

Dry beneficiation of -13mm South African coal using an air dense medium fluidization bed

ES Diedericks



orcid.org / 0000-0001-9597-0621

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degree Doctor of Philosophy in Chemical Engineering at
the North-West University

Promoter: Prof M le Roux
Co-promoter: Prof QP Campbell

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Student number: 21683883

Preface

Document overview

This thesis is compiled in completion of a degree in Ph.D.Eng.Chemical at the North-West University. An overview of the thesis submitted for examination is summarized in the preface. Student identification and relevant information is provided in the preface. The student identification and relevant information is followed by the main rules and guidelines detailed by the North West University is listed. Moreover, signed statements of consent to use publication in completion of the thesis is provided by all co-authors. The deliverables of the study are discussed in short and lastly, the abstract of the thesis is provided.

Student identification and relevant information:

Full names and surname:	Elmarie Sunette Diedericks
Student number:	21683883
Ethics number	NWU-01709-19-A1
Orcid number	0000-0001-9597-0621
Highest qualification:	M.Eng Chemical
Current study level:	Ph.D. Eng Chemical
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Department:	Department of Chemical and Minerals Engineering Coal research group
Promotors:	Professor M. Le Roux Professor Q.P. Campbell

SOLEMN DECLARATION

I, **Elmarie Sunette Diedericks**, declare herewith that the dissertation entitled:

“Dry beneficiation of -13mmSouth African coal using an air dense medium fluidization bed”

which I hereby submit, in fulfilment of the requirements set for the degree Doctor of Philosophy in Engineering, to the North-West University is my own individual work and has not previously been submitted to another institute.

Credit has been given to all references used during this study. I recognize that I am responsible for acquiring any necessary copyright permissions. I have therefore obtained included written permission declarations from each co-writer that will be included in this dissertation. With this, I grant The North West University and its representatives the unrestricted license to make accessible this work in all forms of media.

Thesis format

This thesis is composed of original articles, either published or ready for publication, as permitted by rule 5.3.1 of the General Academic Rules 2017, approved on the 21st of September 2017 and revised in March 2020. The General Academic Rules 2017 is available at http://www.nwu.ac.za/content/policy_rules (accessed 12th July 2021). In addition to this rule, there are other rules stipulated in the General Academic Rules 2017, which are relevant to the publishing a doctor's degree in article format. These rules are 5.10.4; 5.10.5; 5.10.8 and 5.10.9.

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Thesis formatting, referencing, and numbering style

It should be noted that the formatting, numbering of all figures and tables and referencing style were normalized, in order to maintain a uniform thesis layout. A list of references consulted during each chapter is provided at the end of that respective chapter. A comprehensive list of all of the references consulted during the thesis is provided at the end of the thesis.

Co-Author Statements

This section contains the statements of consent of the co-authors of the various articles that are contained in this thesis. These consents agree with Rule 5.10.8 and Rule 5.10.9 of the General Academic Rules 2017 of the North-West University as provided on page vii.

The following co-authors contributed to the articles either published or presented in this thesis:

- Prof. Marco Le Roux (page ix)
- Prof. Quentin P. Campbell (page x)
- Mrs Nikki Hughes (page xi)
- Mr Louis Le Grange (page xii)

Prof. M le Roux

I, **MARCO LE ROUX**, understand and give full consent to **ELMARIE SUNETTE DIEDERICKS**, to use the publications listed below, of which I am co-author, in fulfilment of the thesis titled: “**Dry beneficiation of -13mm South African coal using an air dense medium fluidization bed**”.

- Diedericks, E.S., **Le Roux, M.**, Campbell, Q.P. & Hughes, N. 2020. Dry beneficiation of small South African coal using an air dense medium fluidization bed. *International Journal of Coal Preparation and Utilization*, 42:1582-1595. DOI: 10.1080/19392699.2020.1758682
- Diedericks, E.S., **Le Roux, M.**, Campbell, Q.P. & Hughes, N. 2019. Dry beneficiation of small South African coal using an air dense medium fluidization bed. XIX International Coal Preparation Congress, New Delhi, India. Woodhead Publishing India Pvt Ltd: Vol 2: 1-9.
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- Diedericks, E.S., Le Grange, L., **Le Roux, M.**, & Campbell, Q.P. 2022. Use of CFD modelling to assist with air dense medium fluidization of South African coal.

These articles form part of the completion of the theses, “**Dry beneficiation of -13mm South African coal using an air dense medium fluidization bed**”, of which I am promotor. The study is focussed on investigating the response and performance of -13.2mm South African coal in an air dense medium fluidization bed. I was involved throughout the duration of the study, more specifically conceptual, experimental and article formulation. Moreover, I also proofed and edited the publications listed above, as written by Elmarie Diedericks. This declaration fulfils Rule 5.10.8 and Rule 5.10.9 of the General Academic Rules 2017 of the North-West University.

Prof. M. Le Roux

Prof. Quentin P. Campbell

I, **QUENTIN PETER CAMPBELL**, understand and give full consent to **ELMARIE SUNETTE DIEDERICKS**, to use the publications listed below, of which I am co-author, in fulfilment of the thesis titled: “***Dry beneficiation of -13mm South African coal using an air dense medium fluidization bed***”.

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- Diedericks, E.S., Le Roux, M. & **Campbell, Q.P.** Submitted 2021. Dry beneficiation of +0.5mm-5.6mm South African coal using an air dense medium fluidization bed. *Journal of the Southern African Institute for Mining and Metallurgy*. Pg. 1-19. (Submission ID: 1702-5380-1SM)
- Diedericks, E.S., Le Grange, L., Le Roux, M., & **Campbell, Q.P.** 2022. Use of CFD modelling to assist with air dense medium fluidization of South African coal

Ms. N. Hughes

I, **NIKKI HUGHES**, understand and give full consent to **ELMARIE SUNETTE DIEDERICKS**, to use the publications listed below, of which I am co-author, in fulfilment of the thesis titled: ***“Dry beneficiation of -13mm South African coal using an air dense medium fluidization bed”***.

- Diedericks, E.S., Le Roux, M., Campbell, Q.P. & **Hughes, N.** 2020. Dry beneficiation of small South African coal using an air dense medium fluidization bed. *International Journal of Coal Preparation and Utilization*, 42:1582-1595. DOI: 10.1080/19392699.2020.1758682
- Diedericks, E.S., Le Roux, M., Campbell, Q.P. & **Hughes, N.** 2019. Dry beneficiation of small South African coal using an air dense medium fluidization bed. XIX International Coal Preparation Congress, New Delhi, India. Woodhead Publishing India Pvt Ltd: Vol 2: 64-75.

Mr. L. Le Grange

I, **LOUIS LE GRANGE**, understand and give full consent to **ELMARIE SUNETTE DIEDERICKS**, to use the publication listed below, of which I am co-author, in fulfilment of the thesis titled: “***Dry beneficiation of -13mm South African coal using an air dense medium fluidization bed***”.

- Diedericks, E.S., **Le Grange, L.**, Le Roux, M., & Campbell, Q.P. 2020. Use of CFD modelling to assist with air dense medium fluidization of South African coal.

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"From life's first cry to final breath – Jesus commands my destiny."

