maximum at the optimal air flow rate is critically important for optimizing the performance of multiphase reactors. In this study the implementation of the Hurst analysis to quantify the hydrodynamics was proven to be successful for a BCR both with and without the hollow fiber membrane module. By evaluating the H value at various air flow rates using the dynamic pressure sensing approach, the optimal air flow rate can be identified within a much shorter time compared with the conventional membrane filtration tests.

4.4.2.4 Chaos Analysis

In addition to the statistical, spectral and Hurst analysis, deterministic chaos analysis was also employed to analyze the complex flow dynamics inside multiphase reactors (Kikuchi *et al.* 1997). Chaotic systems are deterministic systems whose behavior shows irregularity. Dynamic systems can be defined as chaotic if they are non-periodic, unpredictable and very sensitive to the initial conditions (Grassberger and Procaccia 1983). Generally, the evolution of a dynamic system can be determined using mathematical functions at any time if complete information on the initial conditions is known. However, it is nearly impossible to obtain the full information on the initial conditions for most practical dynamic systems such as the flow behavior inside a BCR. Thus, for chaotic PFSs the evolution can be predicted within a certain time scale and coherence can be used to examine the relation between the two sets of signals. Nevertheless, the predictions become grossly inaccurate beyond the time scale (Ruelle 1979; Doherty and Ottino 1988; Ruelle 1995; Letzel *et al.* 1997; van Ommen *et al.* 2011). In order to explore and extract the information hidden in the PFSs, the largest Lyapunov exponent (*LLE*) was calculated to quantify the hydrodynamics of the BCR.

The Lyapunov exponent λ measures the divergent speed of two initially near trajectories and characterizes the instability of dynamic systems. The number of Lyapunov exponent is equal to the dimension of the phase space and at least one of the Lyapunov exponents is positive if the dynamic system is chaotic (Ruelle 1979; Wolf *et al.* 1985). Normally, it is unnecessary to reckon all Lyapunov exponents of the time series since the *LLE* actually indicates the chaos degree of the system (lishun *et al.* 2009). The value of *LLE* is calculated using the following equation:

$$\lambda_{\max} = \frac{1}{t_M - t_0} \sum_{k=1}^M \log \frac{L(t_k)}{L(t_{k-1})} \quad (4.14)$$

where $L(t_k)$ is the distance between the two points evolved from the earlier separation $L(t_{k-1})$ (Swinney and Gollub 1986).

Figure 4.12 shows the values of *LLE* against the air flow rates at the 0.3 and 0.6 m locations in clean tap water. All the positive values of the *LLE* indicated the chaotic nature of the flow pattern inside the BCR. Despite the positive values, the *LLE* decreased continuously with the air flow rate for both locations from 0.2 to 0.9 L/min. From 0.9 to 1.3 L/min, the *LLE* still decreased but at a much slower pace. Beyond 1.3 L/min, the *LLE* gradually flattened out. The results suggested that with an increase in air flow rate, the flow became more and more chaotic and the *LLE* eventually reached a plateau which implied the establishment of the turbulence saturation condition inside the BCR. Hence, an optimal air flow could be attained at 1.3 L/min since further increase in the air flow rate did not change the *LLE* value significantly.

Figure 4.12 Development of the *LLE* with the air flow rate at the 0.3 and 0.6 m locations in clean tap water.

Figure 4.13 shows the development of the *LLE* with the air flow rate at the two locations in clean tap water with the submerged hollow fiber membrane module.

Compared with Figure 4.12, the magnitudes of the *LLE* fluctuated within a smaller range and a three-stage development was also observed. When the air flow rate increased from 0.2 to 0.8 L/min, the *LLE* values fluctuated and no particular trend could be identified. From 0.8 to 1.2 L/min, a local minimum existed for both locations. Beyond 1.2 L/min, the development of the *LLE* tended to be stable which indicated no further enhancement of the mixing condition inside the BCR with increasing air flow rate. Since the range of the *LLE* was about the same in and Figure 4.13, it could be concluded that with the submersion of the membrane module, the overall degree of flow instability was not changed substantially. In addition, the optimal air flow rate of 1.2 L/min could be determined since the highest level of flow turbulence was achieved under this condition.

Figure 4.13 Development of the *LLE* with the air flow rate at the 0.3 and 0.6 m locations in clean tap water with the submerged hollow fiber membrane module.

4.5 Conclusions

In this chapter the utilization of a novel dynamic pressure sensing approach was examined to quantify the hydrodynamic characteristics of the BCR with and without the submersion of the membrane module. The experimental results show that this technique is applicable under various operating conditions, and is potentially useful for industrial applications. The standard deviation of the PFSs reaching a maximum at the optimal air flow rate implied the saturation turbulence condition inside the BCR. The presence of the solid particles was confirmed to have a significant impact on the flow hydrodynamics. By applying the Hurst and chaos analysis, it was confirmed that the nonlinear parameters, LLE and H , could be used to determine the optimal air flow rate. Overall, this novel dynamic pressure sensing technique can be applied to optimize the BCR system performance within a shorter time compared with the conventional membrane filtration tests. This robust and low cost method can be further deployed in industry such as water and wastewater treatment plants for system optimization and cost reduction purposes.

5. Computational Fluid Dynamics (CFD)

Modelling

5.1 Introduction

Investigations of the hydrodynamics of BCRs mainly rely on experiments and numerical modeling. Since the acquisition of reliable data for the flow dynamics in three-dimensional laboratory scale BCRs is difficult (Sanyal *et al.* 1999), numerical simulations of the hydrodynamic conditions are gaining popularity. The development of a well-calibrated numerical model will also allow the achievement of an optimal design configuration within a much shorter time frame, compared to just basing it on trial-and-error approaches with laboratory or pilot scale testing.

By definition CFD is the analysis of systems involving fluid flow, heat transfer and associated phenomena such as chemical reactions by means of computer based simulations (Versteeg and Malalasekera 2007). CFD codes are structured around the numerical algorithms to tackle fluid flow problems. Basically, five steps are involved for numerical model simulations. First of all, the 2D or 3D geometry is established. Subsequently, the volume of the domain is divided into discrete cells. Next, physical equations including the governing equations and boundary conditions are defined. After this, numerical simulations are performed and the equations are solved iteratively until converged results are achieved. Finally, a post-processor is used to analyse and visualize the numerical solutions.

By applying CFD the gas-liquid flow fields inside the BCR can be obtained with the primary features which are necessary for system control and operation. However, since the flow inside the BCR is highly turbulent, the complex flow pattern can be hard to predict. Some excessive simplifications of the 3D model further limit the

accuracy of the simulation of the hydrodynamics (Silva *et al.* 2012). In particular, the proper description of the interfacial forces and the turbulence remain the major concern (Tabib *et al.* 2008).

In this chapter 3D numerical simulations are conducted to predict the transient as well as the time-averaged flow behavior inside a laboratory-scale BCR. The simulation of a single BCR with an air source is established first, and the experimental results measured by Gan *et al.* (2011) are used to validate the numerical solutions. Subsequently, the BCR geometry is re-generated to include a hollow fiber membrane module. In terms of the membrane configuration, initially it is approximated as a solid block and later refined to twenty solid tubes so as to consider the flow around individual fibers. The numerical solutions are compared and verified with those obtained by the membrane filtration tests and dynamic pressure sensing measurements. Overall, the results from the numerical model are able to provide supplementary information on the flow dynamics which helps determine the optimal air flow rate for a specific membrane system. Furthermore, the mixing conditions can be assessed based on the numerical predictions for better fouling control.

5.2 Literature Review

5.2.1 Multiphase Models

Two CFD approaches, the Euler-Euler and Euler-Lagrange models, are normally adopted to simulate gas-liquid two-phase flow. For the Euler-Euler model each element of the finite volume contains a respective fraction of the disperse gas phase and the continuous liquid phase whose sum is equal to one. For the Euler-Lagrange model, each individual bubble movement is tracked and the continuous phase is

solved using the equation-of-motion (Sokolichin and Eigenberger 1994). In the following a summary of these two approaches is given.

Sokolichin and Eigenberger (1994) reviewed the modeling of gas-liquid flow in BCRs and loop reactors using both the Euler-Euler and Euler-Lagrange approaches. Compared with steady-state solutions, dynamic simulations were found to be suitable for uniformly aerated BCRs above a minimum gas flow velocity. The long-term averaging of the dynamic solutions led to the well-known flow structure with the reduced gas content along the column and downflow near the wall. In addition, the pros and cons between the Euler-Euler and Euler-Lagrange formulations were considered. However, no particular turbulence model was suggested in this study.

Becker *et al.* (1994) compared their experimental results measured for a flat BCR with the numerical results solved using the 2D two-phase Euler-Euler model by Sokolichin and Eigenberger (1994). It was found that the level of turbulence was not well described since the actual turbulent viscosity was not taken into account. The standard $k-\varepsilon$ closure was found to overestimate the turbulent viscosity to a great extent, and hence no satisfactory agreement could be achieved.

Delnoij *et al.* (1997) studied the two-dimensional, time-dependent motion of the gas bubbles based on a mixed Eulerian-Lagrangian approach. All the relevant forces acting on the bubbles such as the lift, drag, added or virtual mass forces were involved in the model. The numerical results were shown to be consistent with Becker *et al.*'s (1994) experimental data. They concluded that the added mass force and the lift force could not be neglected in the simulation and the history of the two-phase flow behavior strongly affected the flow structure ultimately obtained in the BCR.

Sokolichin *et al.* (1997) established a dynamical, two-phase flow model using two discretization methods, namely, UPWIND and Total Variation Diminution (TVD), to investigate the numerical diffusion effects inside a BCR. Locally and uniformly aerated BCRs were simulated respectively, using both Euler-Euler and Euler-Lagrange models. They concluded that the numerical solutions agreed quantitatively between the two models with the TVD method while for the

UPWIND technique, the results overestimated the naturally appearing diffusion considerably.

Sokolichin and Eigenberger (1999) conducted 2D and 3D simulations on a locally aerated flat BCR using laminar and turbulent models of Euler-Euler type. They claimed that neither the 2D laminar model nor the 2D $k-\varepsilon$ turbulent closure could fully reproduce the flow structure inside the BCR. Only with the application of the 3D turbulent model, the numerical predictions showed a good agreement with the experimental results. Hence, they believed that the 3D effect as well as the influence of discretization significantly affect the accuracy of the numerical predictions. Pfleger et al. (1999) and Borchers et al. (1999) also drew similar conclusions that the use of 3D simulations could obtain better results compared with 2D models.

Mudde and Simonin (1999) carried out 2D and 3D numerical simulations of a meandering BCR in a rectangular flat geometry. They illustrated that an oscillating flow was present using three-dimensional calculations, which was closer to reality. They showed that the change of the standard $k-\varepsilon$ closure to a low Reynolds $k-\varepsilon$ closure did not lead to a significant difference in the numerical results.

Lain *et al.* (1999) adopted the Euler-Lagrange model to study the time-dependent flow fluctuations in a BCR. They suggested that the bubble size distribution affected the predictions of the gas velocities. Moreover, the appropriate source term also needed to be considered to calculate the bubble-induced turbulence.

Lain *et al.* (2002) studied the modeling of bubble-induced turbulence using the Euler-Lagrange framework. They proposed that the modeling of the bubble source terms in the k - and ε - equations was the main issue to be solved in order to correctly predict the hydrodynamic behavior of a BCR. The numerical results were compared quantitatively with the experimental data. A good agreement was observed which indicated the possibility of obtaining better results using 3D transient simulations.

Krishna *et al.* (1999) (2000) and van Baten (2001) provided a sophisticated study on the flow structures inside two-phase BCRs. The Eulerian framework was employed to describe each phase and the $k-\varepsilon$ turbulent closure was used to predict the

turbulence. They showed that basically three different phases existed inside a BCR, namely, the liquid phase, small bubbles and large bubbles without including the interactions between the two bubble classes. They believed the three-phase Eulerian simulation can be a powerful tool for design and scale-up of BCRs.

Ranade and Tayalia (2001) used the Euler-Euler approach with the $k-\varepsilon$ turbulence closure to study the effect of different sparger designs on the hydrodynamics of a BCR. The computational results for 2D rectangular and 3D cylindrical geometries were compared. It was shown that the 3D simulations relatively improved the agreement with the experimental data.

Deen *et al.* (2000) simulated the gas-liquid flow using the standard $k-\varepsilon$ closure with the consideration of only the drag force. They found that the bubble plume moved into one corner and stayed there for the rest of the simulation time, giving rise to an asymmetric velocity profile. This was proven to be caused by the overestimation of the viscosity by the $k-\varepsilon$ closure. In 2001 they used Large Eddy Simulation (LES) to study the hydrodynamic behavior and compared the results with those obtained using the $k-\varepsilon$ closure in Deen *et al.* (2000). This time, both the lift and virtual mass forces were taken into account. It was found that good quantitative agreement with experimental data was obtained with the successful prediction of transient fluid behavior. Furthermore, LES showed better results than the simulations using the $k-\varepsilon$ closure since it was able to capture the transient movement of the bubble plume (Deen *et al.* 2001).

Pfleger and Becker (2001) used the Euler-Euler two-phase flow model including the standard $k-\varepsilon$ closure to study the turbulent effect in a cylindrical BCR. The computational results showed over-prediction of the gas holdup due to grid coarseness and simplification of the inlet configuration. The incorporation of bubble-induced turbulence enhanced the predictability of the radial profile of the axial velocity. Better quantitative results could be obtained using dynamic simulations despite the higher computational effort.

Lakehal *et al.* (2002) adopted the LES model to study isothermal, highly turbulent bubbly shear flow at very low void fraction (1.9%). They suggested that the optimal

grid concentration had to be such that the ratio of the cut-off filter to the bubble diameter was about 1.5 times the bubble diameter. For 3D simulations they confirmed that the dynamic approach of Germano (Germano *et al.* 1991) did not perform better than the Smagorinsky model.

Bombardelli *et al.* (2006) conducted numerical simulations of the meandering movement of bubble plumes using both the $k-\varepsilon$ and LES models. They showed that with the finite element LES approach in 3D, the meandering could be reproduced with a very simple bubble-slip model and Smagorinsky's subgrid closure. For the $k-\varepsilon$ closure, meandering appeared within a relatively short simulation time but with a longer period, the flow settled to a steady state. The period of meandering predicted by the simulation was found to agree with the measured data.

Zhang *et al.* (2006) investigated the gas-liquid flow in two square-section BCRs and employed the sub-grid scale model and the $k-\varepsilon$ closure to simulate the turbulent viscosity of the liquid phase. It was found that with an increase in the sub-grid scale model constant C_s , the turbulent viscosity increased, and the mean velocity profiles became steeper. For C_s ranging from 0.08 to 0.10, the numerical solutions had a fairly good agreement with the experimental data. It was also found that in a taller column ($H/D=6$), Tomiyama's (2004) interfacial force closures produced a very good solution, while in the lower column, Ishii and Zuber's (1979) drag model yielded a better prediction of the slip velocity. In addition, they suggested that it might be necessary to monitor the bulk water free surface and obtain a good solution for the liquid phase velocity in the top section of the column.

Kulkarni *et al.* (2007) conducted a comprehensive study on BCR flow behavior both experimentally and numerically with two different air spargers. The modeling was carried out using the standard $k-\varepsilon$ closure and both the lift and virtual mass forces were considered. The prediction gave good results in bulk the regime but some discrepancy occurred in the near sparger and disengagement zone. They also pointed out the necessity of conducting a sensitivity analysis of different interfacial forces.

Simonnet *et al.* (2008) simulated a quasi-2D BCR in order to predict the regime transitions. They mentioned that the regime transitions were governed by the bubble-bubble interactions. They used the drag force correlation which was a function of the local gas volume fraction and the bubble diameter, and showed that the use of this drag correlation could reproduce some typical characteristics of the different regimes.

Laborde-Boutet *et al.* (2009) carried out 3D simulations of a BCR using the Euler-Euler framework, and focused on the importance of the numerical setup and the effect of turbulence closures in the RANS approach. They addressed the reason to choose third-order numerical schemes and recommended the application of the RNG $k-\varepsilon$ closure for the turbulence modeling due to its better depiction of the turbulent dissipation rate and turbulent viscosity.

Ekambara and Dhotre (2010) conducted 3D simulations of a BCR using the Euler-Euler approach. They assessed the performance and applicability of different turbulence closures and compared the numerical solutions against the experimental data of Kulkarni *et al.* (2007). They concluded that LES gave more accurate results especially near the sparger region although more computer resources were required. On the other hand, RANS showed good performance for the steady and time-averaged features of the flow.

Silva *et al.* (2012) conducted 3D Eulerian simulations of gas-liquid flows in a cylindrical BCR to study the interfacial forces in the heterogeneous flow regime. The interfacial forces including drag, lift and turbulence dispersion were analyzed. A comparison between the $k-\varepsilon$ and Reynolds Stress (RSM) turbulence closures were performed as well. The numerical results showed that the RSM model was able to give better representations of the flow without any simplifications of the boundary conditions near the sparger.