- Keith Getty/Stuart Townend -

Dedicated to
my father,
Henk Peters
and
my mother,
Gisela Peters

November 2022

DELIVERABLES FROM THIS STUDY

- **Locally Published Paper**

Diedericks, E.S., Le Roux, M. & Campbell, Q.P. Submitted 2021. Dry beneficiation of +0.5mm-5.6mm South African coal using an air dense medium fluidization bed. Journal of the Southern African Institute for Mining and Metallurgy. Pg. 1-19. (Submission ID: 1702-5380-1SM)

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ABSTRACT

Coal remains an important energy-resource base worldwide as it is considered abundant, affordable and reliable (WEC, 2022). South Africa's national energy-resource base is dominated by coal; with coal-fired power stations meeting about 83% of the maximum electricity generation in 2018 (DOE, 2019). South Africa's need for coal-based energy is not likely to change in the coming 20 years, due to the lack of suitable energy alternatives.

Coal is inherently heterogeneous by nature, and it requires liberation and separation of fractional impurities prior to consumption. The process is known as coal beneficiation and is achieved by a broad range of techniques and water treatment practices. Currently industrialized, wet beneficiation processes are known to produce high separation efficiencies however, as the name implies, it requires substantial amounts of water. Difficulties in handling, transport and storage as well as increased capital, operating and transportation cost result from excess water retained in coal (Choung *et al.*, 2006). Furthermore, in water scarce coal producing and processing countries, including Mongolia, Australia and South Africa, suitable water supply is limited and as a result, wet coal beneficiation techniques is not ideal (Chen & Yang, 2003; Zhao *et al.*, 2015).

The industry is researching dry coal beneficiation methods in the hopes of finding a suitable alternative that is capable of producing good separation efficiencies. Dry beneficiation techniques, such as Air Dense Medium Fluidized Bed (ADMFB) system, can prove to be an effective solution to remove ash-forming mineral fractions from ROM coal to improve coal quality and mitigate application issues as well as minimize carbon footprint.

This investigation has been divided into three main sections:

1. Dry beneficiation research studies. A comprehensive literature review compiled from numerous sources investigating ADMFB technology as a viable dry beneficiation technique.
2. Experimental evaluation and testing of a range of operating parameters to identify the best performance of -13.2mm South African coal in a batch ADMFB.
3. Particle segregation and fluidization behaviour in the ADMFB was explored by simulating -13mm South African coal with CFD Euler-Euler approach and by employing, Neptune, a commercial fluid dynamic software.

A batch-wise ADMFB apparatus was used during the evaluation and testing stage, to investigate the performance of high- and low-density ROM particles combined with magnetite as dense medium (Geldart B particles). An experimental plan was set up to

capture the effect of the operating parameters including, PSD, coal to dense medium ratio, addition of vibration and fluidization velocity on the performance of the ADMFB system. Product, middlings and discard ash content (%_{wt}), calorific value (CV), density (g/cm³) as well as system air flow velocity (m/s) were used as process evaluation criteria. Local coal fired power station statistics such as ash content (%_{wt}) and calorific value (CV) levels were also selected as reference point for the process evaluation.

The quality of the coal used in this study was upgraded to varying degrees according to the set variables. During operation of the bed, a large fraction of high ash coal was found in the bottom layer of the bed, producing lower ash yield values to the top layer of the bed (13.0% to 22.4%) compared to ash yields of 82.2% to 74.8% in the bottom of the bed. Insignificant impacts on separation performance were observed for all +2.8mm fractions when adding dense medium (+300µm magnetite with $d_{50} = 490\mu\text{m}$) or applying vibration to the batch setup. Application of vibration to the bed lowered the minimum fluidization velocity of the bed, improved bed stability and assisted with particle striation. Little to no upgrading was seen for - 2.8mm coal particle size ranges. It was observed that the ash forming minerals in the -2.8mm feed coal were lower than that recorded for the +2.8mm, which lead to the conclusion that most of the valuable macerals have already been liberated and therefore the feed coal had a smaller fraction of mineral matter content. Separation of coal and magnetite at these size ranges proved to be challenging due to tiny magnetite particles attaching to the additional exposed surface area of the coal. Nevertheless, the product produced still adhered to demands and standards of the local thermal coal market and in some cases also meet metallurgical standards. This operation philosophy aids in downstream processing by eliminating magnetic separation and recycling of dense media.

For the majority of the ADMFB experimental runs, the best cumulative products were achieved for runs that employed no DM in the absence of vibration. In a coal-magnetite bed, the magnetite is fluidized (in suspension) and the coal particles float or sink in accordance with density. The selection of the best DM (density, size and volume) is difficult as its properties, coal properties and operational parameters of the bed are interdependent when optimizing the performance of the ADMFB. The ADMFB density can be manipulated by fine tuning the DM (fluidization velocity, DM material density and DM particle size).

Computational Fluid Dynamics (CFD) is a useful tool to study the dynamic mechanisms of fluidized beds and is applied in this study to better comprehend the selection of a suitable DM for effective particle segregation. From literature, it was reported that dense medium (DM) had a significant effect on the separation performance of the ADMFB, however during the experimental portion of this study this could not be substantiated. For the most part, it

was found that the experimental beds performed better in the absence of DM. As a result, investigation into DM material density, DM fluidization velocity and DM to coal ratio was conducted by using the CFD simulation software, Neptune. A suitable DM will be selected by evaluating the performance of a binary mixture of dense medium and fine coal by employing CFD simulation. Throughout the study the CFD model was able to demonstrate key characteristics of an ADMFB as found by various researchers in the field. An improved understanding of the role of the DM during density-based segregation of particles was obtained. From the simulation it was found that the best segregation was obtained at a fluidization velocity of 23cm/s using a binary mixture DM consisting of 75% fine coal and 25% magnetite to separate high- and low-density coal particles.

Keywords: Small South African coal, dry coal beneficiation, dense medium, magnetite, air dense medium fluidization, minimum fluidization velocity. CFD.

TABLE OF CONTENTS

Solemn declaration	v
Acknowledgements	xiii
Deliverables from this study	xv
Abstract	xvi
List of figures	i
List of tables	i
List of Acronyms	i
List of symbols	i
CHAPTER 1	3
introduction	3
1.1 Background and motivation	3
1.2 Aim and objectives	6
1.3 Scope of study	8
1.4 Thesis Structure	8
CHAPTER 2	12
Literature review	12
2.1 Overview	12
2.2 Historic development of coal	13
2.1.1 Coal rank	14
2.1.2 Lithotypes	15
2.1.3 Coal macerals	15
2.1.4 Coal characterization and mineralogy	16
2.1.5 Other properties of relevance	16
2.1.6 Density	17
2.3 Significance of coal in South Africa	18
2.3.1 Coal products and uses	19
2.4 Coal preparation and beneficiation	21
2.4.1 Preparation	21