Generally, the vast majority of turbulent flow computations have been carried out with procedures based on the Reynolds-Averaged-Navier-Stokes (RANS) equations. The most common RANS turbulence closures (see Table 5.1) are classified on the

basis of additional transport equations that need to be solved along with the RANS flow equation.

Table 5.1: The summary of RANS-based turbulence models

No. of extra transport equations	Name
Zero	Mixing length model
One	Spalart-Allmaras model
Two	$k-\varepsilon$ model
	$k-\omega$ model
	Algebraic stress model
Seven	Reynolds stress model

For the mixing length model, as it is completely incapable of describing flows with separation and recirculation, it is almost completely ignored in commercial CFD programmes today. For the Spalart-Allmaras model, it is not suitable in computing massively separated flows, free shear flows and decaying turbulence. Hence, it is also seldom used now. The $k-\varepsilon$ turbulence closure is robust, economical and converges relatively easily. It is widely used and predicted reasonably results for many flow. In principle, $k-\omega$ model can also be used as it is also well established for aerospace applications. In the present study, as the $k-\varepsilon$ turbulence closure is widely preferred for industrial internal flow computations, it is employed to predict the flow behaviours inside the BCR. For the Algebraic stress model, the algebraic equations themselves are not very stable. However, the computational time is significantly longer than that of the standard $k-\varepsilon$ turbulence closure. For the Reynolds stress model, although it is the physically most complete, the computational time is quite long due to the seven additional equations needed to be solved. Although LES is a more advanced method for turbulence simulations over the $k-\varepsilon$ closure since it is based on both spatial and temporal simulations (as opposed to spatial only in $k-\varepsilon$ closure), the intensive need for computing resources as well as the necessity of an efficient multi-phase numerical sub-model to simulate the bubbling-induced flow make the simulation difficult. Hence, in this study the standard $k-\varepsilon$ closure was applied.

5.2.2 Simulation of MBRs

Ndinisa *et al.* (2006) performed two-phase numerical simulations with the submersion of flat sheet membranes to investigate the fouling control processes induced by aeration. The particle imaging software, AnalySIS, was adopted. A different number of baffles used in the experiments for mixing enhancement was also simulated. The results showed that higher gas flow rate, larger bubble size and the presence of baffles all contributed to fouling control.

Kang *et al.* (2008) conducted 3D, multiphase CFD simulations for both pilot and plant scale MBRs to study the global hydrodynamics of the systems. An integrated numerical model with the membrane module simulated as porous medium was used for approximation. Good agreement was achieved between the numerical predictions and experiment data on the hydrodynamics, which validated the applicability of the integrated model. They also recommended that different simulation criteria should be applied for the pilot and plant scale MBR systems.

Prieske *et al.* (2008) examined the hydrodynamics of a pilot scale MBR with different air diffuser configurations experimentally and numerically. The results showed that the airlift loop configuration was better in fouling reduction owing to the higher cross-flow velocities generated. They were also the first to propose an approach for calculating the circulation velocity. However, the comparison between the experimental results and numerical solutions showed that the determination of a proper bubble size greatly affected the accuracy of the numerical predictions.

Brannock *et al.* (2010) developed a numerical model which accounted for aeration, sludge rheology and membrane and reactor configurations, and applied it to full-scale MBR systems for studying the hydrodynamics. Experiments were also conducted on two different full-scale MBRs for validation. The predictions generally agreed well with the experimental data.

Based on Brannock *et al.*'s (2010) model, Wang *et al.* (2010) simulated the membrane configuration using a porous medium to consider the hydrodynamic conditions surrounding the membranes. Experiments were carried out to measure the flow resistance of the hollow fibre membrane bundle which was then used to simulate the porosity of the filtration zone. The model was then applied to the full-scale submerged MBR using the calibration information. They found that the flow behavior predicted by the numerical model was consistent with the observations in the full-scale plant.

Although CFD provides a method for prediction of the global hydrodynamics of MBRs, the detailed flow conditions around the membrane modules was not reported extensively in the past. As discussed above, generally, the membrane modules were approximated as a solid or porous medium. This can be attributed to the huge number of hollow fibres typically used in industrial MBRs, which makes it very difficult and computationally costly to simulate the individual fibres. However, the hydrodynamics around and within the fibres is the key information for fouling mitigation. If the flow conditions near the fibres are known, improved fouling control, such as modifying the membrane configuration to reduce non-even flow distribution around the fibres, can be performed. In this study we make the first attempt by simulating the hollow fibre membrane bundle as real individual fibres in a laboratory scale BCR so as to obtain an in-depth understanding on the hydrodynamics of the MBR including the flow fields around and within the fibers.

5.3 Conservation Laws of Fluid Motion

5.3.1 Governing Equations

The numerical simulations presented here are based on the two-phase Euler-Euler approach with the mixture model employed to describe the flow behaviour inside the BCR. Currently there are two approaches for the numerical calculation of

multiphase flows, namely, the Euler-Lagrange approach and the Euler-Euler approach. In the Euler-Lagrange approach, particles are tracked on the level of a single particle where particle refers to either a solid particle or a gas/fluid bubble/droplet. In Euler-Euler models, all phases are treated as continuous. However, it can also be used to model dispersed flows when the overall motion of particles is of interest rather than tracking individual particles. As, the bubble plume flow pattern is the interest of the present study, the Euler-Euler approach is employed. The model solves the equations of continuity, momentum and energy equations for the mixture. The volume fraction equations for the secondary phase and algebraic expressions for the relative velocities are also considered (Wang *et al.* 2010).

The continuity equation for the mixture is:

$$\frac{\partial}{\partial t}(\rho_m) + \nabla(\rho_m \vec{v}_m) = 0 \quad (5.1)$$

where ρ_m is the mixture density and can be computed as follows:

$$\rho_m = \sum_{k=1}^n \alpha_k \rho_k \quad (5.2)$$

$\vec{v}_m$ is the mass-averaged mixture velocity computed as follows:

$$\vec{v}_m = \frac{\sum_{k=1}^n \alpha_k \rho_k \vec{v}_k}{\rho_m} \quad (5.3)$$

and α_k is the volume fraction for any phase k and n is the number of phases.

The momentum equation for the mixture is:

$$\frac{\partial}{\partial t}(\rho_m \vec{v}_m) + \nabla(\rho_m \vec{v}_m \vec{v}_m) = -\nabla p + \nabla[\mu_m \vec{T}] + \rho_m \vec{g} + \vec{F} + \nabla(\sum_{q=1}^n \alpha_q \rho_q \vec{v}_{dr,q} \vec{v}_{dr,q}) \quad (5.4)$$

where $\vec{F}$ is the body force and μ_m is the mixture viscosity expressed as follow:

$$\mu_m = \sum_{k=1}^n \alpha_k \mu_k \quad (5.5)$$

$\vec{v}_{dr,k}$ is the drift velocity for phase k with respect to the mixture velocity:

$$\vec{v}_{dr,k} = \vec{v}_k - \vec{v}_m = \vec{v}_{kq} - \sum_{k=1}^n \alpha_k \rho_k \vec{v}_{kq} \quad (5.6)$$

and $\vec{v}_{dr,p}$ is the drift velocity for the secondary phase p :

$$\vec{v}_{dr,p} = \vec{v}_p - \vec{v}_m = \vec{v}_{pq} - \sum_{k=1}^n \alpha_k \rho_k \vec{v}_{kq} \quad (5.7)$$

where $\vec{v}_{kq}$ is the slip velocity of phase k with respect to the velocity of the primary phase q . $\vec{v}_{pq}$ is the slip velocity of secondary phase p with respect to the velocity of the primary phase q .

$\vec{T}$ is the rate-of-deformation tensor, which can be computed as:

$$\vec{T} = \nabla \vec{v}_m + \nabla \vec{v}_m^T \quad (5.8)$$

where $\nabla \vec{v}_m^T$ means the transpose of the rate-of-strain tensor.

The volume fraction equation for the secondary phase p is:

$$\frac{\partial}{\partial t}(\alpha_p \rho_p) + \nabla(\alpha_p \rho_p \vec{v}_m) = -\nabla(\alpha_p \rho_p \vec{v}_{dr,p}) + \sum_{q=1}^n (\dot{m}_{pq} - \dot{m}_{qp}) \quad (5.9)$$

where $\dot{m}_{pq}$ is the mass transfer from phase p to phase q , and $\dot{m}_{qp}$ is the mass transfer from phase q to phase p .

Other than the mixture model shown above, the Eulerian model can also be used by solving a set of continuity and momentum equations for each individual phase. Both the mixture and Eulerian models are suitable for simulations of bubbly flows when the gas volume fraction is low. However, the mixture model is superior when there is a wide distribution of the dispersed phase. In addition, the complexity of the Eulerian model makes it less computationally stable than the mixture model. Therefore, the mixture multiphase model is employed here.

5.3.2 Interfacial Forces

The interfacial force term represents the interaction forces between the continuous and dispersed phases (Joshi, 2001; Tabib *et al.*, 2008; Silva *et al.*, 2012). When stationary, the forces exerted on a motionless bubble are solely due to pressure and gravity. With motion, however, interfacial forces are induced with resulting momentum transfer across the interface between the two phases (Tabib *et al.*, 2008). In terms of simulations of the hydrodynamics inside the BCR, there are mainly three types of interfacial forces, namely, drag, lift, and virtual mass force (Joshi, 2001). The lift force is induced by the velocity gradients in the primary phase flow field (Joshi, 2001; Tabib *et al.*, 2008; Silva *et al.*, 2012). The inclusion of lift force is in general computationally excessive for convergence and it is often assumed to be significant only for large bubbles. The virtual mass force occurs due to the encounter between the inertia of the primary phase mass and the accelerating secondary phase. Hunt *et al.* (1987) suggested that for large diameter bubble columns ($D > 0.15\text{m}$), the interfacial coupling term is dominated by the drag force and the role of virtual mass force can be ignored (Hunt *et al.*, 1987; Joshi, 2001).

In this case only the drag force per unit particle mass is included due to the small bubble size distributions. The equations are listed below:

$$F_D = \frac{18\mu_q}{\rho_p d_p^2} \frac{C_D \text{Re}}{24} (\vec{v}_p - \vec{v}_q) \quad (5.10)$$

$$C_D = \begin{cases} 24(1 + 0.15 \text{Re}^{0.0687}) / \text{Re}, & \text{Re} \leq 1000 \\ 0.44, & \text{Re} > 1000 \end{cases} \quad (5.11)$$

$$\text{Re} = \frac{\rho_q |\vec{v}_p - \vec{v}_q| d_p}{\mu_q} \quad (5.12)$$

where C_D is the drag coefficient, Re is the relative Reynolds number, μ_q is the dynamic viscosity of the primary phase q , ρ_p is the density of the secondary phase p , and d_p is diameter of secondary phase p .

5.4 Turbulence Modelling

Turbulence causes the appearance in the flow of eddies with a wide range of length and time scales that interact in a dynamically complex way (Versteeg and Malalasekera 2007). In this regard various numerical methods have been developed to study the important features of the turbulence. Basically, three categories are most widely adopted. The best known ones are the turbulence models for Reynolds-averaged Navier-Stokes (RANS) including the k - ε and Reynolds Stress model, etc. In addition to this method, LES and direct numerical simulation (DNS) are also commonly used. In this section, the k - ε and LES models are discussed.

5.4.1 k - ε Turbulence Closures for RANS Models

5.4.1.1 *Standard k - ε closure*

Based on the RANS equations, instantaneous variables in the momentum equations are defined as the sum of an averaged value and its fluctuation. The additional unknown variables, Reynolds stresses, $\bar{T}_t = \rho \overline{\vec{v}_m' \vec{v}_m'}$, which represent the effects of turbulence are added to the total stress. Over the years, many different turbulent closures had been developed to model the Reynolds stresses, among which the most widely used is the k - ε turbulence closure. The two-equation models focus on the mechanisms that affect the turbulent kinetic energy k and simulate the transport equations for k (Eqn. 5.13) and its dissipation rate ε (Eqn. 5.14).

$$\frac{\partial}{\partial t}(\rho k) + \frac{\partial}{\partial x_i}(\rho k u_i) = \frac{\partial}{\partial x_j} \left[\left(\mu + \frac{\mu_t}{\sigma_k} \right) \frac{\partial k}{\partial x_j} \right] + G_k + G_b - \rho \varepsilon - Y_M + S_K \quad (5.13)$$

$$\frac{\partial}{\partial t}(\rho \varepsilon) + \frac{\partial}{\partial x_i}(\rho \varepsilon u_i) = \frac{\partial}{\partial x_j} \left[\left(\mu + \frac{\mu_t}{\sigma_\varepsilon} \right) \frac{\partial \varepsilon}{\partial x_j} \right] + C_{1\varepsilon} \frac{\varepsilon}{k} (G_k + C_{3\varepsilon} G_b) - C_{2\varepsilon} \rho \frac{\varepsilon^2}{k} + S_\varepsilon \quad (5.14)$$

where μ_t is the turbulent (or eddy) viscosity calculated by:

$$\mu_t = \rho C_\mu \frac{k^2}{\varepsilon} \quad (5.15)$$

and C_μ is a constant and equals to 0.09 for standard k - ε closures.

G_k represents the generation of turbulent kinetic energy due to the mean velocity gradients and is defined as follows from the exact equation for the transport k :

$$G_k = -\overline{\rho u'_i u'_j} \frac{\partial u_j}{\partial x_i} \quad (5.16)$$

G_b is the generation of turbulent kinetic energy due to buoyancy:

$$G_b = \beta g_i \frac{\mu_t}{Pr_t} \frac{\partial T}{\partial x_i} \quad (5.17)$$

where Pr_t is the turbulent Prandtl number for energy and equals to 0.85 for the standard k - ε closure. g_i is the component of the gravitational vector in the i th direction. The coefficient of thermal expansion β , is defined as:

$$\beta = -\frac{1}{\rho} \left(\frac{\partial \rho}{\partial T} \right)_p \quad (5.18)$$

Y_M represents the contribution of the fluctuating dilatation in compressible turbulence to the overall dissipation rate:

$$Y_M = 2\rho\varepsilon M_t^2 \quad (5.19)$$

where M_t is the turbulent Mach number. Since we assume incompressible flow in this case, the item Y_M is not included in the computation.

$C_{1\varepsilon}$, $C_{2\varepsilon}$ and $C_{3\varepsilon}$ are constants that have the following default values:

$$C_{1\varepsilon} = 1.44, \quad C_{2\varepsilon} = 1.92$$

σ_k and σ_ε are the turbulent Prandtl numbers for k and ε and are equal to 1.0 and 1.3, respectively. S_k and S_ε are the user-defined source terms.

5.4.1.2 Realizable k - ε closure

The realizable k - ε closure is a relatively recent development (Versteeg and Malalasekera 2007). It computes the turbulent viscosity with a variable coefficient and determines the dissipation equation from an exact equation for the transport of the mean-square vorticity fluctuation. Hence, it gives more precise solutions than the standard k - ε closure in theory in predicting flows such as separated flows and flows with complex secondary flow features. In this regard it is adopted in this study.

For the realizable k - ε closure, the transport equation for k is the same as the standard k - ε closures. The equation for ε is however expressed alternatively as:

$$\frac{\partial}{\partial t}(\rho\varepsilon) + \frac{\partial}{\partial x_i}(\rho\varepsilon u_i) = \frac{\partial}{\partial x_j}[(\mu + \frac{\mu_t}{\sigma_\varepsilon}) \frac{\partial \varepsilon}{\partial x_j}] + \rho C_1 S \varepsilon - \rho C_2 \frac{\varepsilon^2}{k + \sqrt{\nu \varepsilon}} + C_{1\varepsilon} \frac{\varepsilon}{k} C_{3\varepsilon} G_b + S_\varepsilon \quad (5.20)$$

$$C_1 = \max[0.43, \frac{\eta}{\eta + 5}] \quad (5.21)$$

$$\eta = S \frac{k}{\varepsilon} \quad (5.22)$$

$$S = \sqrt{2S_{ij}S_{ij}} \quad (5.23)$$

where ν is the turbulent kinematic viscosity. η represents the scalar measure of the deformation tensor and S is the module of the mean rate-of-strain tensor.