2.4.2	Coal beneficiation	22
2.4.3	Dry beneficiation techniques	25
2.5	Fluidization	26
2.5.1	Fluidization principle	26
2.5.2	Minimum fluidizing velocity.....	29
2.5.3	Pressure and velocity relations	31
2.5.4	Bubble formation and behavior	33
2.5.5	Classification of particles	34
2.6	Dry beneficiation of coal using ADMFB technology	35
2.6.2	Dense medium.....	37
2.6.3	Vibrated air dense medium fluidized beds.....	39
2.6.4	Effect of moisture	41
2.6.5	Particle size, shape and density.....	42
2.6.6	Static bed height and column shape	43
2.7	Development of a numerical model.....	44
2.7.1	Simulation model	47
2.7.2	Governing Equations	48
2.7.2.1	Mass transport equation	48
2.7.2.2	Momentum transport equation	48
2.7.2.3	Interphase transfer modelling.....	49
2.7.2.4	Particle stress modeling.....	50
2.7.2.5	Particle fluctuant kinetic energy transport equation	51
2.7.2.6	Fluid–particle velocity covariance transport equation	52
2.8	Previous Case Studies.....	52
2.9	Conclusion	56
2.10	References	58
3	CHAPTER 3.....	67
3.1	Dry beneficiation of small South African coal using an air dense medium fluidization bed	69

4	CHAPTER 4.....	87
4.1	Dry beneficiation of +0.5mm-5.6mm South African coal using an air dense medium fluidization bed	89
5	CHAPTER 5.....	113
5.1	Dry beneficiation of small South African coal using an air dense medium fluidization bed	115
6	CHAPTER 6.....	128
6.1	Additional Information (Paper 1 to Paper 3)	129
6.1.1	Overview of study	129
6.1.2	Experimental.....	130
6.1.3	Results and discussion	134
7	CHAPTER 7.....	153
7.1	Use of CFD modelling to assist with air dense medium fluidization of South African coal	155
7.2	Additional Information (Paper 4).....	184
8	CHAPTER 8.....	196
8.1	Conclusion.....	197
8.2	Recommendations for future works.....	199

LIST OF FIGURES

Figure 1.1-1: South African statistics taken from CEIC (2019).....	4
Figure 1.1-2: South African rainfall pattern change taken from SAWS (2022)	5
Figure 2.2-1: Stages of coal maturity adapted from WCI (2009).....	14
Figure 2.3-1: Main coalfields in South Africa, taken from SACPS (2015).....	19
Figure 2.4-1: Principle of dense medium separation	23
Figure 2.4-2: Typical colliery coal washing circuit adapted from SANEDI (2011).....	24
Figure 2.5-1: Stages of contacting during fluidization, adapted from Kunii & Levenspiel (2001)	27
Figure 2.5-2: Pseudo fluid properties, adapted from Luo & Chen (2001); Mohanta <i>et al.</i> (2011)	28
Figure 2.5-3: External forces acting on a particle during fluidization, adapted from Sahu <i>et al.</i> (2009)	29
Figure 2.5-4: Pressure vs. velocity diagram of uniform sand particles, adapted from Kunii & Levenspiel (1991)	32
Figure 2.5-5: Geldart powder classification diagram taken from Richardson <i>et al.</i> (2008) ...	34
Figure 2.6-1: DDM (dry dense medium) plant located in Shenhua, China (Kunii & Levenspiel, 1991).....	36
Figure 2.6-2: ΔP across a fluidized bed as a function of superficial velocity adapted from, He <i>et al.</i> (2016 (a)).....	38
Figure 2.6-3: Comparison between magnetite sizes from three sources (adapted from de Korte (2016)).....	39
Figure 2.6-4: ΔP fluctuation of the bed with and without vibration (adapted from He <i>et al.</i> (2015))	40
Figure 2.6-5: Effect of particle shape on the ADMFB separation (taken from Prusti <i>et al.</i> (2015))	43
Figure 2.8-1: World's first modularized dry coal beneficiation plant (Zhao <i>et al.</i> , 2016)	55
Figure 2.8-2: Engineering flowsheet of the modularized system (Zhao <i>et al.</i> , 2016).....	55
Figure 3.1-1: Standard precipitation index for South Africa adapted from SAWS (2017)	70
Figure 3.1-2: Three-dimensional illustration of the ADMFB	74
Figure 3.1-3: Ash values of control run according to bed layers (PSD-05; DM 0:1; VD)	75
Figure 3.1-4: Comparison of bed performance curves for various PSD's (DM 0:1; VD).....	77
Figure 3.1-5: Comparison of bed performance curves for various DM (PSD-01; VD)	80
Figure 3.1-6: Comparison of vibration (PSD-05; DM 0:1)	81
Figure 3.1-7: Comparison of bed performance curves for vibration deactivated (VD) and vibration activated (VA) (PSD-05; DM 0:1).....	82

Figure 4.1-1: Density profiles for +0.5mm-5.6mm coal	95
Figure 4.1-2: ADMFB with (1) airflow system, (2) fluidized bed system and (3) dust control section	96
Figure 4.1-3: Pressure vs. velocity diagram, adapted from Kunii & Levenspiel (1991).....	98
Figure 4.1-4: Experimental determination of minimum fluidization points for pure coal beds	98
Figure 4.1-5: Minimum fluidization points determined for various experimental runs	99
Figure 4.1-6: Comparison of bed performance curves for various PSD's (DM 0:1; VD and VA).....	100
Figure 4.1-7: Comparison of bed performance for +1.0mm-2.0mm and +4.0mm-4.75mm without vibration	101
Figure 4.1-8: Degree of segregation for PSD-02 and PSD-03 compared to literature (He et al., 2016(a)).	103
Figure 4.1-9: Ash analyses for +4.75-5.6mm particles and 30% amplitude and 20 Hz	105
Figure 4.1-10: Ash analyses for +4.75-5.6mm particles at 21.97 Hz	106
Figure 5.1-1: Three-dimensional illustration of the ADMFB with (1) airflow system, (2) fluidized bed system and (3) dust control section	118
Figure 5.1-2: Ash values of control run according to bed layers (PSD-06; DM 0:1; VD)	119
Figure 5.1-3: Comparison of bed performance curves for various PSD's (DM 0:1; VD).....	121
Figure 6.1-1: Investigation matrix	129
Figure 6.1-2: Three-dimensional illustration of the ADMFB	132
Figure 6.1-3: (a) Shapes of the crushed and sieved coal (b) Visual analysis of coalification of a sample	132
Figure 6.1-4: Sampling procedure	133
Figure 6.1-5: Pressure drop relative to airflow for -5.6+4.75mm particles (VD).....	134
Figure 6.1-6: Particle segregation as per the ash yield for +11.2mm-13.2mm (DM 0:1 and VD)	135
Figure 6.1-7: Mass yield compared to bed height for +11.2mm-13.2mm (DM 0:1 and VD)	136
Figure 6.1-8: Ash yield compared to bed height for +11.2mm-13.2mm (DM 0:1 and VD)..	137
Figure 6.1-9: Cumulative ash yield compared to mass yield for +11.2mm-13.2mm (DM 0:1 and VD).....	138
Figure 6.1-10: Comparison of bed performance and washability curve (limiting performance curve) for +11.2mm-13.2mm (DM 0:1 and VD)	139
Figure 6.1-11: Comparison of bed performance for DM 0:1, DM 1:1 and DM 2:1 (+11.2mm-13.2mm and VD).....	140
Figure 6.1-12: Comparison of bed performance for DM 0:1, DM 1:1 and DM 2:1 (+11.2mm-13.2mm and VD).....	141