As mentioned earlier, the constant C_μ for standard k - ε closures is represented by a function of the mean strain and rotation rates, the angular velocity of the system rotation, and the turbulent fields. It is computed as follows:

$$C_\mu = \frac{1}{A_0 + A_s \frac{kU^*}{\varepsilon}} \quad (5.24)$$

$$U^* \equiv \sqrt{S_{ij}S_{ij} + \underline{\Omega}_{ij}\underline{\Omega}_{ij}} \quad (5.25)$$

$$\underline{\Omega}_{ij} = \Omega_{ij} - 2\varepsilon_{ijk}\omega_k \quad (5.26)$$

$$\Omega_{ij} = \overline{\Omega}_{ij} - 2\varepsilon_{ijk}\omega_k \quad (5.27)$$

where $\overline{\Omega}_{ij}$ is the mean rate-of-rotation tensor viewed in a rotating reference frame with the angular velocity ω_k . The model constants A_0 and A_s are given as:

$$A_0 = 4.04 \quad (5.28)$$

$$A_s = \sqrt{6} \cos \phi \quad (5.29)$$

$$\phi = \frac{1}{3} \cos^{-1}(\sqrt{6}W) \quad (5.30)$$

$$W = \frac{S_{ij}S_{jk}S_{ki}}{\underline{S}^3} \quad (5.31)$$

$$\underline{S} = \sqrt{S_{ij}S_{ij}} \quad (5.32)$$

$$S_{ij} = \frac{1}{2} \left(\frac{\partial u_j}{\partial x_i} + \frac{\partial u_i}{\partial x_j} \right) \quad (5.33)$$

5.4.2 Large Eddy Simulation (LES)

LES is more precise in modelling turbulence compared with traditional RANS turbulent closures (Deen *et al.* 2001). The wide application of a general-purpose turbulence model is constrained mainly due to the different behaviour of large and

small eddies. Large eddies are dictated by the geometries and boundary conditions of the flow involved. Moreover, the momentum, mass, energy and other passive scalars are transported mainly by large eddies. On the other hand, the smaller eddies are nearly isotropic and have a more universal behaviour (Versteeg and Malalasekera 2007). While RANS closures treat all eddies with a single turbulence closure, LES resolves the large integral scales and models the dissipation scales separately.

In principle, LES filters the smaller scale eddies to be modelled using a sub-grid scale model whereas the larger scale eddies are retained and solved using an instantaneous calculation. First of all, the governing equations employed for LES are obtained by filtering the time-dependent Navier-Stokes equations. The filtered variable is defined by:

$$\bar{\phi}(\mathbf{x}) = \int_D \phi(\mathbf{x}') G(\mathbf{x}, \mathbf{x}') d\mathbf{x}' \quad (5.34)$$

where D is the fluid domain, G is the filter function that determines the scale of the resolved eddies. The overbar indicates space-filtering.

The filtered unsteady Navier-Stokes equations are listed as below:

$$\frac{\partial \bar{\rho}}{\partial t} + \frac{\partial}{\partial x_i} (\bar{\rho} \bar{u}_i) = 0 \quad (5.35)$$

$$\frac{\partial}{\partial t} (\bar{\rho} \bar{u}_i) + \frac{\partial}{\partial x_j} (\bar{\rho} \bar{u}_i \bar{u}_j) = \frac{\partial}{\partial x_j} (\sigma_{ij}) - \frac{\partial \bar{p}}{\partial x_i} - \frac{\partial \tau_{ij}}{\partial x_j} \quad (5.36)$$

where σ_{ij} is the stress tensor due to molecular viscosity defined by:

$$\sigma_{ij} \equiv [\mu (\frac{\partial \bar{u}_i}{\partial x_j} + \frac{\partial \bar{u}_j}{\partial x_i})] - \frac{2}{3} \mu \frac{\partial \bar{u}_i}{\partial x_i} \delta_{ij} \quad (5.37)$$

and τ_{ij} is the sub-grid scale stress defined by

$$\tau_{ij} \equiv \overline{\rho u_i u_j} - \bar{\rho} \bar{u}_i \bar{u}_j \quad (5.38)$$

Since the sub-stresses are unknown and require modelling, it is computed from the following equation basing on the Boussinesq hypothesis:

$$\tau_{ij} - \frac{1}{3} \tau_{kk} \delta_{ij} = -2\mu_t \bar{S}_{ij} \quad (5.39)$$

where μ_t is the sub-grid scale turbulent viscosity. The isotropic part of the subgrid scale stresses τ_{kk} is not modelled, but added to the filtered static pressure term. $\bar{S}_{ij}$ is the rate-of-strain tensor for the resolved scale defined by:

$$\bar{S}_{ij} = \frac{1}{2} \left(\frac{\partial \bar{u}_i}{\partial x_j} + \frac{\partial \bar{u}_j}{\partial x_i} \right) \quad (5.40)$$

Several models have been developed to compute μ_t and the most common one is the Smagorinsky-Lily model (Murthy and Joshi 2008), whereby the turbulent viscosity is modelled by:

$$\mu_t = \rho L_s^2 |\bar{S}| \quad (5.41)$$

where L_s is the mixing length for the subgrid scale and can be computed by:

$$L_s = \min(\kappa d, C_s \Delta) \quad (5.42)$$

where κ is the von Karman constant, d is the distance to the closest wall, C_s is the Smagorinsky constant and Δ is the local grid scale.

and

$$|\bar{S}| \equiv \sqrt{2\bar{S}_{ij}\bar{S}_{ij}} \quad (5.43)$$

Due to the unsteady nature of the LES model, the computational requirement is much higher than those of many classical turbulence closures (Versteeg and Malalasekera 2007). However, it is superior in flows where the main phenomena are dominated by the existence of large coherent structures or eddies including BCRs (Dhotre *et al.* 2008). Choosing a proper turbulence model is essential to study the flow behaviour in a chaotic system.

5.5 Model of BCR

5.5.1 Model Geometry

GAMBIT 2.4.6 is used to reproduce the geometry of the experimental BCR simulated and establish meshes. It allows the user to create basic 2D and 3D geometries and further manipulate to change any other physical properties including the geometry nature (Cimbala 2008). The simulated geometry of the cylindrical BCR, which is identical to that used by Gan *et al.*(2011), is shown in Figure 5.1. A hemispherical air stone with a diameter of 30 mm was simulated as the air sparger mounted at the central bottom of the column.

Initially, numerical simulations were carried out for three air flow rates of 0.13, 0.25 and 0.38 L/min, respectively. The simulated radial profiles of the time-averaged liquid axial velocity and air volume fraction were compared with the experimental data of Gan *et al.* (2011) at six axial heights as shown in Figure 5.1, i.e., 100, 130, 160, 200, 400 and 600 mm from the bottom of the reactor.

Figure 5.1 Schematic of the BCR (all dimensions in mm) and measurement axial locations (reproduced from Gan et al., 2011).

5.5.2 Meshing

Meshing is the construction of grids on the surface and interior of the model. The construction of grids is very important as the grids designate the cells on which the flow is being modelled and solved. Therefore, the selection of the type of meshing and grids to be used determines the accuracy and time required to compute the solutions. In this case the unstructured meshes, TGrid, were utilized to generate tetrahedral meshes for the entire domain since this kind of mesh is commonly used to perform meshing for complicated large surfaces.

A size function was employed to ensure smooth growth of the grid size in highly curved regions. The size function enabled the modification of the grid density to allow the user to obtain higher accuracy data over the region of interest, while reducing the computational cost. It was operated via a controlled parameter over the growth rate of the cell size to maintain absolute mesh integrity over the geometric areas of interest, while allowing other areas to achieve the necessary degree of accuracy. The growth rate, which represented the rate of increase of the edge length of the cells with each succeeding layer of meshing, was set at 1.1. The starting mesh size was set at 0.005 m with the maximum mesh size at 0.01 m. After meshing, the mesh quality was examined and the largest value of the equiangle skewness of the grids was kept under 0.7. Figure 5.2 shows the meshed structure of (a) the BCR and (b) the top view and (c) side view of the hemispherical inlet, respectively. It can be seen that the majority of the cells were concentrated around the air sparger.

5.5.3 Physical Properties and Boundary Conditions

After the establishment of the meshed geometry, the commercial software FLUENT was used to solve the flow and turbulent equations. FLUENT supports a wide range of modelling capabilities which enables the user to simulate various models which is particularly useful for research applications. For BCR modelling, FLUENT also offers special tools including multiphase systems which are particularly helpful.

Figure 5.2 Numerical meshes: (a) the BCR and the hemispheric inlet boundary for (b) top view and (c) side view.

1. Physical properties

The numerical simulation was conducted based on the following assumptions. First, both the air and water phase were assumed to be incompressible. Second, the air phase was taken to be spherical bubbles with a constant diameter. Note that the bubble size increased with the air flow rate (Fukuma *et al.* 1987; Saxena *et al.* 1990). Thus, the constant bubble size used should increase with the air flow rate as well. In order to achieve this correlation, we conducted visualization experiments again with the laboratory scale BCR using a high speed camera. For a specific air flow rate, the sizes of the bubbles were determined from the images right above the air inlet. Subsequently, the bubble sizes were number-averaged and this average bubble size was used in the simulation. The same procedures were employed for all the air flow rates, for which the correlation is shown in Figure 5.3. Given the small bubble size and relatively small column height involved, bubble coalescence and breakup was not observed. When the air flow rate was 0.13 L/min, a constant bubble size of 0.79 mm was used for the simulation. Table 5.2 shows the basic physical properties of the air and water phases. Here, only the predominant bubble size was used.

Figure 5.3 Estimated average bubble size for different air flow rates

Table 5.2 Physical properties of the air and water phases

Physical properties	Air	Water
Density	1.225 kg/m ³	998.2 kg/m ³
Viscosity	1.79e-05 kg/m-s	9.79e-04 kg/m-s

2. Boundary conditions

The column was filled with water to a height of 1 m. At the entrance, the mass flow rate of the gas phase was specified according to the experimental volumetric flow rate. Atmospheric pressure was defined at the exit for the gas phase pressure and the gas volume fraction was set to 1 for the backflow. Non-slip conditions were defined for all the walls of the BCR.

The governing equations were solved using second-order upwind methods for spatial discretization and the first-order implicit method for time discretization. The PISO solver was used in this study for the pressure-velocity coupling. The under-relaxation factors, time step sizes and iterations for the variables were tuned to optimize the balance between the convergence criteria (10^{-6}) and the number of iterations required. The time averaged profiles of the hydrodynamics were obtained by averaging over a period of 120 s.

5.5.4 Results and Discussion

5.5.4.1 Grid Independence Test

As the accuracy of the numerical solutions is highly dependent on the grid strategy, it is very important to ensure that the results are grid independent (Silva *et al.* 2012). Here, the grid independence test was conducted using five different numbers of tetrahedral meshing, i.e., 43531, 251976, 769068, 1609472 and 1786858 cells,

respectively. In the mesh independency evaluation, tests were performed with the realizable $k-\varepsilon$ turbulence closure. The profiles of liquid axial velocity at the centreline were compared as shown in Figure 5.4, in which D was the column diameter. The numerical results appeared to be stable when the number of cells exceeded 769068. Therefore, for all the subsequent simulations, the meshing with 769068 cells was selected for simulation.

Figure 5.4 Grid independence test: liquid axial velocity at the centreline

5.5.4.2 Liquid Velocity

Figure 5.5 shows the transient and time-averaged liquid axial velocity contours at a 0.13 L/min air flow rate using the realizable $k-\varepsilon$ turbulent closure and LES model, respectively. The liquid axial velocity is actually the velocity component along the z -axis of the BCR when Cartesian coordinate system is employed. It can be seen from Figure 5.5 (a) that the bubble wandered back and forth near the bottom portion. With the development of the bubble plume along the column, the bubble plume

became more developed and dispersed radially. Similar phenomena were found in Figure 5.5(c) where the LES turbulence model was applied. Besides, coherent structures and eddies were captured and the flow behaved highly chaotic and unstable in Figure 5.5(c). By comparing Figure 5.5(a) and (c), it was suggested that the LES model was superior in predicting the flow with the existence of large coherent structures and flow circulation. Also, the LES model showed a higher liquid velocity along the axis compared with the values obtained with the realizable $k-\varepsilon$ closure. As far as the time-averaged flow pattern was concerned, Figure 5.5(b) and (d) show symmetric profiles in terms of the liquid axial velocity distribution. With air injection, the initially static liquid phase was accelerated and entrained into the central region. The movement of liquid was slowed down by the boundary wall and thus the flow circulation was formed. In terms of the velocity magnitude, again, the LES model gave higher values along the central axis. Furthermore, the bubble plume was more concentrated along the centreline and developed to a higher axial height but with a narrower plume width for the LES model.

Figure 5.6 shows the comparison between the liquid axial velocity computed at 0.13, 0.25 and 0.38 L/min using the two turbulent closures and the experimental measurement from Gan *et al.* (2011). It is noted that the measurement from Gan *et al.* (2011)'s model is 2D while our numerical solutions were derived based on a 3D model. The spatial profile of the liquid axial velocity at six axial elevations was compared, namely, 100, 130, 160, 200, 400 and 600 mm from the bottom of the reactor. Generally, all the three air flow rates showed similar trends. In addition, the liquid axial velocity obtained from the same location of the BCR increased accordingly with the increase in air flow rates. The liquid flow was induced mainly by the drag force of the bubbles rising through the column (Gan *et al.*, 2011). The experimental observation showed the radial distribution of large scale circulation with the maximum upward velocity at the centre and downward flows near the wall. The profile also indicated that the point of flow reversal shifted towards the wall from $r/R=0.5$ to 0.8 with increasing axial elevations for all the three air flow rates. This suggested that the bubble plume was developing and extended to the cross-section of the BCR along the column height. It was found that both numerical approaches managed to capture these flow patterns as observed in the experiments.

However, the realizable $k-\varepsilon$ turbulence closure showed a flatter radial distribution profile of the liquid axial velocity with the flow reversal location taking place at around $r/R=0.6-0.8$. While for LES, it produced a narrower distribution with the flow reversal location ranging from $r/R=0.3$ to 0.8 along with a higher peak value at the centreline.

Figure 5.5 Contours of the liquid axial velocity magnitude (m/s) at a 0.13 L/min air flow rate: (a) $k-\varepsilon$ transient, (b) $k-\varepsilon$ time-average, (c) LES transient, and (d) LES time-average.

It was also found that the LES predictions agreed with the experimental measurements near the sparger ($z = 100$ and 130 mm) relatively better than the

realizable $k-\varepsilon$ closure except the center value, which the LES predictions overestimated by up to 25%, while for the realizable $k-\varepsilon$ closure, the discrepancy was within 6%. Nevertheless, the predictive performance of LES deteriorated at higher axial elevations as discussed in the following paragraph. It was known that strong meandering patterns dominated near the air sparger where the flow exhibited highly chaotic behaviour. Thus it was believed that the LES model was able to better predict the transient flow dynamics with a high level of flow instabilities.

At $z = 160$ and 200 mm, both numerical approaches were not able to reproduce the experimental measurements. The realizable $k-\varepsilon$ closure predicted a parabolic profile that was different from the relatively flat profile measured experimentally, while the LES significantly overestimated the center value with a steeper radial gradient. This discrepancy might be attributed to the single constant bubble diameter used in the numerical simulation, since in the experiments the bubble size had multiple sizes with a normal distribution peaking around 0.79 mm. In real systems, the bubble size varied and the bubble size distribution changed with the air flow rate.

On the other hand, the realizable $k-\varepsilon$ turbulent closure predicted the velocity more accurately in the middle portion while overestimating the velocity value near the air sparger. When $z=400$ and 600 mm, the numerical solutions fit perfectly with the experimental measurements for all the three air flow rates. In contrast, the LES overestimated the centre value by almost four times. Overall, the realizable $k-\varepsilon$ turbulent closure presented a good performance when the bubble plume was fully developed whereby the flow was stabilized and relatively ordered compared with the meandering regime.

Figure 5.6 Comparison between the CFD simulation results and the experimental data (Gan et al., 2011) of the liquid axial velocity at six axial elevations ($z = 100, 130, 160, 200, 400$ and 600 mm) for three air flow rates of (a) 0.13, (b) 0.25 and (c) 0.38 L/min (r is the radial co-ordinate, and R is the radius of the BCR)

5.5.4.3 Gas Volume Fraction

Figure 5.7 shows the contours of the air volume fraction at a 0.13 L/min air flow rate using the two numerical approaches. Clearly, the air volume fraction was the highest at the point of injection into the BCR. The gas fraction tended to decrease along the column height and as a result of the bubble plume development, the gas fraction also decreased radially. Again, it could be observed that the meandering pattern was dominant at the bottom portion. Comparing the two turbulence closures, the bubble plume developed at a slower pace for the LES than the realizable $k-\epsilon$ closure. The value of the gas fraction was also higher for the LES when the time-averaged contours were compared. Similar to the liquid velocity contour, symmetrical profiles were obtained for the time-averaged contours.

Figure 5.7 Contours of the air volume fraction at 0.13 L/min air flow rate: (a) $k-\epsilon$ transient, (b) $k-\epsilon$ time-average, (c) LES transient, and (d) LES time-average.

Figure 5.8 shows the comparison of the air volume fraction at the three air flow rates predicted by the two turbulent closures and the experimental results from Gan et al. (2011). The radial distributions of the air volume fraction were plotted at the same axial elevations. It can be seen that the general shapes were similar for the three air flow rates. From Figure 5.8 (a), it was apparent that the centre value of air volume fraction decreased notably with increased axial elevations in the experiments. The realizable $k-\varepsilon$ closure yielded better agreement with this trend at the centreline than the LES, which overestimated the centre value by almost 50% compared to the experiments.

Both numerical solutions showed that within a short distance above the air sparger, the width of the bubble plume was narrow with a higher gas void fraction value near the centreline. The air volume fraction gradient created the density difference that further drove the liquid circulation. Similar to the profile of the liquid axial velocity, the realizable $k-\varepsilon$ closure showed a flatter radial distribution of the air volume fraction compared with the LES predictions, except at axial elevations of 600 mm. For $z=600$ mm, both numerical approaches predicted similar profiles with a close-to-uniform air volume fraction distribution.

Near the sparger, the predictions of the air volume fraction by the realizable $k-\varepsilon$ closure agreed better with the experimental data than the LES at the column center and near the walls, which implied that the physical structure of the sparger was properly accounted for in the boundary conditions. However, at higher axial elevations, the values of air volume fraction near the wall were underestimated. Another important point was that in the experiments, the maximum local air volume fraction tended to shift towards the wall with increasing heights, which was not predicted by both numerical approaches. The discrepancy between the simulation predictions with the experimental data near wall was possibly due to the lack of the inclusion of lift forces in the calculations as well as the single size bubble used in the numerical simulations. The agreement between the that of simulation results for the air flow rates of 0.25 and 0.38 L/min was comparable to 0.13 L/min.

Figure 5.8 Comparison between the CFD simulation results and the experimental data (Gan et al., 2011) for the air volume fraction at six axial elevations ($z=100, 130, 160, 200, 400$ and 600 mm) for the three air flow rates of (a) 0.13, (b) 0.25 and (c) 0.38 L/min

5.6 Model MBR with Submerged Membrane Module

After simulating a two-phase BCR and comparing the numerical solutions with the experimental results from Gan *et al.* (2011) for model validation, a membrane module was added into the model to predict the instantaneous and time-averaged flow behaviour around the membrane module. Six air flow rates, namely, 0.2, 0.5, 1.0, 1.4, 1.7 and 2.0 L/min, were simulated. The membrane module configuration was identical to the one used in the membrane filtration tests. Initially, the membrane module was considered to be a solid block and later refined to the twenty solid tubes. In this regard, water flows around the individual fibers were considered.

5.6.1 Membrane Module Specifications

5.6.1.1 Block Membrane Module

Initially, the hollow fiber membrane module was modelled as a solid block. The detailed structure and dimensions are shown in Figure 5.9. The faces were set to be solid walls so that the velocity at the membrane surface was zero.

Unstructured grids were used for meshing and typical grid distributions are shown in Figure 5.10. It can be seen that most of the grids were concentrated around the membrane and near the air sparger, while for the rest the cell size was relatively coarse. The time-averaged profiles of the hydrodynamics were obtained by averaging over a period of 100 s. The physical properties and boundary conditions were the same as those used in Section 5.5.3.

Figure 5.9 Geometry of the block membrane: (a) top view and (b) side view

Figure 5.10 Numerical meshes for the BCR with the membrane module located at (a) 0.3 and (b) 0.6 m from the bottom of the reactor.

5.6.1.2 Hollow Fiber Membrane Module

In order to improve the predictions around the membrane module, the structure of the module was further refined from a solid block to twenty solid tubes. The geometry of the refined membrane bundle is shown in Figure 5.11 with the detailed dimensions shown in Figure 5.12.

Figure 5.11 Geometry of the fiber bundle constructed of solid tubes: (a) top view and (b) side view

Figure 5.12 Detailed dimensions for the fiber model

5.6.2 Results and Discussion

In this section the results are compared between the block membrane module (block model) and the fiber membrane module (fiber model). In order to investigate the effects of air sparging on the membrane fouling performance, specific data observation points near the membrane module were selected (see Figure 5.13). Numerical data was exported at the 0.3 and 0.6 m locations to study the hydrodynamic conditions surrounding the membrane bundle. For both locations the air and liquid axial velocities, air volume fraction and turbulent kinetic energy (TKE) were predicted and compared using the two approaches.

Figure 5.13 Top view of the BCR showing specific data coordinates ($\pm 0.03, 0$)

5.6.2.1 Air Axial Velocity

Figure 5.14 shows the contours of the air axial velocity at the two locations at the air flow rate of 1.0 L/min for the block and fiber model, respectively. It can be seen that the air axial velocity decreased along the column height for both locations with

the development of the bubble column. Comparing the block and fibre models, the fibre model was able to yield finer details around the membrane module, including the air volume fraction distribution around the fibres, i.e., trapping of bubbles. Also, the air axial velocity was higher near the air injection point for the fibre model due to the reduced blockage.

Figure 5.14 Contours of the air axial velocity magnitude (m/s): (a) 0.3 m block, (b) 0.3 m fiber, (c) 0.6 m block and (d) 0.6 m fiber

Figure 5.15 shows the development of the air axial velocity that is plotted against the air flow rate for the two models at the 0.3 and 0.6 m locations. Generally, the four cases showed a similar trend that the air axial velocity increased with the air flow rate. Initially, the air axial velocity increased rapidly when the air flow rate was small (<1.0 L/min). Beyond 1.0 L/min, the increase in air axial velocity became to moderate. The magnitudes of the air axial velocity for both the block and fibre models were around 10^{-1} m/s and smaller at the 0.6 m than the 0.3 m location, which demonstrated the significance of the air plume entrainment of the slower surrounding liquid at the bottom portion of the BCR, resulting in a decrease in the

air axial velocity at the higher column heights (Gan *et al.* 2011). Again, the fibre model generally predicted the highest air axial velocities due to its reduced blockage.

Figure 5.15 Air axial velocity against the air flow rate for the block and fiber models

Figure 5.16 shows the radial profile of the air axial velocity profile at $z =$ (a) 0.9 m, (b) 0.6 m and (c) 0.3 m locations. The membrane module was simulated at the 0.3 m location with the air flow rate of 1.0 L/min. It can be seen that the block and fibre models predicted approximately the same velocity distributions. Numerical predictions using the two models at 0.6 and 0.9 m also showed a similar trapezoidal trend which remained approximately constant from $r/R = 0$ to 0.75 and then decreased steeply and eventually reached 0 m/s at the column wall. The occurrence of the constant magnitude could be attributed to the fully developed plume behaviour when the air bubbles spread evenly across the entire column cross-section at a higher column height.

However, a noticeable difference could be observed around the membrane at 0.3 m. Comparing the block and fibre models, the velocities were much higher with the fibre model, which has a strong implication on the fouling analysis. The fibre model showed an air axial velocity magnitude of 0.35 m/s at the centre of the membrane gap which then decreased sharply close to 0 m/s at the membrane fibre surface. It sharply increased again beyond the membrane bundle at $r/R=0.07$ and gradually increased to reach its peak of 0.48 m/s at $r/R=0.35$. From $r/R=0.8-1.0$, a steep decline of the air axial velocity occurred. Comparatively, the block model showed a much simpler trend. The air axial velocity increased sharply from 0 m/s at the block surface to peak at $r/R=0.25$ at 0.5 m/s and gradually decreased until $r/R=0.9$ and then a sharp reduction to 0 m/s at the column wall. The air axial velocity of the fibre model reached and remained at 0 m/s at a distance further away from the column wall in contrast to the block wall due to the refined configuration of the membrane module.

Overall, the sharp decrease in the air axial velocity was due to the submersion of the membrane module. After refinement of the membrane from a solid block to individual tubes, the range of steep decrease near the centreline was reduced. The improvement of the membrane structure should better depict the flow near the fiber edge. Within the free flow region, the two models delivered similar results which indicated that in these areas, the hydrodynamics was not much affected by the membrane configuration.

Figure 5.16 Air axial velocity for the block and fiber models at the 0.3 m membrane position at a 1 L/min air flow rate along the x-axis at (a) 0.9 m, (b) 0.6 m and (c) 0.3 m column heights

Figure 5.17 shows the radial profile of the air axial velocity at $z =$ (a) 0.9 m, (b) 0.6 m and (c) 0.3 m for the block and fibre models. The membrane module was simulated at the 0.6 m location at the air flow rate of 1.0 L/min. Similar to Figure 5.16, both models displayed similar trends, i.e., a decreasing air axial velocity with increasing column heights. At the 0.9 m location, the velocity displayed a trapezoidal data trend whereby it was approximately constant from the column centre to $r/R=0.8$ and then decreased steeply to reach 0 m/s at the column wall.. Around the membrane module at the 0.6 m location, a significant difference between the two models could be clearly observed. The fiber model showed an air axial velocity magnitude of 0.35 m/s inside the membrane gap at the centre which decreased sharply to 0 m/s at the membrane fiber surface. The air axial velocity decreased again at $r/R=0.15$ and became gently increasing to reach a peak of 0.46 m/s at $r/R=0.2$. After that, it reduced gradually and decreased to 0 m/s close to the column wall. Comparatively, with the block model the velocity sharply increased from 0 m/s at $r/R=0.15$ to a peak of 0.45 m/s at $r/R=0.2$. It then diminished to 0 m/s at the column wall. At the 0.3 m location, both models showed a peak value at the column centre which gradually decreased to 0.35 m/s at $r/R=0.8$ and then sharply decreased to 0 m/s at the column wall. In terms of the magnitude, the fibre model had a significantly greater peak value of 0.6 m/s compared to 0.52 m/s determined using the block model due to the reduced blockage.

Overall, both figures showed a sharp decrease and significantly low velocity values near the membrane edge and column wall for all column heights which could be attributed to a boundary layer effect. Besides, the non-slip condition set at the wall led to the zero velocity of bubbles along the column wall. In reality, the bubble velocity was not zero. Particularly, near the column wall, the low air axial velocity might be because that the bubbles near the wall were slowed down by the drag of the downward moving liquid which was the recirculation. On top of this, the air axial velocity predicted by the fiber model was shown to have a higher magnitude compared with that derived from the block model, which was consistent with Figure 5.16.

Figure 5.17 Air axial velocity for both models at the 0.6 m membrane position at 1 L/min air flow rate along the x-axis at (a) 0.9 m, (b) 0.6 m and (c) 0.3 m column

5.6.2.2 Air Volume Fraction

Figure 5.18 illustrates the contours of the air volume fraction using the two approaches at the two measuring locations, respectively. For both the block and fiber models, the air volume fraction was highest at the injection point and decreased gradually along the column height. Due to the entrainment of the liquid phase, the air volume fraction also reduced radially. In terms of the magnitude, the fiber model showed higher air volume fraction value comparing with the block model for both locations.

Figure 5.18 Contours of the air volume fraction for the two models: (a) 0.3 m block, (b) 0.3 m fiber, (c) 0.6 m block and (d) 0.6 m fiber

Figure 5.19 shows the variations of the mean air volume fraction with the air flow rate for the two models. It can be seen that the air volume fraction generally

increased with the air flow rate, which was consistent with the previous findings (Deckwer et al. 1980; Schumpe and Grund 1986; Saxena et al. 1990; Daly et al. 1992; Pino et al. 1992; Hyndman et al. 1997; Krishna et al. 1997; Li and Prakash 2000; Prakash et al. 2001). Both the block and fibre models had higher air volume fractions at the lower column height of 0.3 m location. Comparatively, the fibre model predicted higher magnitudes of the air volume fraction at both locations.

Figure 5.19 Air volume fraction against the air flow rate

Figure 5.20 shows the radial distribution of the air volume fraction at different column heights. The membrane module was at the 0.3 m location and the air flow rate was 1.0 L/min. At the 0.3 m location, the two models exhibited markedly different trends especially near the membrane edge. The air volume fraction of the fibre model had a peak of 0.025 at $r/R=0.27$ whereas the block model had a lower peak of 0.019 at $r/R=0.18$. This indicated that the block membrane had many

bubbles gathered right at the edge of the module whereas the fiber module had higher bubble concentrations of air bubbles at a distance away from the fibre edge. This could be due to the increased recirculation of the liquid medium through and around the fibers.

At the 0.6 and 0.9 m heights, a similar flat data trend which was relatively constant across the column radius was obtained. Furthermore, the values of the air volume fraction decreased from 0.6 to 0.9 m. This was due to the fully developed bubble plume at a higher column height which caused air bubbles to distribute uniformly across the cross-section of the column. Moreover, the smaller value of the air volume fraction at the 0.9 m location could also be a result of the energy dissipation near the membrane surface.

Figure 5.21 shows the radial distribution of air volume fraction at $z =$ (a) 0.9 m, (b) 0.6 m and (c) 0.3 m. The membrane module was at the 0.6 m membrane position and the air flow rate was 1.0 L/min. Similar to Figure 5.20, both models showed a higher air volume fraction at a lower column height. The fibre model had a relatively higher value of the air volume fraction compared to the block model. At the 0.9 m height, similar flat data trend which was relatively constant across the column radius was found for both models. With the membrane module at the 0.6 m location, the air volume fraction of the fibre model had a peak value of 0.009 at $r/R=0.23$ whereas the block model had a lower peak of 0.006 at $r/R=0.15$. At the 0.3 m position, both models illustrated a similar data trend with a peak value at the column centre that decreased gradually to 0 at the column wall.

Overall, the radial profiles of both models at the 0.3 m membrane position showed higher magnitudes of the air volume fraction compared to the 0.6 m position. This was consistent with the analysis done earlier across the range of air flow rates. The air volume fraction profiles were also distinctly different between the block and fibre models, which indicated the importance incorporating the membrane configuration design in the simulations for fouling analysis.

Figure 5.20 Air volume fraction for the two models at the 0.3 m membrane position at a 1 L/min air flow rate along the x-axis at (a) 0.9 m, (b) 0.6 m and (c) 0.3 m column heights

Figure 5.21 Air volume fraction for the two models at the 0.6 m membrane position at a 1 L/min air flow rate along the x-axis at (a) 0.9 m (b) 0.6 m and (c) 0.3 m column heights

5.6.2.3 Liquid Axial Velocity

Figure 5.22 shows the variation of the liquid axial velocity at the six air flow rates for the block and fiber models at the two respective locations. Both the block and fibre models had general liquid axial velocities of 10^{-2} m/s. The significantly lower value of the liquid axial velocity compared with the air axial velocity could be attributed to the fact that the liquid velocity was induced by the rising air bubbles (Gan *et al.* 2011). The liquid axial velocity was smaller at the higher column region of 0.6 m for both models. Moreover, the fibre model showed a higher magnitude of the liquid axial velocity at both membrane positions, which could be directly related to the plume effect whereby a higher air volume fraction induced a higher liquid velocity.

Figure 5.22 Liquid axial velocity versus air flow rate

Figure 5.23 shows the radial distribution of the liquid axial velocity at $z =$ (a) 0.9 m, (b) 0.6 m and (c) 0.3 m. The liquid axial velocity generally decreased with increasing column height. The development of the radial velocity was similar at the

0.6 and 0.9 m locations. At higher column heights, a peak value occurred at the centreline and the liquid flow reversed its direction at $r/R = 0.78$, which implied the formation of liquid circulation at the measuring regions.

At the 0.3 m location, the trends were substantially different. For the block model, the highest liquid axial velocity was found at the membrane edge and then the value decreased continuously and eventually reached close to 0 m/s at the column wall. When the fiber model was considered, the liquid axial velocity achieved a maximum at a distance from the membrane edge, which was at around $r/R = 0.3$. From $r/R = 0$ to 0.7, the liquid axial velocity decreased progressively. From $r/R = 0.7$ to 0.9, the liquid velocity increased gradually and then dropped to zero at the boundary wall due to the non-slip condition set earlier. The increase of liquid axial velocity from $r/R = 0.7$ to 0.9 might be attributable to the refined structure of the individual fibers. Thus, the bubble plume was separated by the membrane holder, with part of the plume moving upward near the fiber bundle and the rest developing near the column wall, which led to a low liquid velocity region in-between.

Figure 5.24 shows the radial profile of the liquid axial velocity at $z =$ (a) 0.9 m, (b) 0.6 m and (c) 0.3 m for both models at the 0.6 m membrane position. At the 0.9 m location, the development of the liquid axial velocity was similar to Figure 5.23 (a). With the submersion of the membrane module at the 0.6 m location, the liquid axial velocity achieved a peak value of 0.073 m/s at $r/R = 0.23$ and decreased to 0.027 m/s at $r/R = 0.95$ for the fiber models. For the block model, the liquid axial velocity reached a maximum right at the membrane edge and reversed its direction at $r/R = 0.85$. For all three axial locations, the flow direction reversed near wall and approached 0 m/s at the column wall. Due to the air injection and liquid entrainment at the 0.3 m location, the liquid axial velocities were the largest in the meandering region in comparison with higher axial elevations.