Figure 6.1-13: Comparison of bed performance for DM 0:1, DM 1:1 and DM 2:1 (+2.8mm-4.0mm and VD).....	142
Figure 6.1-14: Oscillating vibration motor (A) adjustments of weights (B and C)	143
Figure 6.1-15: Ash values for +4.75mm-5.6mm coal with vibration deactivated and activated (DM 0:1).....	145
Figure 6.1-16: Ash values for +4.75mm-5.6mm coal at 21.97 Hz	146
Figure 6.1-17: Ash values for +4.75mm-5.6mm particles at 30% amplitude	147
Figure 6.1-18: Ash values for +4.75mm-5.6mm coal particles at 19.86 Hz	148
Figure 6.1-19: Performance curves for +4.75mm-5.6mm coal particles.....	149
Figure 6.1-20: Performance curves for +2.8mm-4.0mm coal particles.....	150
Figure 7.1-1: Three-dimensional illustration of the ADMFB	159
Figure 7.1-2: Case 11 – complete mixing – fluidization velocity too high	164
Figure 7.1-3: Case 11 – complete mixing - vertical solid fractions of components	165
Figure 7.1-4: Case 11 – successful segregation.....	165
Figure 7.1-5: Case 11 – successful segregation - vertical solid fractions of components...	166
Figure 7.1-6: Case 11 – DM settling – fluidization velocity too low.	166
Figure 7.1-7: Case 11 – DM settling - vertical solid fractions of components.....	167
Figure 7.1-8: Case 11 – Segregation of 5mm coal fluidized with 0.3mm dense medium at various fluidization velocities.....	168
Figure 7.1-9: Case 23 – Complete mixing occurs over the entire range of fluidization velocities.....	169
Figure 7.1-10: Case 18 – DM settling in fluidized bed.	170
Figure 7.1-11: Case 19 – Particle segregation with 0.1mm DM particle size.	171
Figure 7.1-12: Case 20 – Particle segregation with 0.3mm DM particle size.	171
Figure 7.1-13: Case 6 – DM volumetric ratio to coal (25 : 75).....	172
Figure 7.1-14: Case 14 – DM volumetric ratio to coal (50 : 50).....	173
Figure 7.1-15: Case 20 – DM volumetric ratio to coal (75 : 25).....	173
Figure 7.1-16: Binary mixture (DM ratio magnetite to coal 92:8).....	176
Figure 7.1-17: Binary mixture (DM ratio of magnetite to coal 88:12).....	176
Figure 7.1-18: Binary mixture (DM ratio of magnetite to coal 75:25).....	177
Figure 7.1-19: Binary mixture (DM ratio of magnetite to coal 63:37).....	177
Figure 7.1-20: Particle Solid fraction versus bed-height	178
Figure 7.1-21: Particle Solid fraction versus bed-height	179
Figure 7.2-1: ADMFB CFD Case 1.....	184
Figure 7.2-2: ADMFB CFD Case 2.....	184
Figure 7.2-3: ADMFB CFD Case 3.....	185
Figure 7.2-4: ADMFB CFD Case 4.....	185

Figure 7.2-5: ADMFB CFD Case 5.....	186
Figure 7.2-6: ADMFB CFD Case 6.....	186
Figure 7.2-7: ADMFB CFD Case 7.....	187
Figure 7.2-8: ADMFB CFD Case 8.....	187
Figure 7.2-9: ADMFB CFD Case 9.....	188
Figure 7.2-10: ADMFB CFD Case 10.....	188
Figure 7.2-11: ADMFB CFD Case 11.....	189
Figure 7.2-12: ADMFB CFD Case 12.....	189
Figure 7.2-13: ADMFB CFD Case 13.....	190
Figure 7.2-14: ADMFB CFD Case 14.....	190
Figure 7.2-15: ADMFB CFD Case 15.....	191
Figure 7.2-16: ADMFB CFD Case 16.....	191
Figure 7.2-17: ADMFB CFD Case 17.....	192
Figure 7.2-18: ADMFB CFD Case 18.....	192
Figure 7.2-19: ADMFB CFD Case 19.....	193
Figure 7.2-20: ADMFB CFD Case 20.....	193
Figure 7.2-21: ADMFB CFD Case 21.....	194
Figure 7.2-22: ADMFB CFD Case 22.....	194
Figure 7.2-23: ADMFB CFD Case 23.....	195
Figure 7.2-24: ADMFB CFD Case 24.....	195

LIST OF TABLES

Table 2.3-1: General characteristics of South African ROM and coal products (IPC, 2016; SANEDI, 2011)	20
Table 2.3-2: ESKOM coal specification and rejection ranges (Steyn, 2009)	20
Table 2.4-1: Classification and characteristics of coal size fractions (SANEDI, 2011)	21
Table 2.4-2: Ep's of beneficiation techniques taken from Chikerema & Moys (2012).....	26
Table 2.5-1: Symbols for Fluidization Equations 2.1 to 2.8.....	31
Table 3.1-1: Separation results of various coal size fractions	71
Table 3.1-2: Average initial proximate and calorific values for each coal particle size range	72
Table 3.1-3: Degree of difficulty of washing of the coal samples	73
Table 3.1-4: Minimum (U_{mf}) and operating (U) fluidizing conditions for PSD-05, PSD-03 and PSD-01	76
Table 3.1-5: Ash, CV and density value comparison of PSD-05, PSD-03 and PSD-01	76
Table 3.1-6: Comparison of dense medium ratio of +8.0mm-9.5mm coal.....	78
Table 3.1-7: U_{mf} (m/s) and U (m/s) of PSD-05 (+11.2mm-13.2mm) and PSD-01 (+5.6mm-6.7mm) coal	79
Table 3.1-8: Comparison of minimum and operating fluidizing conditions for varying vibration	81
Table 3.1-9: Best cumulative results obtained from study	83
Table 4.1-1: Summary of experimental investigations on ADMFB, adapted from Mohanta <i>et al.</i> (2011).....	92
Table 4.1-2: Best cumulative results obtained from study	92
Table 4.1-3: Average initial proximate and calorific values for each coal particle size range	94
Table 4.1-4: Influence fluidization factor on ADMFB performance	102
Table 4.1-5: Cumulative ash yield and mass yield at a 50mm cut-point height (discarding only the bottom layer).....	107
Table 5.1-1: Separation results of various coal size fractions	116
Table 5.1-2: Range of average initial proximate and calorific values for feed coal.....	117
Table 5.1-3: Minimum (U_{mf}) and operating (U) fluidizing conditions for various PSD's.....	119
Table 5.1-4: Ash, CV and density value comparison of PSD-11, PSD-09 and PSD-07	120
Table 5.1-5: Comparison of dense medium ratio of PSD-11 coal (VD).....	122
Table 5.1-6: U_{mf} (m/s) and U (m/s) of PSD-11 and PSD-07 coal	122
Table 5.1-7: Comparison of minimum and operating fluidizing conditions for varying vibration (DM 0:1).....	123
Table 6.1-1: Average initial proximate and calorific values for each coal particle size range	130

Table 6.1-2: Total mass and ash balance for +11.2mm-13.2mm (DM :1 and VD)	135
Table 7.1-1: Average initial proximate and calorific values for each coal particle size range	158
Table 7.1-2: Properties of gas and particulate phases.....	161
Table 7.1-3: Parametric study variables	161
Table 7.1-4: ADMFB case definitions	161
Table 7.1-5: Binary mixture cases	175

LIST OF ACRONYMS

ADMFB	<i>Air Dense Medium Fluidization Bed</i>
AD	<i>Air dried</i>
AR	<i>As received</i>
df	<i>degrees of freedom</i>
ARD	<i>Apparent relative density</i>
BD	<i>Bulk density</i>
TRD	<i>True relative density</i>
ROM	<i>Run Off Mine</i>
SACPS	<i>South African Coal Processing Society</i>
SAWS	<i>South African Weather Service</i>
WCI	<i>World Coal Institute</i>
CFD	<i>Computational Fluid Dynamics</i>
IEA	<i>International Energy Agency</i>
SPI	<i>Standardized Precipitation Index</i>
Ep	<i>Ecart Probable</i>
DOE	<i>Department of Energy</i>
DDM	<i>Dry Dense Medium</i>
Hz	<i>Hertz</i>
L	<i>Length</i>
H	<i>Height</i>
W	<i>Width</i>
WEC	<i>World Energy Council</i>

POT *Pneumatic oscillating table*

SANEDI *South African National Energy Development Institute*

LIST OF SYMBOLS

cm	<i>centimetre</i>
CO₂	<i>carbon dioxide</i>
g	<i>gram</i>
H₂O	<i>water</i>
kg	<i>kilogram</i>
mm	<i>millimetre</i>
%	<i>percentage</i>
°C	<i>degrees Celsius</i>
%_w	<i>percentage by weight</i>
<i>C_d</i>	drag coefficient (dimensionless)
<i>H</i>	height above the distributor (m)
<i>I_k</i>	interphase momentum exchange (kg/m ² /s)
<i>M_B</i>	total mass of biomass (dimensionless)
<i>MI_{vol}</i>	volume based mixing index (dimensionless)
<i>N</i>	total
<i>P_k</i>	pressure of phase <i>k</i> (Pa)
<i>Re_p</i>	particle Reynolds number (dimensionless)
<i>U</i>	superficial gas velocity (m/s)
<i>U_{ki}</i>	mean velocity of phase <i>k</i> (m/s)
<i>V_{d,i}</i>	drift velocity (m/s)
<i>V_{rp,i}</i>	mean relative velocity gas and particle (m/s)
<i>X</i>	volume fraction of biomass (dimensionless)

d_p	particle diameter (m)
e_c	elasticity coefficient during inter-particle collision (dimensionless)
g	gravitational constant (m/s ²)
g_0	particle pair correlation function (dimensionless)
k	turbulent kinetic energy (m ² /s ²)
$m_{B,i}$	mass of biomass in i^{th} slice (dimensionless)
$m_{i,N}$	normalized mass of biomass in i^{th} slice (dimensionless)
m_p	mass of particle p (kg)
n_p	number of particles per volume unit (m ⁻³)
q_{fp}	fluid–particle fluctuating covariance (m ² /s ²)
q^2	particle fluctuating kinetic energy (m ² /s ²)

CHAPTER 1

INTRODUCTION

This report is titled “Dry beneficiation of small South African coal using an air dense medium fluidization bed”. Chapter 1 consists of Sections 1.1 to 1.4. Section 1.1 contains a brief background discussion of coal including a short discussion on the motivation for the need for beneficiation of coal in South Africa. This is followed by Section 1.2, which details the scope of the investigation. Section 1.3 identifies the main objectives of the study. Lastly, the outline of the project is given in Section 1.4.

1.1 Background and motivation

Coal has been an important energy-resource base worldwide for centuries due to its abundance, affordability and availability (WEC, 2022). Currently, coal is employed as fuel source that meets about 30 percent of the worldwide energy demand and will continue to do so for the next couple of decades (WEC, 2022). South Africa’s national energy-resource base is dominated by coal; with coal-fired power stations meeting about 83% of the maximum electricity generation in 2018 (DOE, 2019). South Africa’s need for coal-based energy is not likely to change drastically in the next 20 years, due to the lack of suitable energy alternatives. The South African economy is greatly dependent on coal as a source of foreign proceeds (Reddick, 2006). In addition to the extensive use of coal in the national economy, about 24% of South Africa’s coal contributed to the local export market in 2016, while only 1% reported to the import market (DOE, 2019). This coal exported from South Africa reaches various international merchants including China, India and Europe. Moreover, steel, iron and cement production sectors in South Africa also relies on coal, as reducing agent during metallurgical processes as well as a source of energy to operate processes (IEA, 2011). The coal mining industry is an important national establishment that provides thousands of South Africans with job opportunities. The South African mining industry as a whole was able to secure jobs for more than 485000 individuals in 2019, with 92000 reporting to the coal and lignite mining division (Minerals Council of South Africa, 2020). Therefore, by employing roughly 19% of the total mining workforce, the coal mining industry of South Africa was ranked 3rd largest national mining employer after the gold and platinum groups (Minerals Council of South Africa, 2020). Coal is a key source of energy

worldwide and from an economic stance, the sustainability thereof is of vital importance to South Africa.

According to Jeffrey *et al.* (2014), the persistent decline in accessible coal deposits is the main issue that is faced by prominent coal producing countries worldwide.

In Figure 1.1.1, a summary of South Africa's total coal production, total export, total national consumption as well as population growth is displayed.