Figure 5.23 Liquid axial velocity predicted by the models at the 0.3 m membrane position at 1 L/min air flow rate along the x-axis at (a) 0.9 m (b) 0.6 m and (c) 0.3 m column heights

Figure 5.24 Liquid axial velocity predicted by the models at 0.6 m membrane position at a 1 L/min air flow rate along the x-axis at (a) 0.9 m (b) 0.6 m and (c) 0.3 m column heights

5.6.2.4 Turbulent Kinetic Energy (TKE)

Figure 5.25 shows the development of the average TKE as a function of the air flow rate for the block and fiber models at the 0.3 and 0.6 m locations. For the block model, the TKE generally increased with the air flow rate, which indicated that the level of turbulence inside the BCR was enhanced. For the fiber model, the rate of increase of the TKE was slower at the 0.3 m location and then flattened out when the air flow rate exceeded 1.0 L/min for both locations. For the smaller air volume fraction, the interactions between the air bubbles and liquid medium became more intense with increasing air flow rate (Joshi 2011). Hence, better mixing was achieved owing to enhanced agitation induced by the air bubbles. When the membrane geometry was refined to account for the effect of the individual fibers, the increase of the TKE was slowed down and eventually achieved a maximum at the 0.6 m location. Since the rate of TKE is the key parameter for turbulence characterization prediction that TKE approached constant beyond an optimal air flow rate indicated that the turbulence achieved a maximum saturation state. In addition, higher TKE values at the lower column height suggested that the chaotic instability of the flow was stronger in the region near the air sparger.

Figure 5.25 TKE against the air flow rate for the block and fiber models

Figure 5.26 shows the radial distribution of the TKE at different column heights. The membrane module was simulated at the 0.3 m location at an the air flow rate of 1.0 L/min. Generally, the highest TKE values appeared at 0.6 m, followed by the lower column height of 0.3 m and lastly of 0.9 m. The smaller TKE values near the water surface might be attributed to the bubbles being dispersed over the entire cross-section of the BCR, yielding a locally smaller momentum transfer from the bubbles to the liquid medium. This observation was in agreement with previous studies for small bubbles (Sommerfeld and Bröder 2009).

It can be seen that the fibre model had greater values of the TKE except at the 0.3 m location. At 0.9 m, both models predicted relatively constant TKE values across the column radius with the fibre model at a higher value of $5 \times 10^{-4} \text{ m}^2/\text{s}^2$ and the block model at a value of $6 \times 10^{-5} \text{ m}^2/\text{s}^2$. At the 0.6 m location, the two models predicted a similar data trend, remaining constant from the column centre to $r/R = 0.5$ and then decreasing to $0 \text{ m}^2/\text{s}^2$ at the wall. Meanwhile, two peak values were identified for both models, with a higher maximum value of $2.2 \times 10^{-3} \text{ m}^2/\text{s}^2$ for the fibre model and $1.7 \times 10^{-3} \text{ m}^2/\text{s}^2$ for the block model. At the 0.3 m location, both models showed a similar data trend, gradually increasing from the column centre to reach a peak at $r/R = 0.5$. The fibre model predicted a minimum value of $6 \times 10^{-5} \text{ m}^2/\text{s}^2$ and $9 \times 10^{-4} \text{ m}^2/\text{s}^2$ at the column centre, whereas the block model predicted TKE values of $5 \times 10^{-4} \text{ m}^2/\text{s}^2$ and $1.9 \times 10^{-3} \text{ m}^2/\text{s}^2$ at $r/R = 0.15$ where the membrane module edge located. This different trend could be attributed to the different membrane configurations for these two models. Apart from this, higher data fluctuations were also observed for the fibre model as opposed to the block model.

Figure 5.27 shows the radial distribution of the TKE at $z =$ (a) 0.9 m, (b) 0.6 m and (c) 0.3 m for both models at the 0.6 m membrane position. Similar to Figure 5.26, higher TKE values were obtained for the fibre model at the three axial locations. It also shows that the TKE values were smaller at the higher column regions. At 0.9 m, both models showed relatively constant TKE values across the column radius, with the fibre model at a higher magnitude of around $7 \times 10^{-4} \text{ m}^2/\text{s}^2$ and the block model around $2 \times 10^{-5} \text{ m}^2/\text{s}^2$. Furthermore, the difference between the TKE trends was not

as significant at the 0.6 m compared to 0.3 m. Both models demonstrated a similar data trend which gradually increased from the column centre and reached a peak value at $r/R=0.5$. From $r/R=0.5$ to 1.0, the TKE values decreased gradually to zero. In addition, the fibre model had a minimum value of $3 \times 10^{-5} \text{ m}^2/\text{s}^2$ at the column centre and a higher maximum value of $8 \times 10^{-4} \text{ m}^2/\text{s}^2$ at $r/R=0.5$. For the block model, a minimum TKE value of $2 \times 10^{-4} \text{ m}^2/\text{s}^2$ at $r/R=0.15$ and a maximum value of $5 \times 10^{-4} \text{ m}^2/\text{s}^2$ could be identified.

At the 0.3 m height, both models showed a similar data trend. The TKE values increased from $3.3 \times 10^{-3} \text{ m}^2/\text{s}^2$ at the column centre and then reached a peak value at $r/R=0.33$. After that, they gradually decreased to $7 \times 10^{-4} \text{ m}^2/\text{s}^2$ at the column wall. The fibre model had a higher maximum value of $4.4 \times 10^{-3} \text{ m}^2/\text{s}^2$, while for the block model, the maximum TKE value was $3.9 \times 10^{-3} \text{ m}^2/\text{s}^2$. The higher TKE values near the centre were due to the air plume effect with higher air velocities along the centreline leading to a higher level of turbulence.

Figure 5.26 The TKE for the two models at the 0.3 m membrane position at a 1 L/min air flow rate along the x-axis at (a) 0.9 m, (b) 0.6 m and (b) 0.3 m column heights

Figure 5.27 The TKE for the two models at the 0.6 m membrane position at a 1 L/min air flow rate along the x-axis at (a) 0.9 m, (b) 0.6 m and (c) 0.3 m column heights

5.7 Data Validation using PIV

PIV is a whole field non-intrusive measurement technique that enables examination of spatial flow structures in both steady and unsteady flows. It has been used in a wide range of applications to study flow characteristics. With PIV two-dimensional velocity vectors are captured in a plane in the flow field. This technique is unlike the intrusive methods which require a probe be placed where the velocities are to be measured. The presence of probes is likely to affect the flow within the boundary layer which will further influence the accuracy of data readings especially when small magnitude velocities are to be measured (Yeo *et al.* 2006).

The general principle of PIV is to illuminate the tracer particles in a flow field of interest with a plane sheet of light, and then obtain two images of the diffracted light pattern of the particles with a known time separation. The displacement of the particles is calculated by cross-correlating the two images and with the known time interval leading to determination of the velocity vector field. The size of the interrogation area should also avoid velocity variation within the interrogation area itself to reduce errors.

In this study average air axial velocity data of BCR with the membrane submerged at either 0.3 or 0.6 m under different air flow rates were measured accordingly and compared with the numerical results for validation. The bubbles themselves were used as the seeding particles which could be illuminated by the laser sheet. Hence, the velocities determined were the air-phase velocities. The frame rate was 20 per second and a total duration of 30 s was captured after the system was stabilized. In this regard, for a particular measuring point, a total of 600 air axial velocities were collected. After signal filtering to remove outliers, the adjusted air axial velocities were averaged at the 0.3 and 0.6 m locations to obtain the radial distribution profiles.

Figure 5.28 illustrates the radial air axial velocity profile for the block and fiber models and PIV data with the membrane submerged at 0.3 m at the air flow rate of (a) 0.2 and (b) 0.5 L/min respectively. Generally, the prediction of the two models was of the same order-of-magnitude. For Figure 5.28 (a) the fiber model showed a

better agreement with the PIV data at the membrane edge and the near wall region compared to the block model. However, both models underestimated the air velocity in the free stream field ($r/R=0.3-0.7$). When the air flow rate was increased to 0.5 L/min (see Figure 5.28 (b)), both models showed a similar trend of the data but overestimated the air velocity 26.86% as the maximum air velocity was around 0.38 m/s for the two models while the peak air velocity was 0.28 m/s measured by PIV.

Figure 5.29 shows the air axial velocity profile for the block and fiber models and PIV measurements with the membrane submerged at 0.6 m location at the two air flow rates of (a) 0.2 and (b) 0.5 L/min, respectively. Similar to Figure 5.28 (a), the PIV data had a more significant peak of 0.21 m/s at $r/R=0.4$. It also exhibited a wider peak as compared to the simulated data using the two models. It was found that the air velocity predicted by the fiber model fit well within the fibers with the data measured by PIV. It demonstrated the superiority of the fiber model in terms of depicting the flow conditions within the individual fibers, which is critical for fouling analysis. Generally, the predictions of the two models were of the same order-of-magnitude. For the 0.5 L/min air flow rate, the trend was also similar to that shown in Figure 5.28 (b), the block and fibre models both overestimated the air axial velocity by about 30% when the peak values were compared.

Overall, it can be concluded that the fibre model was more accurate than the block model in predicting the flow conditions in the area surrounding the membrane module. This is significant and has important implication for fouling alleviation analysis. The results validate refining the membrane geometry in order to improve the accuracy of the numerical solutions.

(a)

(b)

Figure 5.28 Air axial velocity profile for block, fiber and PIV data with the membrane submerged at the 0.3 m location at (a) 0.2 and (b) 0.5 L/min air flow rates

(a)

(b)

Figure 5.29 Air axial velocity profile for block, fiber and PIV data with the membrane submerged at the 0.6 m location at (a) 0.2 and (b) 0.5 L/min air flow rates

5.8 Comparisons among Different Approaches

In this study, three different approaches, namely, the membrane filtration tests, dynamic pressure sensing measurements and CFD were applied to investigate the hydrodynamic characteristics of a laboratory scale BCR with and without a submerged hollow fiber membrane module. The three methods are introduced in Chapters Three, Four and Five, respectively. The major objective of this chapter is to corroborate the results from them and propose an integrated method for fast characterization of the hydrodynamics.

First of all, membrane filtration tests were carried out under constant flux conditions. The primary finding was that there existed an optimal air flow rate of around 1.2 L/min for the air sparging for a specific membrane filtration system (refer to Figure 3.11). Beyond this optimal value, further increases in the air flow rate had no benefit on fouling mitigation. Since aeration requires a significant part of the total energy consumption in multiphase reactors such as MBRs and gas-liquid FBRs, the quick determination of this optimal air flow rate is essential to lower the energy consumption and optimize the system operation.

Although the optimal air flow rate could be identified via the membrane filtration experiments, the trial-and-error processes are rather time-consuming since the detailed flow behaviour remains unknown. In order to determine the optimal air flow rate more efficiently, the quantification of the hydrodynamics inside the reactors through measurements is important. However, the complexity of the multiphase flow behaviour and the aggressive environment inside the multiphase reactors lead to the difficulties in the application of most measurement techniques. In this study a submersible pressure sensor was used to characterize the flow dynamics at different positions inside the BCR. The hydrodynamic information was extracted from the transient dynamic pressure signals. Data analysis was then carried out to convert the hydrodynamic measurements into nonlinear parameters that described the instantaneous flow dynamics.

Figure 5.30 and Figure 5.31 show the development of the Hurst exponent and *LLE* along with the membrane fouling rate respectively, at different air flow rates with the membrane module submerged at the 0.3 m location. It can be seen that with an the increase in the air flow rate, the membrane fouling rate decreased continuously and reached a plateau at around 1.2 L/min. Meanwhile, the *H* value increased progressively and also flattened out from 1.2 L/min onwards. Furthermore, the *LLE* also approached a constant at around the same air flow rate. Hence, it is suggested that the development of *H* and *LLE* values correlates with the membrane filtration performance and can be used to quantify the flow fluctuations inside the BCR. Since the Hurst exponent is normally used to characterize the long-term correlation of the pressure signals, the flattened tail of the *H* curve here indicated the flow behaviour was deemed stabilized beyond the optimal air flow rate. Since the *LLE* is used to indicate the predictability of a chaotic system, similar to the Hurst exponent, the *LLE* approaching a constant implied the chaotic instability of the system was not further enhanced beyond the optimal air flow. Overall, by plotting the *H* or *LLE* against the air flow rate, the optimal air flow rate could be identified within a much shorter time frame compared with the trial-and-error experiments.

Figure 5.30 Development of the Hurst exponent and the fouling rate as a function of the air flow rate at the 0.3 m location

Figure 5.31 Development of the *LLE* and the fouling rate as a function of the air flow rate at the 0.3 m location

Beside the dynamic pressure sensing measurements, 3D CFD models were also developed to quantify the flow behaviour inside the BCR. Relative to the dynamic pressure sensing technique which focuses on the local flow hydrodynamics characterization, numerical simulation is more powerful in terms of predicting the complex flow behaviour over the entire domain of interest. At the same time, the approximation due to turbulence modelling and grid discretization should be noted. By employing this approach, the transient and time-averaged flow field can be obtained. Moreover, many hydrodynamic features like gas and liquid velocity, gas fraction as well turbulent kinetic energy and energy dissipation rate can be computed. In this regard, the hydrodynamics can be re-tuned to achieve maximal turbulence fluctuations around the membrane modules for effective fouling control.

Figure 5.32 shows the development of the air axial velocity and the membrane fouling rate at various air flow rates when the membrane module is submerged at the 0.3 m location. It can be observed that the optimal air flow rate occurred where the air axial velocity no longer increased in proportion to with the air flow rate.

Figure 5.32 Development of the air axial velocity for the fiber model and the fouling rate as a function of the air flow rate at the 0.3 m location

Figure 5.33 shows the development of the TKE and the membrane fouling rate at various air flow rates with the membrane module submerged at the 0.3 m location. Basically, the optimal air flow rate also occurred when the increasing rate of the TKE slowed down.

Since both the air axial velocity and TKE are important parameters to characterize the hydrodynamics. The plateau appearance for both quantities near the optimal air flow rate indicated that the level of turbulence was not further enhanced with an increase in the air flow rate. Thus, the CFD simulations reinforced the earlier conclusions. In addition, by conducting the numerical simulation, the mixing conditions were clearly displayed which was essential for system optimization. The region that had little or no flow circulation indicated by the small magnitude of parameters such as the air axial velocity and TKE, should be re-tuned to maximize the flow fluctuations.

Figure 5.33 Development of the TKE for the fiber model and the fouling rate against the air flow rate at the 0.3 m location

As a result, for a particular MBR system, the novel dynamic pressure sensing measurement can be carried out at desired positions with a multiphase pressure sensor installed. Nonlinear parameters can be computed via various data analysis methods and plotted against the air flow rate. In this regard, the magnitude of the optimal air flow rate could be determined efficiently. In addition, the CFD simulation also provides a cost-effective tool to predict the flow hydrodynamics of the multiphase reactors. The determination of the optimal air flow rate can be achieved by identifying the change of the air axial velocity and TKE trend near the measured region of interest. The successful application of these two approaches would allow operators to optimize the air sparging conditions within a short time, which is essential for energy reduction and cost minimization.

5.9 Conclusions

This chapter examined the hydrodynamic characteristics of a BCR with and without a submerged hollow fiber membrane module using CFD. The numerical model was first established for the BCR without the submersion of the membrane module. It was proven that this model was able to predict the instantaneous and time-averaged multiphase flow behaviour and that the realizable $k-\varepsilon$ turbulence closure was superior in predicting the time-averaged flow behaviour while the LES was more capable of investigating the unsteady flows.

Later, the BCR geometry was re-constructed to include the hollow fiber membrane module. Two membrane configurations were considered, namely, the solid block and individual tubes. The numerical results were analysed and assessed to quantify the flow dynamics around the membrane module. Generally, the numerical model was able to provide supplementary information on the flow hydrodynamics. On the other hand, the refinement of the membrane structure from a solid block to individual tubes better predicted the flow behaviour around the membrane module as the flow dynamics within and around the fibers was included. It was found that the air flow rate was the primary parameter that determined the level of flow instability inside the BCR. Furthermore, the plot of the TKE versus the air flow rates computed using the fiber model was shown to level off beyond a critical air flow rate. Thus, with the proper construction of the membrane configuration, the turbulence inside the BCR was shown to achieve a saturation maximum at the optimal air flow rate.

6. Conclusions and Recommendations

6.1 Conclusions

In the present study, the membrane filtration tests that were carried out confirmed the existence of an optimal superficial air velocity for the air sparging of a membrane module. Factors including the aeration strength (in terms of air flow rate), bubble-induced turbulent flow, air sparger geometry, membrane module orientation as well as the nature of the feed solutions were examined. Compared with the central aeration scheme, the alternate aeration produced by the two air stones with 1.0 min switching interval greatly reduced the aeration demand. In addition, the testing with both inorganic and organic feeds for membrane filtration demonstrated that the particle composition as well as particle size distribution all contributed to the cake formation on the membrane surface.