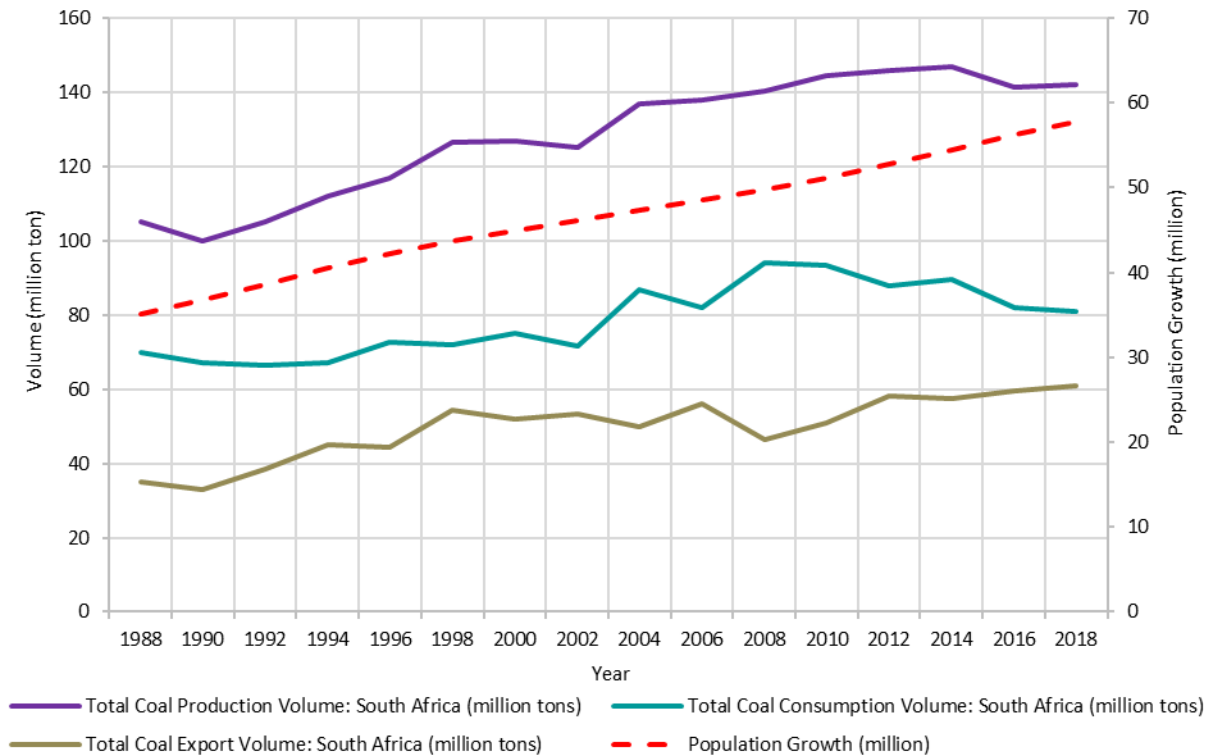


Figure 1.1-1: South African statistics taken from CEIC (2019)

From Figure 1.1.1, it is apparent the demand for coal will continue to increase along with the population growth. As such, the focus of the coal processing industries worldwide has shifted from quantity to quality and as a result, efficient processing, in an aid to secure an energy sustainable future.

Due to the inherently heterogeneous nature of coal, it is usually necessary to attempt to separate some of the impurities from the coal prior to consumption. By doing this, a coal product of higher overall quality is produced which curbs environmental penalties and limits energy losses (Chen & Yang, 2003). This reduction process is known as coal beneficiation and is achieved by a broad range of mechanisation techniques and various water treatment practices. Currently industrialized, wet beneficiation processes are known to produce high

separation efficiencies however, as the name implies, it involves water – substantial amounts at that. Difficulties in handling, transport and storage as well as increased capital, operating and transportation cost result from excess water retained in coal (Choung *et al.*, 2006). Furthermore, in water scarce coal producing and processing countries, including Mongolia, Australia and South Africa, suitable water supply is limited and as a result, wet coal beneficiation techniques is not ideal (Chen & Yang, 2003; Zhao *et al.*, 2015).

South Africa's water resource base is currently strained and upon further exploration, it was found that the average yearly rainfall is $\pm 490\text{mm}$, which is significantly lower than the worldwide average of $\pm 980\text{mm}$ (Rand Water, 2017). South Africa is characterized by irregular rainfall patterns and periods of abrupt, intense drought (SAWS, 2017). Figure 1.1.2 shows a map of South Africa showing the rainfall pattern change for the past 5 years - January 2017 to January 2022.

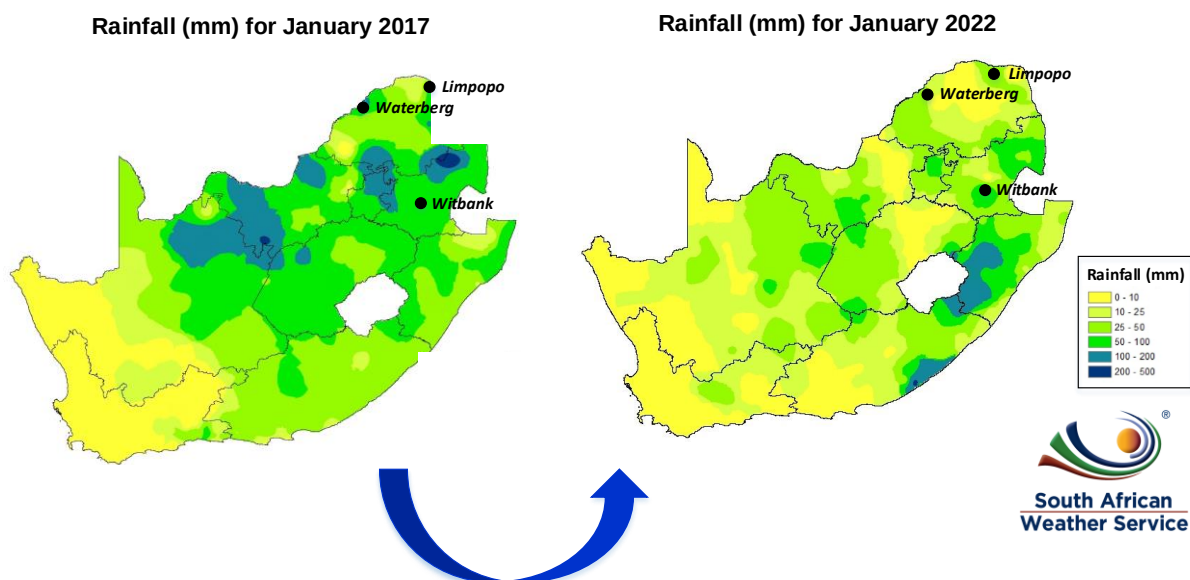


Figure 1.1-2: South African rainfall pattern change taken from SAWS (2022)

A notable reduction in rainfall is visible in the northeast portion of the country, which is South Africa's coal rich regions. From this, it is apparent that South Africa has endured periods of inconsistent rainfall, which brings concern especially to industries relying on water for day-to-day operation.

These undesired rainfall patterns, especially in the coal rich areas of South Africa and various other countries, calls for sustainable use of process water. As such, the industry has been researching dry coal beneficiation methods in the hopes of finding a suitable alternative

that is capable of producing comparably high separation efficiencies, without water consumption. Globally, a wide range of dry beneficiation techniques have been studied, including pneumatic oscillating tables, air jigging and air dense medium fluidized bed (ADMFB). These methods yielded efficiencies that fall within a narrow range with wet beneficiation methods (Chen & Wei, 2003; Dwari & Rao, 2007, Zhao *et al.*, 2015). Air dense medium fluidization (ADMF) has grown popular over recent years as an alternative dry beneficiation process and has successfully been implemented in China for +6mm coal particles (Mohanta *et al.*, 2011). Air dense medium fluidization is known to yield separation efficiencies that are in close range of wet beneficiation techniques, decrease capital and operating costs as well as lowering the levels of air and water pollution (Chen & Yang, 2003).

ADMF operates on the same principle as wet, density separation-based beneficiation processes. This process consists of a bed of dense medium particles, at a specific size range, through which air percolates from the bottom of the bed, at a velocity somewhat above the point of minimum fluidization. The air creates a dense medium particle suspension that behaves like a pseudo-fluid with a stable, uniform specific density. When coal particles are added to the dense medium suspension, segregation occurs in which the high-density coal particles sink to the bottom while the low-density coal particles float to the bed surface. In some cases, by applying parameters such as altering the amount of dense medium added per unit of coal, or vibration with various frequency and amplitude, fluidization is improved, and separation efficiency is increased. Therefore, it is possible to obtain a high-quality coal product that is suitable for export and a lower quality beneficiated product that still adheres to the local thermal coal market demands.