The feasibility of the novel dynamic pressure sensing technique to quantify the local flow dynamics with and without the submersion of the hollow fiber membrane module was demonstrated. This robust and inexpensive technique is also potentially useful for industrial applications. The nonlinear parameters, of LLE and H , were found to be able to quantify the local hydrodynamics and could be used to determine the optimal air flow rate for a particular membrane filtration system within a much shorter time frame compared with the trial and error filtration tests.

The application of the CFD approach was also confirmed to be a powerful tool for hydrodynamics prediction in multiphase reactors. The velocity of the respective phase, air volume fraction and the TKE were investigated using realisable $k-\epsilon$ turbulence closure and the LES model. In this regard, not only the local flow behaviour was predicted, the instantaneous and time-averaged flow dynamics were

described over the whole reactor. By carefully simulating the membrane module as twenty individual solid tubes, the flow within and around the fibers was better predicted. Therefore, the fouling mechanisms could be assessed in more detail and the optimal air flow rate could be determined using this numerical modelling.

The integrated research using the three approaches in this study enabled the in-depth understanding of the hydrodynamic characteristics of multiphase reactors which was rarely reported in the past. The knowledge gained in this research should be useful for the design of air sparger and system operation optimization in multiphase reactors.

6.2 Recommendations

Based on the results obtained and challenges that have been encountered during this study, the following suggestions are proposed:

1. Constant flux conditions should be monitored all the time when conducting the membrane filtration tests irrespective of the automation system used. The loss of flexibility of the tygon tubing could cause the gradual reduction of permeate flux.
2. The noise level should be minimized when conducting the dynamic pressure sensing measurements. The overly high level of redundant signals could lead to the removal of real signals when signal filtering was conducted.
3. Multiple sensors should be applied when large scale multiphase reactors are tested. The assessment of the local flow dynamics using multiple sensors simultaneously could help determine the optimal air flow rate faster and more precisely.
4. Further improvement of the fiber bundle configuration should be conducted to better predict the flow between the fibers. Fiber looseness could also be considered to study the flow behaviour around the individual fibers.

5. The meshing should be refined around the fiber tubes and right above the air inlet to account for the boundary effect. The convergence requirement must be met to guarantee the accuracy of the numerical solutions.
6. The conclusions drawn from the present research shall be tested on pilot-and plant-scale MBRS. When scaling-up, the optimal operating conditions including the optimal air flow rate and air sparger arrangement shall be re-determined to meet the specific requirements of the treatment system. The methods and algorithms employed in this study can be utilized for system optimization.
7. The numerical solutions derived from the 3D model were found to be unsatisfactory. The inaccuracy of the results might be attributable to the improper description of the interfacial forces, namely, the lift, virtual mass and drag forces, in the numerical model. The PIV measurements also showed substantially scatterings. Due to time limitations in this study, direct PIV measurements were taken with the cylindrical column, and effort was attempted to compensate for the refraction of the laser light due to the circular walls in the analysis stage. However, this approach was proven to be unsatisfactory and the refraction effects had introduced uncertainties to the measurements. Besides, the laser light was also scattered by the bubbles inside the reactor which affected the data accuracy, particularly when the air fraction was high. For future work, it is therefore suggested that the three interfacial forces shall be included for further numerical simulations. In addition, the PIV measurements should be improved to include an observation box with plane glass windows to eliminate the refraction effects, as well as to explore the use of fluorescent particles to eliminate the scattered laser light on the images due to the rising bubbles.

References

- Akita, K. and F. Yoshida (1974). "Bubble Size, Interfacial Area, and Liquid-Phase Mass Transfer Coefficient in Bubble Columns." Industrial & Engineering Chemistry Process Design and Development **13**(1): 84-91.
- Al-Dahhan, M. (2002). Flow Visualization of multiphase flow systems. The 6th Engineering Conference. KFUPM. Dhahran. **5**: 661-668.
- Al-Masry, W. A., E. M. Ali and Y. M. Aqeel (2005). "Determination of Bubble Characteristics in Bubble Columns Using Statistical Analysis of Acoustic Sound Measurements." Chemical Engineering Research and Design **83**(10): 1196-1207.
- Albasi, C., S. Felipe and S. C. S. Rocha (2004). "Time series analysis of pressure fluctuation in gas-solid fluidized beds." Brazilian Journal of Chemical Engineering **21**: 497-507.
- Bakshi, B. R., H. Zhong, P. Jiang and L. S. Fan (1995). "Analysis of flow in gas-liquid bubble columns using multi-resolution methods." Chemical Engineering Research and Design **73**: 608-614.
- Becker, S., A. Sokolichin and G. Eigenberger (1994). "Gas—liquid flow in bubble columns and loop reactors: Part II. Comparison of detailed experiments and flow simulations." Chemical Engineering Science **49**(24, Part 2): 5747-5762.
- Behkish, A., Z. Men, R. J. Inga and B. I. Morsi (2002). "Mass transfer characteristics in a large-scale slurry bubble column reactor with organic liquid mixtures." CHem Eng Sci **57**(3307-3324).

- Bella, G. D., F. Durante, M. Torregrossa, G. Viviani, P. Mercurio and A. Cicala (2007). "The role of fouling mechanisms in a membrane bioreactor." Water Science & Technology **55**(8-9): 455-464.
- Bellara, S. R., Z. F. Cui and D. S. Pepper (1996). "Gas sparging to enhance permeate flux in ultrafiltration using hollow fibre membranes." Journal of Membrane Science **121**(2): 175-184.
- Benna, M., N. Kbir-Ariguib, C. Clinard and F. Bergaya (2001). "Static filtration of purified sodium bentonite clay suspensions. Effect of clay content." Applied Clay Science **19**(1-6): 103-120.
- Benna, M., N. Kbir-Ariguib, A. Magnin and F. Bergaya (1999). "Effect of pH on Rheological Properties of Purified Sodium Bentonite Suspensions." Journal of Colloid and Interface Science **218**(2): 442-455.
- Bennett, M. A., R. M. West, S. P. Luke, X. Jia and R. A. Williams (1999). "Measurement and analysis of flows in a gas-liquid column reactor." Chemical Engineering Science **54**(21): 5003-5012.
- Bhole, M. R., S. Roy and J. B. Joshi (2006). "Laser Doppler Anemometer Measurements in Bubble Column: Effect of Sparger." Industrial & Engineering Chemistry Research **45**(26): 9201-9207.
- Bi, H. T. and J. R. Grace (1995). "Flow regime diagrams for gas-solid fluidization and upward transport." International Journal of Multiphase Flow **21**(6): 1229-1236.
- Böhm, L., A. Drews, H. Prieske, P. R. Bérubé and M. Kraume (2012). "The importance of fluid dynamics for MBR fouling mitigation." Bioresource Technology **122**(0): 50-61.

- Borchers, O., C. Busch, A. Sokolichin and G. Eigenberger (1999). "Applicability of the standard $k-\epsilon$ turbulence model to the dynamic simulation of bubble columns. Part II:: Comparison of detailed experiments and flow simulations." Chemical Engineering Science **54**(24): 5927-5935.
- Bouaifi, M., G. Hebrard, D. Bastoul and M. Roustan (2001). "A comparative study of gas hold-up, bubble size, interfacial area and mass transfer coefficients in stirred gas-liquid reactors and bubble columns." Chemical Engineering and Processing: Process Intensification **40**(2): 97-111.
- Braak, E., M. Alliet, S. Schetrite and C. Albasi (2011). "Aeration and hydrodynamics in submerged membrane bioreactors." Journal of Membrane Science **379**(1-2): 1-18.
- Brannock, M., Y. Wang and G. Leslie (2010). "Mixing characterisation of full-scale membrane bioreactors: CFD modelling with experimental validation." Water Research **44**(10): 3181-3191.
- Briens, C. L., L. A. Briens, J. Hay, C. Hudson and A. Margaritis (1997). "Hurst's analysis to detect minimum fluidization and gas maldistribution in fluidized beds." AIChE Journal **43**(7): 1904-1908.
- Brookes, S. J., E Reid, E Germain, S Smith, H Alvarez, P LeClech, T Stephenson, E Turra, B. Jefferson (2003). "Biomass characterisation in membrane bioreactors." Proceedings of the IMSTEC, Sydney, Australia.
- Buer, T. and J. Cumin (2010). "MBR module design and operation." Desalination **250**(3): 1073-1077.
- Cabassud, C., S. Laborie, L. Durand-Bourlier and J. M. Lainé (2001). "Air sparging in ultrafiltration hollow fibers: relationship between flux enhancement, cake

- characteristics and hydrodynamic parameters." Journal of Membrane Science **181**(1): 57-69.
- Cassanello, M., F. Larachi, A. Kemoun, M. H. Al-Dahhan and M. P. Dudukovic (2001). "Inferring liquid chaotic dynamics in bubble columns using CARPT." Chemical Engineering Science **56**(21–22): 6125-6134.
- Chan, C. C. V., P. R. Bérubé and E. R. Hall (2007). "Shear profiles inside gas sparged submerged hollow fiber membrane modules." Journal of Membrane Science **297**(1–2): 104-120.
- Chan, C. C. V., P. R. Bérubé and E. R. Hall (2011). "Relationship between types of surface shear stress profiles and membrane fouling." Water Research **45**(19): 6403-6416.
- Chang, I.-S. and C.-H. Lee (1998). "Membrane filtration characteristics in membrane-coupled activated sludge system -- the effect of physiological states of activated sludge on membrane fouling." Desalination **120**(3): 221-233.
- Chang, I. S. and S. N. Kim (2005). Wastewater treatment using membrane filtration: effect of biosolids concentration on cake resistance. Kidlington, ROYAUME-UNI, Elsevier.
- Chang, I. S., Pierre Le Clech, Bruce Jefferson, and Simon Judd (2002). "Membrane Fouling in Membrane Bioreactors for Wastewater Treatment " J. Envir. Engrg. ASCE **128**: 1018-1029.
- Chang, S., A. G. Fane and S. Vigneswaran (2002). "Modeling and optimizing submerged hollow fiber membrane modules." AIChE Journal **48**(10): 2203-2212.

- Chen, R. C., J. Reese and L. S. Fan (1994). "Flow structure in a three-dimensional bubble column and three-phase fluidized bed." AICHE Journal **40**(7): 1093-1104.
- Chen, W., T. Hasegawa, A. Tsutsumi, K. Otawara and Y. Shigaki (2003). "Generalized dynamic modeling of local heat transfer in bubble columns." Chem Eng J **96**: 37-44.
- Choi, H., K. Zhang, D. D. Dionysiou, D. B. Oerther and G. A. Sorial (2005). "Effect of permeate flux and tangential flow on membrane fouling for wastewater treatment." Separation and Purification Technology **45**(1): 68-78.
- Choi, H., K. Zhang, D. D. Dionysiou, D. B. Oerther and G. A. Sorial (2005). "Influence of cross-flow velocity on membrane performance during filtration of biological suspension." Journal of Membrane Science **248**(1-2): 189-199.
- Cimbala, J. M. (2008). "Fluent and Gambit Learning Modules." from http://www.mne.psu.edu/cimbala/Learning/Fluent/gambit_introduction.htm.
- Cui, Z. F., S. Chang and A. G. Fane (2003). "The use of gas bubbling to enhance membrane processes." Journal of Membrane Science **221**(1-2): 1-35.
- Çulfaz, P. Z., M. Wessling and R. G. H. Lammertink (2011). "Fouling behavior of microstructured hollow fiber membranes in submerged and aerated filtrations." Water Research **45**(4): 1865-1871.
- Daly, J. G., S. A. Patel and D. B. Bukur (1992). "Measurement of gas holdups and sauter mean bubble diameters in bubble column reactors by dynamics gas disengagement method." Chemical Engineering Science **47**(13-14): 3647-3654.

- de Martín, L., J. V. Briongos, N. García-Hernando and J. M. Aragón (2011). "Detecting regime transitions in gas-solid fluidized beds from low frequency accelerometry signals." Powder Technology **207**(1–3): 104-112.
- Deckwer, W.-D., Y. Louisi, A. Zaidi and M. Ralek (1980). "Hydrodynamic Properties of the Fischer-Tropsch Slurry Process." Industrial & Engineering Chemistry Process Design and Development **19**(4): 699-708.
- Deen, N. G., T. Solberg and B. H. Hjertager (2000). Numerical Simulation of the Gas-Liquid Flow in a Square Crosssectioned Bubble Column CHISA 14th Interantional Congress of Chemical and Process Engineering. Praha, Czech Republic.
- Deen, N. G., T. Solberg and B. H. Hjertager (2001). "Large eddy simulation of the Gas-Liquid flow in a square cross-sectioned bubble column." Chemical Engineering Science **56**(21–22): 6341-6349.
- Defrance, L. and M. Y. Jaffrin (1999). "Reversibility of fouling formed in activated sludge filtration." Journal of Membrane Science **157**(1): 73-84.
- Degaleesan, S., M. Dudukovic and Y. Pan (2001). "Experimental study of gas-induced liquid-flow structures in bubble columns." AIChE Journal **47**(9): 1913-1931.
- Delnoij, E., F. A. Lammers, J. A. M. Kuipers and W. P. M. van Swaaij (1997). "Dynamic simulation of dispersed gas-liquid two-phase flow using a discrete bubble model." Chemical Engineering Science **52**(9): 1429-1458.
- Dewes, I., A. Kuksal and A. Schumpe (1995). "Gas density effect on mass transfer in three-phase sparged reactors." Chemical Engineering Research & Design **73**: 697.

- Dhotre, M. T., B. Niceno and B. L. Smith (2008). "Large eddy simulation of a bubble column using dynamic sub-grid scale model." Chemical Engineering Journal **136**(2–3): 337-348.
- Doherty, M. F. and J. M. Ottino (1988). "Chaos in deterministic systems: Strange attractors, turbulence, and applications in chemical engineering." Chemical Engineering Science **43**(2): 139-183.
- Dong, F., Z. X. Jiang, X. T. Qiao and L. A. Xu (2003). "Application of electrical resistance tomography to two-phase pipe flow parameters measurement." Flow Measurement and Instrumentation **14**(4–5): 183-192.
- Dong, F., Y. B. Xu, L. J. Xu, L. Hua and X. T. Qiao (2005). "Application of dual-plane ERT system and cross-correlation technique to measure gas–liquid flows in vertical upward pipe." Flow Measurement and Instrumentation **16**(2–3): 191-197.
- Drahoš, J., J. Zahradník, M. Punčochář, M. Fialová and F. Bradka (1991). "Effect of operating conditions on the characteristics of pressure fluctuations in a bubble column." Chemical Engineering and Processing: Process Intensification **29**(2): 107-115.
- Drews, A. (2010). "Membrane fouling in membrane bioreactors—Characterisation, contradictions, cause and cures." Journal of Membrane Science **363**(1–2): 1-28.
- Ekambara, K. and M. T. Dhotre (2010). "CFD simulation of bubble column." Nuclear Engineering and Design **240**(5): 963-969.
- Fang, H. H. P. and X. Shi (2005). "Pore fouling of microfiltration membranes by activated sludge." Journal of Membrane Science **264**(1-2): 161-166.

- Farag, H. I., T. Mejdell, K. Hjarbo, P. Ege, M. Lysberg, A. GrislingÅs and H. De Lasa (1997). "Fiber optic and capacitance probes in turbulent fluidized beds." Chemical Engineering Communications **157**(1): 73-107.
- Flemming, H. C. (1995). Biofouling bei Membranprozessen. Berlin, Heidelberg, Springer-Verlag.
- Froment, G. F. and K. B. Bischoff (1990). Chemical Reactor Analysis and Design. New York, John Wiley and Sons.
- Fukuma, M., K. Muroyama and S. Morooka (1987). "Properties of bubble swarm in a slurry bubble column." J Chem Eng Jpn **20**: 28-33.
- Gan, Z. W., S. C. M. Yu and A. W. K. Law (2011). "Hydrodynamic stability of a bubble column with a bottom-mounted point air source." Chemical Engineering Science **66**(21): 5338-5356.
- Geldart, D. and J. R. Kelsey (1972). "The use of capacitance probes in gas fluidised beds." Powder Technology **6**(1): 45-50.
- Germain, E., T. Stephenson (2005). "Biomass characteristics, aeration and oxygen transfer in membrane bioreactors: their interrelations explained by a review of aerobic biological processes." Reviews in Environmental Science and Biotechnology **4**(4): 223 - 233.
- Germano, M., U. Piomelli, P. Moin and W. H. Cabot (1991). "A dynamic subgrid-scale eddy viscosity model." Physics of Fluids **3**(7): 1760-1765.
- Glassgow, L. A., J. Hua, T.-Y. Yiin and L. E. Erickson (1992). "Acoustic studies of interfacial effects in airlift reactors." Chemical Engineering Communications **113**(1): 155-181.

- Grassberger, P. and I. Procaccia (1983). "Measuring the strangeness of strange attractors." Physica D: Nonlinear Phenomena **9**(1-2): 189-208.
- Guibert, D., R. B. Aim, H. Rabie and P. Côté (2002). "Aeration performance of immersed hollow-fiber membranes in a bentonite suspension." Desalination **148**(1-3): 395-400.
- Güngör, N. (2000). "Effect of the adsorption of surfactants on the rheology of Na-bentonite slurries." Journal of Applied Polymer Science **75**(1): 107-110.
- Gupta, N., N. Jana and C. B. Majumder (2008). "Submerged membrane bioreactor system for municipal wastewater treatment process: An overview." India Journal of Chemical Technology (IJCT) **15**(6): 604-612.
- Hamachi, M. and M. Mietton-Peuchot (1999). "Experimental investigations of cake characteristics in crossflow microfiltration." Chemical Engineering Science **54**(18): 4023-4030.
- Han, S.-S., T.-H. Bae, G.-G. Jang and T.-M. Tak (2005). "Influence of sludge retention time on membrane fouling and bioactivities in membrane bioreactor system." Process Biochemistry **40**(7): 2393-2400.
- He, Y., P. Xu, C. Li and B. Zhang (2005). "High-concentration food wastewater treatment by an anaerobic membrane bioreactor." Water Research **39**(17): 4110-4118.
- Heinrich, S., N. G. Deen, M. S. van Buijtenen, J. A. M. Kuipers, P. Tripp, M. Schönherr and M. Peglow (2007). Detailed investigation of granulation processes using a fibre-optical probe and discrete element simulations 3rd International Conference on Population Balance Modelling. Quebec, Canada.

- Hikita, H., S. Asai, K. Tanigawa, K. Segawa and M. Kitao (1980). "Gas hold-up in bubble columns." The Chemical Engineering Journal **20**(1): 59-67.
- Höller, V., M. Ruzicka, J. Drahos, L. Kiwi-Minsker and A. Renken (2003). "Acoustic and visual study of bubble formation processes in bubble columns staged with fibrous catalytic layers." Catalysis Today **79–80**(0): 151-157.
- Hong, S. P., T. H. Bae, T. M. Tak, S. Hong and A. Randall (2002). "Fouling control in activated sludge submerged hollow fiber membrane bioreactors." Desalination **143**(3): 219-228.
- Hyndman, C. L., F. Larachi and C. Guy (1997). "Understanding gas-phase hydrodynamics in bubble columns: a convective model based on kinetic theory." Chemical Engineering Science **52**(1): 63-77.
- Ishii, M. and N. Zuber (1979). "Drag coefficient and relative velocity in bubbly, droplet or particulate flows." AIChE Journal **25**(5): 843-855.
- J.C. Orantes, C. W., M. Heran, A. Grasmick (2004). "Influence of total sludge retention on the performance of a submerged membrane bioreactor." Proceedings of the Water Environment-Membrane Technology Conference, Seoul, Korea.
- Jakobsen, H. A. (2008). Fluidized bed reactors. Chemical Reactor Modeling, Springer Berlin Heidelberg: 867-952.
- Jefferson, A. B., P. Le Clech, S. J. Judd (2004). "Methods for understanding organic fouling in MBRs." Water Science & Technology **49**: 237-244.
- Jeffery, C. (2005). ZeeWeed MBR Technology Update. SAWEA Workshop

- Ji, L. and J. Zhou (2006). "Influence of aeration on microbial polymers and membrane fouling in submerged membrane bioreactors." Journal of Membrane Science **276**(1-2): 168-177.
- JINSONG, Z., C. H. CHUAN, JITI, #160, Zhou, FANE and A. G. (2006). Effect of sludge retention time on membrane bio-fouling intensity in a submerged membrane bioreactor. Basingstoke, ROYAUME-UNI, Taylor & Francis.
- Joshi, M. (2011). "CFD Simulation of Hydrodynamics of Bubbly Flow in Bubble Column." Journal of Engineering and Applied Sciences **6**(1).
- Judd, S. (2011). Chapter 3 - Design, Operation and Maintenance. The MBR Book (Second Edition). Oxford, Butterworth-Heinemann: 209-288.
- Julián, I., F. Gallucci, M. van Sint Annaland, J. Herguido and M. Menéndez (2012). "Coupled PIV/DIA for fluid dynamics studies on a Two-Section Two-Zone Fluidized Bed Reactor." Chemical Engineering Journal **207–208**(0): 122-132.
- Kang, C.-W., J. Hua, J. Lou, W. Liu and E. Jordan (2008). "Bridging the gap between membrane bio-reactor (MBR) pilot and plant studies." Journal of Membrane Science **325**(2): 861-871.
- Kang, I.-J., C.-H. Lee and K.-J. Kim (2003). "Characteristics of microfiltration membranes in a membrane coupled sequencing batch reactor system." Water Research **37**(5): 1192-1197.
- Kantarci, N., F. Borak and K. O. Ulgen (2005). "Bubble column reactors." Process Biochemistry **40**(7): 2263-2283.

- Kantarci, N., F. Borak and K. O. Ulgen (2005). "Review: Bubble column reactors." Process Biochemistry **40**: 2263-2283.
- Khalili-Garakani, A., M. R. Mehrnia, N. Mostoufi and M. H. Sarrafzadeh (2011). "Analyze and control fouling in an airlift membrane bioreactor: CFD simulation and experimental studies." Process Biochemistry **46**(5): 1138-1145.
- Kikuchi, R., T. Yano, A. Tsutsumi, K. Yoshida, M. Punchochar and J. Drahos (1997). "Diagnosis of chaotic dynamics of bubble motion in a bubble column." Chemical Engineering Science **52**(21-22): 3741-3745.
- Kim, B. S., A. K. Khambampati, Y. J. Hong, S. Kim and K. Y. Kim (2013). "Multiphase flow imaging using an adaptive multi-threshold technique in electrical resistance tomography." Flow Measurement and Instrumentation(0).
- Kim, H. Y., Yeon, #160, Kyung-Min, Lee, Chung-Hak, Sangho, SWAMINATHAN and Tyagarajan (2006). Biofilm structure and extracellular polymeric substances in low and high dissolved oxygen membrane bioreactors. Basingstoke, ROYAUME-UNI, Taylor & Francis.
- Kim, J., M. Kang, H. Chio, S.Kim (2004). "Characteristics of membrane and module affecting membrane fouling." Proceedings of the Water Environment-Membrane Technology Conference, Seoul, Korea.
- Koide, K., S. Morooka, K. Ueyama and A. Matsuura (1979). "Behavior of bubbles in large scale bubble column reactor." J Chem Eng Jpn **12**: 98-104.

- Kraume, M. and A. Drews (2010). "Membrane Bioreactors in Waste Water Treatment – Status and Trends." Chemical Engineering & Technology **33**(8): 1251-1259.
- Krishna, R., J. W. A. De Swart, J. Ellenberger, G. B. Martina and C. Maretto (1997). "Gas holdup in slurry bubble columns: Effect of column diameter and slurry concentrations." AIChE Journal **43**(2): 311-316.
- Krishna, R., M. I. Urseanu, J. M. van Baten and J. Ellenberger (1999). "Influence of scale on the hydrodynamics of bubble columns operating in the churn-turbulent regime: experiments vs. Eulerian simulations." Chemical Engineering Science **54**(21): 4903-4911.
- Krishna, R., M. I. Urseanu, J. M. van Baten and J. Ellenberger (2000). "Liquid phase dispersion in bubble columns operating in the churn-turbulent flow regime." Chemical Engineering Journal **78**(1): 43-51.
- Krishna, R. and J. M. van Baten (2003). "Mass transfer in bubble columns." Catalysis Today **79–80**(0): 67-75.
- Kulkarni, A. A., K. Ekambara and J. B. Joshi (2007). "On the development of flow pattern in a bubble column reactor: Experiments and CFD." Chemical Engineering Science **62**(4): 1049-1072.
- Kumar, S. B., D. Moslemian and M. P. Duduković (1995). "A γ -ray tomographic scanner for imaging voidage distribution in two-phase flow systems." Flow Measurement and Instrumentation **6**(1): 61-73.
- Kunii, D. and O. Levenspiel (1991). Fluidization Engineering. Boston, Butterworth-Heinemann.

- Laborde-Boutet, C., F. Larachi, N. Dromard, O. Delsart and D. Schweich (2009). "CFD simulation of bubble column flows: Investigations on turbulence models in RANS approach." Chemical Engineering Science **64**(21): 4399-4413.
- Lain, S., D. Bröder and M. Sommerfeld (1999). "Experimental and numerical studies of the hydrodynamics in a bubble column." Chemical Engineering Science **54**(21): 4913-4920.
- Lain, S., D. Bröder, M. Sommerfeld and M. F. Göz (2002). "Modelling hydrodynamics and turbulence in a bubble column using the Euler-Lagrange procedure." International Journal of Multiphase Flow **28**(8): 1381-1407.
- Lakehal, D., B. L. Smith; and M. Milelli; (2002). "Large eddy simulation of bubbly turbulent shear flows." Journal of Turbulence **3**(N25).
- Latifi, M. A., N. Midoux, A. Storck and J. N. Gence (1989). "The use of micro-electrodes in the study of the flow regimes in a packed bed reactor with single phase liquid flow." Chemical Engineering Science **44**(11): 2501-2508.
- Laverman, J. A., I. Roghair, M. van Sint Annaland and J. A. M. Kuipers (2007). Experimental study on solids mixing and bubble behavior in a pseudo-2D, freely bubbling, gas-solid fluidized bed using PIV and DIA. 6th International Conference on Multiphase Flow, ICMF 2007. Leipzig, Germany.
- Le-Clech, P. (2010). "Membrane bioreactors and their uses in wastewater treatments." Applied Microbiology and Biotechnology **88**(6): 1253-1260.

- Le-Clech, P., Z. Cao, P. Y. Wan, D. E. Wiley and A. G. Fane (2006). "The application of constant temperature anemometry to membrane processes." Journal of Membrane Science **284**(1-2): 416-423.
- Le-Clech, P., V. Chen and T. A. G. Fane (2006). "Fouling in membrane bioreactors used in wastewater treatment." Journal of Membrane Science **284**(1-2): 17-53.
- Le-Clech, P., B. Jefferson, I. S. Chang and S. J. Judd (2003). "Critical flux determination by the flux-step method in a submerged membrane bioreactor." Journal of Membrane Science **227**(1-2): 81-93.
- Letzel, H. M., J. C. Schouten, R. Krishna and C. M. van den Bleek (1997). "Characterization of regimes and regime transitions in bubble columns by chaos analysis of pressure signals." Chemical Engineering Science **52**(24): 4447-4459.
- li-shun, H., W. xing-jun, Y. guang-suo, W. yi-fei, Z. zhi-jie, W. fu-chen and Y. zun-hong (2009). "Chaotic analysis of pressure fluctuation signal in the gas–liquid–solid slurry column." Nonlinear Analysis: Real World Applications **10**(1): 410-415.
- Li, H. and A. Prakash (1997). "Heat Transfer and Hydrodynamics in a Three-Phase Slurry Bubble Column." Industrial & Engineering Chemistry Research **36**(11): 4688-4694.
- Li, H. and A. Prakash (2000). "Influence of slurry concentrations on bubble population and their rise velocities in a three-phase slurry bubble column." Powder Technology **113**(1–2): 158-167.

- Lin, T. J., R. C. Juang, Y. C. Chen and C. C. Chen (2001). "Predictions of flow transitions in a bubble column by chaotic time series analysis of pressure fluctuation signals." Chemical Engineering Science **56**(3): 1057-1065.
- Lin, T. J., J. Reese, T. Hong and L. S. Fan (1996). "Quantitative analysis and computation of two-dimensional bubble columns." AIChE Journal **42**(2): 301-318.
- Liu, R., X. Huang, Y. F. Sun and Y. Qian (2003). "Hydrodynamic effect on sludge accumulation over membrane surfaces in a submerged membrane bioreactor." Process Biochemistry **39**(2): 157-163.
- Liu, R., X. Huang, C. Wang, L. Chen and Y. Qian (2000). "Study on hydraulic characteristics in a submerged membrane bioreactor process." Process Biochemistry **36**(3): 249-254.
- Liu, Y. and J.-H. Tay (2004). "State of the art of biogranulation technology for wastewater treatment." Biotechnology Advances **22**(7): 533-563.
- Luo, L., Y. Yan, P. Xie, J. Sun, Y. Xu and J. Yuan (2012). "Hilbert–Huang transform, Hurst and chaotic analysis based flow regime identification methods for an airlift reactor." Chemical Engineering Journal **181–182**(0): 570-580.
- Luo, X., D. J. Lee, R. Lau, G. Yang and L.-S. Fan (1999). "Maximum stable bubble size and gas holdup in high-pressure slurry bubble columns." AIChE Journal **45**(4): 665-680.
- Madaeni, S. S., A. G. Fane and D. E. Wiley (1999). "Factors influencing critical flux in membrane filtration of activated sludge." Journal of Chemical Technology & Biotechnology **74**(6): 539-543.

- Mahto, V. and V. P. Sharma (2004). "Rheological study of a water based oil well drilling fluid." Journal of Petroleum Science and Engineering **45**(1–2): 123-128.
- Matsui, G. (1984). "Identification of flow regimes in vertical gas-liquid two-phase flow using differential pressure fluctuations." International Journal of Multiphase Flow **10**(6): 711-719.
- Meng, F., S.-R. Chae, A. Drews, M. Kraume, H.-S. Shin and F. Yang (2009). "Recent advances in membrane bioreactors (MBRs): Membrane fouling and membrane material." Water Research **43**(6): 1489-1512.
- Michaelides, E. E. S. (2013). Fluidized bed reactors. Heat and Mass Transfer in Particulate Suspensions, Springer New York: 89-119.
- Miyahara, T., K. Tsuchiya and L. S. Fan (1988). "Wake properties of a single gas bubble in three-dimensional liquid-solid fluidized bed." International Journal of Multiphase Flow **14**(6): 749-763.
- Mudde, R. F. and O. Simonin (1999). "Two- and three-dimensional simulations of a bubble plume using a two-fluid model." Chemical Engineering Science **54**(21): 5061-5069.
- Mulder, M. (1996). Basic Principles of Membrane Technology Second Edition, Springer, 1996.
- Mulder, M. (2000). Basic principles of Membrane Technology. Dordrecht, Kluwer Academic Publisher.

- Muller, F. L. and F. Davidson (1995). "On the effects of surfactants on mass transfer to viscous liquids in bubble columns." Chemical Engineering Research & Design **73**: 291-296.
- Murthy, B. N. and J. B. Joshi (2008). "Assessment of standard , RSM and LES turbulence models in a baffled stirred vessel agitated by various impeller designs." Chemical Engineering Science **63**(22): 5468-5495.
- Ndinisa, N. V., A. G. Fane, D. E. Wiley and D. F. Fletcher (2006). "Fouling Control in a Submerged Flat Sheet Membrane System: Part II—Two - Phase Flow Characterization and CFD Simulations." Separation Science and Technology **41**(7): 1411-1445.
- Nedeltchev, S., S. Aradhya, F. Zaid and M. Al-Dahhan (2011). Flow Regime Identification In Different Two-Phase Reactors Based On Extraction of Various Entropies From Gauge Pressure Fluctuations. 11 AIChE annual meeting. Minneapolis, MN.
- Nedeltchev, S., S. B. Kumar and M. P. Dudukovic (2003). "Flow Regime Identification in a Bubble Column Based on both Kolmogorov Entropy and Quality of Mixedness Derived from CAERPT Data." The Canadian Journal of Chemical Engineering **81**(3-4): 367-374.
- Ng, H. Y. and S. W. Hermanowicz (2005). "Membrane bioreactor operation at short solids retention times: performance and biomass characteristics." Water Research **39**(6): 981-992.
- Ní Mhurchú, J. and G. Foley (2008). Dead-end and crossflow microfiltration of yeast and bentonite suspensions: Experimental and modelling studies incorporating the use of artificial neural networks. School of Biotechnology Dublin City University: 287.