1.2 Aim and objectives

The use of fluidized bed technology in the industry has grown popular over in recent years. Fluidized bed uses include producing an array of light petroleum products from crude oil, boiler bed technology, granulator for urea granule production and last but not the least, coal beneficiation. The main aim of this study is to investigate the applicability of air dense medium fluidization separation technology to -13.2mm South African ROM coals.

The performance of the air dense medium fluidization coal separator was tested by considering various parameters when feeding -13.2mm South African ROM coals. The desire is to identify the best operating conditions in order to obtain a high-quality coal product that is suitable for export and/or a beneficiated fraction that adheres to local thermal coal market demands. The performance and process efficiency parameters were calculated to evaluate the beneficiation process.

Recent development in numerical techniques and computing efficacy has advanced computational fluid dynamics (CFD), used extensively in practice, to provide effective design solutions in fluidized bed industry. In addition to the main aim, this thesis includes a review of the current fluidized bed setup simulated with CFD software, which in this case, was done with the code NEPTUNE_CFD. The aim is to illustrate the bed function, identify flow patterns, track particle segregation and trends as well as identify areas where operational or structural alterations is required to the bed.

The following objectives were set out for this thesis:

1. A laboratory scale air dense medium fluidized bed was constructed and commissioned at the North-West University of Potchefstroom. Test work to investigate the feasibility of beneficiating +0.5-13.2mm South African ROM coal will be conducted on this laboratory scale fluidized bed. These experiments will provide insight into the bed performance and operating conditions, and any difficulties experienced will identified in order to further develop and improve this technology. The parameters that will be tested include coal particle size range, the effect of dense medium, vertical induced bed vibration as well as the effect of minimum fluidization.
2. Compare the bed performance by measuring the minimum fluidization velocity, ash yield, calorific value and density of the coal in each of the bed layers. The bed performance will be evaluated by comparing the results of the top layer to the feed values of the coal. The bed performance will also be evaluated by determining the particle segregation in the bed – this will also be done by comparing the results in the top and bottom layers of the bed. The most feasible and effective operating conditions will be identified from the test work.
3. Lastly, the movement and segregation of coal particles in the presence of dense medium as well as the solid-gas interaction in fluidized bed will be investigated by studying the dynamic mechanism in the fluidized bed. In order to achieve this, the experimental ADMFB geometry will be set up in NEPTUNE_CFD and a set of experimental runs will be simulated with various parameters. In this study, CFD will be employed to improve the understanding of the dense medium (DM) selection process for best particle segregation. The material properties and operational parameters that will be studied includes DM particle size, DM material density, fluidization velocity and DM to coal ratio. This portion of the study aims to identify the best suitable DM (size, density and volume ratio) from a binary mixture of magnetite and coal fines by systematically evaluating the separation performance of coal in an ADMFB in the presence of a DM with CFD simulation.

By employing such a numerical simulation, various underlying mechanisms including bubble formation, volume fraction and velocity patterns within the bed can be detected and observed. Mixing during fluidization and the degree of separation throughout the bed can be observed as well as present a screenshot of the state of the bed at any point in time. The economic viability of employing ADMFB as a beneficiation technique for coal fines is governed by the rate of segregation, the extend of segregation as well as the fluidization velocity. By simulating the ADMFB geometry and associated parameters with CFD, the economic viability can easily be verified ahead of engineering and construction.

1.3 Scope of study

The scope of study of this thesis is three-fold. This includes completing a literature review study on the theory relating to ADMFB technology and conducting a comprehensive set of pre-determined experiments on a lab scale batchwise ADMFB technology. Finally, a brief investigation into Eulerian Multiphase Model theory was conducted in order to study the movement and segregation of coal particles and solid-gas interaction in a fluidized bed (same geometry as the experimental setup) with NEPTUNE_CFD Software.

1.4 Thesis Structure

Each chapter deliberates on a specific part of the study as set out in the previous sections. The order and layout of the chapters are provided in the sections that follow.

1.4.1 General

Chapter 1: Introduction

Chapter 1 serves as a background and motivation supporting the intent of the thesis. This section of the thesis identifies and discusses main issues and problems related to the coal in industry and calls attention to the South African mining industry's demand for improved and cost-effective dry beneficiation technologies. Moreover, a proposal of an applicable dry beneficiation technique is discussed as well as the objectives to be addressed throughout the study is identified.

Chapter 2: Literature review

Chapter 2 is a literature review detailing the theory relating to ADMFB technology. The literature review touches on the historic development of coal and elaborates on mining and related industries in South Africa. Coal preparation and beneficiation in South Africa is deliberated on as well as Air Dense Medium Fluidized Bed (ADMFB) technology as a suitable alternative. ADMFB technology is meticulously described with emphases on the

fluidization mechanism. Subsequently, modelling of particle segregation and fluidization behaviour in a fluidized bed with computational fluid dynamics (CFD) is evaluated. Furthermore, case studies are discussed where the evolution, development and success of industrialized ADMFB technology is emphasized.

1.4.2 Papers

Chapter 3 to Chapter 5 contains papers published that investigated dry beneficiation by applying an ADMFB experimental setup to beneficiate +0.5mm-1mm, +1mm-2mm, +2mm-2.8mm, +2.8mm-4mm, +4mm-4.75mm and +4.75mm-5.6mm, +5.6mm-6.7mm, +6.7mm-8.0mm, +8.0mm-9.5mm, +9.5mm-11.2mm and +11.2mm-13.2mm South African coal. These particle size ranges were split into two papers and published according to progression of the research. The first paper titled “Dry beneficiation of small South African coal using an air dense medium fluidization bed” (Chapter 3) investigated dry beneficiation of +5.6mm-13.2mm South African coal in an ADMFB when subjected to various parameters and operating condition. The second paper titled “Dry beneficiation of +0.5mm-5.6mm South African coal using an air dense medium fluidization bed” (Chapter 4) investigated dry beneficiation of +0.5mm-5.6mm South African coal in an ADMFB when subjected to various parameters and operating condition. The third paper was also titled “Dry beneficiation of small South African coal using an air dense medium fluidization bed” (Chapter 5) investigated dry beneficiation of +0.5mm-13.2mm South African coal in an ADMFB when subjected to various parameters and operating condition. This paper included discussed the information included in the first and second paper and the content was presented at the XIX International Coal Preparation Congress (ICPC). The paper was subsequently published in the congress proceedings. Throughout the first three papers, one of the key parameters investigated was addition of DM to the fluidized bed process. The function of DM is to stabilize and enhance separation capabilities of the ADMFB system. However, the experimental results were not providing this effect and the decision was taken to further explore the DM selection process.

The fourth and final paper titled “Use of CFD modelling to assist with air dense medium fluidization of South African coal” (Chapter 7) evaluated the effects of material properties and operational parameters on the segregation of particles in an air dense medium fluidized bed (ADMFB) with the help of Computational Fluid Dynamics (CFD). In this study, CFD was employed to improve the understanding of the DM selection process for best particle segregation.

Chapter 3: ADMFB of +5.6mm-13.2mm South African Coal

This paper investigates the air dense medium fluidized bed (ADMFB) as a suitable dry beneficiation technique to upgrade +5.6mm-13.2mm South African coal. The separation efficiency of the ADMFB was assessed by determining the effect, if any, of particle size range, medium to coal feed ratios and the addition of vibration on the separation ability of the bed. The coal was upgraded to varying degrees, depending on the parameters of the applied methods.