- Nishikawa, K., K. Sekoguchi and T. Fukano (1969). "On the Pulsation Phenomena in Gas-Liquid Two-Phase Flow : Relationship between Pulsating Pressure and Flow Pattern in Upward Two-Phase Flow." Bulletin of JSME **12**(54): 1410-1416.
- Ognier, S., C. Wisniewski and A. Grasmick (2002). "Influence of macromolecule adsorption during filtration of a membrane bioreactor mixed liquor suspension." Journal of Membrane Science **209**(1): 27-37.
- Olmos, E., C. Gentric, S. Poncin and N. Midoux (2003). "Description of flow regime transitions in bubble columns via laser Doppler anemometry signals processing." Chemical Engineering Science **58**(9): 1731-1742.
- öztürk, S. S., A. Schumpe and W. D. Deckwer (1987). "Organic liquids in a bubble column: Holdups and mass transfer coefficients." AIChE Journal **33**(9): 1473-1480.
- Park, J.-S., K.-M. Yeon and C.-H. Lee (2005). "Hydrodynamics and microbial physiology affecting performance of a new MBR, membrane-coupled high-performance compact reactor." Desalination **172**(2): 181-188.
- Park, J. S., K. M. Yeon and C. H. Lee (2005). "Hydrodynamics and microbial physiology affecting performance of a new MBR, membrane-coupled high-performance compact reactor." Desalination **172**(2): 181-188.
- Park, S. H. and S. D. Kim (2003). "Characterization of pressure signals in a bubble column by wavelet packet transform." Korean Journal of Chemical Engineering **20**(1): 128-132.

- Pfleger, D. and S. Becker (2001). "Modelling and simulation of the dynamic flow behaviour in a bubble column." Chemical Engineering Science **56**(4): 1737-1747.
- Pfleger, D., S. Gomes, N. Gilbert and H. G. Wagner (1999). "Hydrodynamic simulations of laboratory scale bubble columns fundamental studies of the Eulerian–Eulerian modelling approach." Chemical Engineering Science **54**(21): 5091-5099.
- Pino, L. Z., R. B. Solari, S. Siquier and L. Antonio (1992). "Effect of operating conditons on gas holdup in slurry bubble columns with a foaming liquid." Chemical Engineering Communications **117**(1): 367-382.
- Pollice, A., A. Brookes, B. Jefferson and S. Judd (2005). "Sub-critical flux fouling in membrane bioreactors -- a review of recent literature." Desalination **174**(3): 221-230.
- Prakash, A., A. Margaritis, H. Li and M. A. Bergougnou (2001). "Hydrodynamics and local heat transfer measurements in a bubble column with suspension of yeast." Biochemical Engineering Journal **9**(2): 155-163.
- Prieske, H., A. Drews and M. Kraume (2008). "Prediction of the circulation velocity in a membrane bioreactor." Desalination **231**(1-3): 219-226.
- Psoch, C. and S. Schiewer (2005). "Long-term study of an intermittent air sparged MBR for synthetic wastewater treatment." Journal of Membrane Science **260**(1-2): 56-65.
- Ranade, V. V. and Y. Tayalia (2001). "Modelling of fluid dynamics and mixing in shallow bubble column reactors: influence of sparger design." Chemical Engineering Science **56**(4): 1667-1675.

- Reilley, I. G., D. S. Scott, T. De Bruijin, A. Jain and J. A. Piskorz (1986). "A correlation for gas holdup in turbulent coalescing bubble columns." Can J Chem Eng **64**(705-17).
- Rosenberger, S., C. Laabs, B. Lesjean, R. Gnirss, G. Amy, M. Jekel and J. C. Schrotter (2006). "Impact of colloidal and soluble organic material on membrane performance in membrane bioreactors for municipal wastewater treatment." Water Research **40**(4): 710-720.
- Rowe, P. N. and B. A. Partridge (1997). "An x-ray study of bubbles in fluidised beds." Chemical Engineering Research and Design **75, Supplement(0)**: S116-S134.
- Roy, S., T. Bauer and M. Al-Dahhan (2004). "Monoliths as multiphase reactors: a review." AIChE Journal **50**(11): 2918.
- Roy, S., A. Kemoun, M. H. Al-Dahhan, M. P. Dudukovic, T. B. Skourlis and F. M. Dautzenberg (2005). "Countercurrent flow distribution in structured packing via computed tomography." Chemical Engineering and Processing: Process Intensification **44**(1): 59-69.
- Ruelle, D. (1979). "SENSITIVE DEPENDENCE ON INITIAL CONDITION AND TURBULENT BEHAVIOR OF DYNAMICAL SYSTEMS." Annals of the New York Academy of Sciences **316**(1): 408-416.
- Ruelle, D. (1995). Turbulence, strange attractors, and chaos, World Scientific Publishing Co.Pte.Ltd.
- Sada, E., S. Katoh and H. Yoshil (1984). "Performance of the gas-liquid bubble column in molten salt systems." Ind Eng Chem Process Des Dev **23**: 151-154.

- Santos, A., W. Ma and S. J. Judd (2011). "Membrane bioreactors: Two decades of research and implementation." Desalination **273**(1): 148-154.
- Sanyal, J., S. Vásquez, S. Roy and M. P. Dudukovic (1999). "Numerical simulation of gas-liquid dynamics in cylindrical bubble column reactors." Chemical Engineering Science **54**(21): 5071-5083.
- Saxena, S. C., N. S. Rao and A. C. Saxena (1990). "Heat-transfer and gas-holdup studies in a bubble column: air-water-glass bead system." Chemical Engineering Communications **96**(1): 31-55.
- Schouten, J. C. and C. M. van den Bleek (1998). "Monitoring the quality of fluidization using the short-term predictability of pressure fluctuations." AIChE Journal **44**(1): 48-60.
- Schumpe, A. and G. Grund (1986). "The gas disengagement technique for studying gas holdup structure in bubble columns." The Canadian Journal of Chemical Engineering **64**(6): 891-896.
- Shah, Y. T., B. G. Kelkar, S. P. Godbole and W. D. Deckwer (1982). "Design parameters estimations for bubble column reactors." AIChE Journal **28**(3): 353-379.
- Shaikh, A. and M. Al-Dahhan (2005). "Characterization of the hydrodynamic flow regime in bubble columns via computed tomography." Flow Measurement and Instrumentation **16**(2-3): 91-98.
- Shaikh, A. and M. H. Al-Dahhan (2007). "A review on flow regime transition in bubble columns." International Journal of Chemical Reactor Engineering **5**(1).

- Shou, M. C. and L. P. Leu (2005). "Energy of Power Spectral Density Function and Wavelet Analysis of Absolute Pressure Fluctuation Measurements in Fluidized Beds." Chemical Engineering Research and Design **83**(5): 478-491.
- Siemens. (2009). "MemPulse™ Membrane Bioreactor (MBR) System." from http://www.water.siemens.com/en/products/biological_treatment/membrane_biological_reactor_systems_mbr/Pages/envirex_product_integrated_mbr.aspx.
- Silva, M. K., M. A. d'Ávila and M. Mori (2012). "Study of the interfacial forces and turbulence models in a bubble column." Computers & Chemical Engineering **44**(0): 34-44.
- Simmonds, C. (1919). Composition of Yeast. Alcohol, Its Production, Properties, Chemistry, And Industrial Applications, Macmillan And Co.
- Simonnet, M., C. Gentric, E. Olmos and N. Midoux (2008). "CFD simulation of the flow field in a bubble column reactor: Importance of the drag force formulation to describe regime transitions." Chemical Engineering and Processing: Process Intensification **47**(9-10): 1726-1737.
- Smith, S. W. (1997). The Scientist and Engineer's Guide to Digital Signal Processing, California Technical Pub.
- Sofia, A., W. J. Ng and S. L. Ong (2004). "Engineering design approaches for minimum fouling in submerged MBR." Desalination **160**(1): 67-74.
- Sokolichin, A. and G. Eigenberger (1994). "Gas—liquid flow in bubble columns and loop reactors: Part I. Detailed modelling and numerical simulation." Chemical Engineering Science **49**(24, Part 2): 5735-5746.

- Sokolichin, A. and G. Eigenberger (1999). "Applicability of the standard k- ϵ turbulence model to the dynamic simulation of bubble columns: Part I. Detailed numerical simulations." Chemical Engineering Science **54**(13-14): 2273-2284.
- Sokolichin, A., G. Eigenberger, A. Lapin and A. Lübert (1997). "Dynamic numerical simulation of gas-liquid two-phase flows Euler/Euler versus Euler/Lagrange." Chemical Engineering Science **52**(4): 611-626.
- Sommerfeld, M. and D. Bröder (2009). "Analysis of Hydrodynamics and Microstructure in a Bubble Column by Planar Shadow Image Velocimetry." Industrial & Engineering Chemistry Research **48**(1): 330-340.
- Sur, H. W. and Z. F. Cui (2005). "Enhancement of microfiltration of yeast suspensions using gas sparging – effect of feed conditions." Separation and Purification Technology **41**(3): 313-319.
- Swinney, H. L. and J. P. Gollub (1986). "Characterization of hydrodynamic strange attractors." Physica D: Nonlinear Phenomena **18**(1-3): 448-454.
- Tabib, M. V., S. A. Roy and J. B. Joshi (2008). "CFD simulation of bubble column—An analysis of interphase forces and turbulence models." Chemical Engineering Journal **139**(3): 589-614.
- Tadaki, T. and S. Maeda (1961). "On the shape and velocity of single air bubbles rising in various liquids." Kagaku Kogaku **25**: 254-264.
- Thimmapuram, P. R., N. S. Rao and S. C. Saxena (1992). "Characterization of hydrodynamic regimes in a bubble column." Chemical Engineering Science **47**(13-14): 3355-3362.

- Tomiyama, A. (2004). Drag, lift and virtual mass forces acting on a single bubble. Third International Symposium on Two-Phase Flow Modeling and Experimentation. Pisa, Italy.
- Trnka, O., V. Veselý, M. Hartman and Z. Beran (2000). "Identification of the state of a fluidized bed by pressure fluctuations." AIChE Journal **46**(3): 509-514.
- Ueda, T., K.Hata, Y. Kikuota and O. Seino (1997). "Effects of aeration on suction pressure in a submerged membrane bioreactor." Water Res. **31**: 489-494.
- van Lare, C. E. J., H. W. Piepers, J. N. Schoonderbeek and D. Thoenes (1997). "Investigation on bubble characteristics in a gas fluidized bed." Chemical Engineering Science **52**(5): 829-841.
- van Ommen, J. R., S. Sasic, J. van der Schaaf, S. Gheorghiu, F. Johnsson and M.-O. Coppens (2011). "Time-series analysis of pressure fluctuations in gas–solid fluidized beds – A review." International Journal of Multiphase Flow **37**(5): 403-428.
- Vázquez, C., J. L. Nombela, M. d. Vega, J. Zubía and D. Sánchez Montero (2007). Plastic fiber-optic probes for characterizing fluidized beds in bubbling regime. 16th International Conference on Plastic Optical Fibers.
- Veera, U. P., K. L. Kataria and J. B. Joshi (2004). "Effect of superficial gas velocity on gas hold-up profiles in foaming liquids in bubble column reactors." Chem Eng J **99**: 53-58.
- Vera, L., S. Delgado and S. Elmaleh (2000). "Dimensionless numbers for the steady-state flux of cross-flow microfiltration and ultrafiltration with gas sparging." Chemical Engineering Science **55**(17): 3419-3428.

- Verma, A. K. and S. Rai (2003). "Studies on surface to bulk ionic mass transfer in bubble column." Chem Eng J **94**: 67-72.
- Versteeg, H. K. and W. Malalasekera (2007). An introduction to computational fluid dynamics: the finite volume method, Pearson Education Limited.
- Versteeg, H. K. and W. Malalasekera (2007). An introduction to computational fluid dynamics: the finite volume method, Pearson Education Limited.
- Vial, C., E. Camarasa, S. Poncin, G. Wild, N. Midoux and J. Bouillard (2000). "Study of hydrodynamic behaviour in bubble columns and external loop airlift reactors through analysis of pressure fluctuations." Chemical Engineering Science **55**(15): 2957-2973.
- Vial, C., R. Lainé, S. Poncin, N. Midoux and G. Wild (2001). "Influence of gas distribution and regime transitions on liquid velocity and turbulence in a 3-D bubble column." Chemical Engineering Science **56**(3): 1085-1093.
- Vial, C., S. Poncin, G. Wild and N. Midoux (2001). "A simple method for regime identification and flow characterisation in bubble columns and airlift reactors." Chemical Engineering and Processing: Process Intensification **40**(2): 135-151.
- Vijayan, M., H. I. Schlager and M. Wang (2007). "Effects of sparger geometry on the mechanism of flow pattern transition in a bubble column." Chemical Engineering Journal **130**(2-3): 171-178.
- Vivek, V. R. (2002). 12 Fluidized bed reactors. Process Systems Engineering, Academic Press. **Volume 5**: 367-402.

- Wang, Y., M. Brannock, S. Cox and G. Leslie (2010). "CFD simulations of membrane filtration zone in a submerged hollow fibre membrane bioreactor using a porous media approach." Journal of Membrane Science **363**(1-2): 57-66.
- Warsito, W. and L. S. Fan (2001). "Measurement of real-time flow structures in gas–liquid and gas–liquid–solid flow systems using electrical capacitance tomography (ECT)." Chemical Engineering Science **56**(21–22): 6455-6462.
- Warsito, W. and L. S. Fan (2005). "Dynamics of spiral bubble plume motion in the entrance region of bubble columns and three-phase fluidized beds using 3D ECT." Chemical Engineering Science **60**(22): 6073-6084.
- Werther, J. (1999). "Measurement techniques in fluidized beds." Powder Technology **102**(1): 15-36.
- Werther, J. (2000). Fluidized-Bed Reactors. Ullmann's Encyclopedia of Industrial Chemistry, Wiley-VCH Verlag GmbH & Co. KGaA.
- Wicaksana, F., A. G. Fane and V. Chen (2005). "The relationship between critical flux and fibre movement induced by bubbling in a submerged hollow fibre system." Water Science & Technology **51**(6-7): 115-122.
- Wicaksana, F., A. G. Fane and V. Chen (2006). "Fibre movement induced by bubbling using submerged hollow fibre membranes." Journal of Membrane Science **271**(1-2): 186-195.
- Wiesendorf, V. and J. Werther (2000). "Capacitance probes for solids volume concentration and velocity measurements in industrial fluidized bed reactors." Powder Technology **110**(1–2): 143-157.

- Wolf, A., J. B. Swift, H. L. Swinney and J. A. Vastano (1985). "Determining Lyapunov exponents from a time series." Physica D: Nonlinear Phenomena **16**(3): 285-317.
- Wu, J. and C. He (2012). "Effect of cyclic aeration on fouling in submerged membrane bioreactor for wastewater treatment." Water Research **46**(11): 3507-3515.
- Xu, N., W. Xing, N. Xu and J. Shi (2002). "Application of turbulence promoters in ceramic membrane bioreactor used for municipal wastewater reclamation." Journal of Membrane Science **210**(2): 307-313.
- Yamato, N., K. Kimura, T. Miyoshi and Y. Watanabe (2006). "Difference in membrane fouling in membrane bioreactors (MBRs) caused by membrane polymer materials." Journal of Membrane Science **280**(1-2): 911-919.
- Yates, J. D., D. J. Cheesman, P. Lettieri and D. Newton (2002). "X-ray analysis of fluidized bed and other multiphase systems." Kona **20**: 133-143.
- Yeo, A., Fane A.G. (2004). "Performnace of individual fibers in a submerged hollow fiber bundle." Proceedings of the Water Environment-Membrane Technology Conference, Seoul, Korea.
- Yeo, A., Fane A.G. (2005). "Performance of individual fibers in a submerged hollow fiber bundle." Water Science & Technology **51**((6-7)): 165-172.
- Yeo, A. P. S., A. W. K. Law and A. G. Fane (2006). "Factors affecting the performance of a submerged hollow fiber bundle." Journal of Membrane Science **280**(1-2): 969-982.

- Yeom, I. T., K. R. Lee, Y. G. Choi, H. S. Kim, Y. Lee (2004). "Evaluation of a membrane bioreactor system coupled with sludge pretreatment for aerobic sludge digestion." Proceedings of the Water Environment-Membrane Technology Conference, Seoul, Korea.
- Yu, H.-Y., M.-X. Hu, Z.-K. Xu, J.-L. Wang and S.-Y. Wang (2005). "Surface modification of polypropylene microporous membranes to improve their antifouling property in MBR: NH₃ plasma treatment." Separation and Purification Technology **45**(1): 8-15.
- Zhang, D., N. G. Deen and J. A. M. Kuipers (2006). "Numerical simulation of the dynamic flow behavior in a bubble column: A study of closures for turbulence and interface forces." Chemical Engineering Science **61**(23): 7593-7608.
- Zhang, J., H. C. Chua, J. Zhou and A. G. Fane (2006). "Factors affecting the membrane performance in submerged membrane bioreactors." Journal of Membrane Science **284**(1-2): 54-66.
- Zhang, J. P., J. R. Grace, N. Epstein and K. S. Lim (1997). "Flow regime identification in gas-liquid flow and three-phase fluidized beds." Chemical Engineering Science **52**(21-22): 3979-3992.
- Zhang, S., Y. Qu, Y. Liu, F. Yang, X. Zhang, K. Furukawa and Y. Yamada (2005). "Experimental study of domestic sewage treatment with a metal membrane bioreactor." Desalination **177**(1-3): 83-93.