Chapter 4: ADMFB of +0.5mm-5.6mm South African Coal

This paper presents the separation performance of solid phase bed material, at various particle size ranges (-5.6mm+0.5mm South African coal particles), in an air dense medium fluidized bed (ADMFB).

The coal particles were separated into +0.5mm-1mm, +1mm-2mm, +2mm-2.8mm, +2.8mm-4mm, +4mm-4.75mm and +4.75mm-5.6mm PSDs (particle size distributions) and fed to the fluidized bed in these fractions. Along with the six coal particle size ranges, three dense media to coal ratios and the addition of vibration was tested to identify the best operating conditions. The separation performance of the bed was measured by ash content (%wt) and calorific value (CV) as well as minimum fluidization velocity (U_{mf}).

Chapter 5: ADMFB of +0.5mm-13.2mm South African Coal

This study investigated the air dense medium fluidized bed (ADMFB) as a suitable dry beneficiation technique to upgrade small (+0.5mm-13.2mm) South African coal. The separation efficiency of the bed was assessed by determining the influence, if any, of particle size range, dense medium (magnetite) to coal feed ratios and the addition of vibration to the bed.

Chapter 6: Additional information (Paper 1 to Paper 3)

Chapter 7: Use of CFD modelling to assist with air dense medium fluidization of South African coal

Results of computational fluid dynamic simulation model for particle segregation in an ADMFB, applying the Euler- Euler n-fluid approach using of Neptune, is discussed in CHAPTER 6.

Chapter 6, the numerical portion of the investigation, is comprised of three aspects. Firstly, a parametric study to better comprehend the physical principles of an ADMFB and the influence of dense medium particle size, dense medium particle density, dense medium to coal ratio and fluidization velocity. Secondly, a parametric study is performed to examine the suitability of the dense medium made up of a binary mixture of fine coal and magnetite. Lastly, the evaluation of a dense medium binary mixture of fine coal and magnetite to efficiently separate a coal particle mixture.

Chapter 8: Conclusions and Recommendations

Conclusions and recommendations are provided in CHAPTER 7.

CHAPTER 2

LITERATURE REVIEW

This chapter starts with a brief overview and short description of the historic development of coal. Coal in South African is also discussed, specifically coal products and utilization thereof. Coal in South Africa is followed by coal preparation and beneficiation, which includes discussions on preparation, beneficiation as well as dry beneficiation techniques. This section serves to explain why coal requires intensive preparation and beneficiation before use in the coal industry. Also included in this chapter is the principle of fluidization and the various related facets. Dry beneficiation of coal using an air dense medium fluidization bed is discussed next and this includes all the parameters or conditions that affect fluidized bed performance. The literature review is concluded with a final section discussing CFD modelling to study particle movement in an air dense medium fluidization bed.

2.1 Overview

Coal is classified a fossil fuel and is composed of coal is an organic sedimentary rock formed through the biochemical transformation and then followed by sedimentation (physical compaction). As coal is heterogeneous of nature, removal of various impurities before consumption is warranted. By liberating the coal from impurities, a higher quality coal is obtainable, that has an increased calorific value and abides by to environmental emission legislation (Chen & Yang, 2003). The process by which impurities are removed from coal is referred to in the coal industry as coal beneficiation and integrates an extensive variety of wet and dry treatment practices. The desired coal portions are separated from the undesired coal portions during beneficiation (World Coal Institute, 2009). Wet beneficiation processes are currently industrialized; producing high quality separation with acceptable results, however, copious amounts of process water are required. Water affiliated industrial processes is accompanied by handling, storage and transportation difficulties as well as higher capital, operational and transportation overheads (Choung *et al.*, 2006). Added to this, coal-producing countries that are suffering from climate extremities, such as Mongolia and South Africa, have unpredictable or insufficient process water, making wet beneficiation practices in such countries impractical (Chen & Yang, 2003; Zhao *et al.*, 2015).

According to the SAWS (2017), with an average annual rainfall of approximately 490mm, half of the recorded global annual average, South Africa is considered a water scarce

country. Over the past decade, South Africa has suffered from climate extremities such as severe phases of draught and irregular rainfall (SAWS, 2017). This irregular rainfall pattern forecast emphasizes the significance of resourceful water usage, especially in the areas with major coalfields. Consequently, the South African coal industry is leaning towards dry beneficiation as a possible solution, as comparable results to wet beneficiation can be achieved without the added water constraint. A number of dry beneficiation techniques which includes pneumatic oscillating tables, air jigging and air dense medium fluidized bed (ADMFB), have been studied and yielded results falling within a narrow range from those produced by wet beneficiation techniques (Chen & Wei, 2003; Zhao *et al.*, 2015). Air dense medium fluidisation has grown popular as an alternate dry beneficiation technique and has been successfully employed for -6mm coal particles in China (Mohanta *et al.*, 2011). Air dense medium fluidization has delivered results that marginally compares to that of results obtained for wet beneficiation. By implementing this dry beneficiation technique, a reduction in capital and operational expenses can effectively be obtained as well as a reduction in air and water pollution (Chen & Yang, 2003). The principle of air dense medium fluidization is based on density separation, where the bed is composed of dense medium particles and air, as a fluidization medium, is driven through the bottom section of the bed, at a velocity above minimum fluidization. When air is introduced into the bed, the particles form a suspension in the air stream, adopting a fluid-like behaviour with a constant, stable density. Once coal is added to the suspension, the coal particles are stratified along the bed height according to density i.e., more dense coal particles sink to the bottom of the bed and less dense coal particles float at the top. This behaviour conforms to Archimedes' buoyancy principle and is essentially dependent on the coal particle's density, size and shape (Zhao *et al.*, 2015; Sahu *et al.*, 2013). Moreover, by altering the feed of dense medium or by adding vibration, fluidization improvement may be experienced which might yield improved results. The fluidization process can be optimized in this manner, attaining a portion of coal with elevated heating value and a clear-cut discard, at a certain height of the fluidized bed.

2.2 Historic development of coal

Coal is a fossil fuel that is created by compaction, densification and induration of various plant and animal remains (SACPS, 2015). This readily combustible sedimentary rock is formed over millions of years and is rich in carbon. Plant remains and debris, along with tectonic movements, cover swamps and peat sumps to substantial depths during the first coal age (Falcon & Ham, 1988). In this interment, decay period, the plant remains are exposed to severe temperatures and pressures that effect the physical and chemical

characteristics of the remains. These biochemical changes are the driving force in the process of converting peat into coal (Falcon & Ham, 1988).

Coal or coal seams consists of organic and inorganic materials; it does not have a crystalline structure and is not of homogeneous nature. Banded strata, known as coal seams, materializes as various types of coal, shale, sandstone sediments and intermittent conglomerates (SACPS, 2015).

2.1.1 Coal rank

The measure of the degree of alteration during maturation of coal is termed “coal rank” and is expressed at the hand of coal stages that range from lignite to anthracite (Falcon & Ham, 1988). Coal rank is based on the increase in carbon content and a decrease in volatile material. Figure 2.1.1, adapted from WCI (2009), shows the stages in which coal matures during coalification.

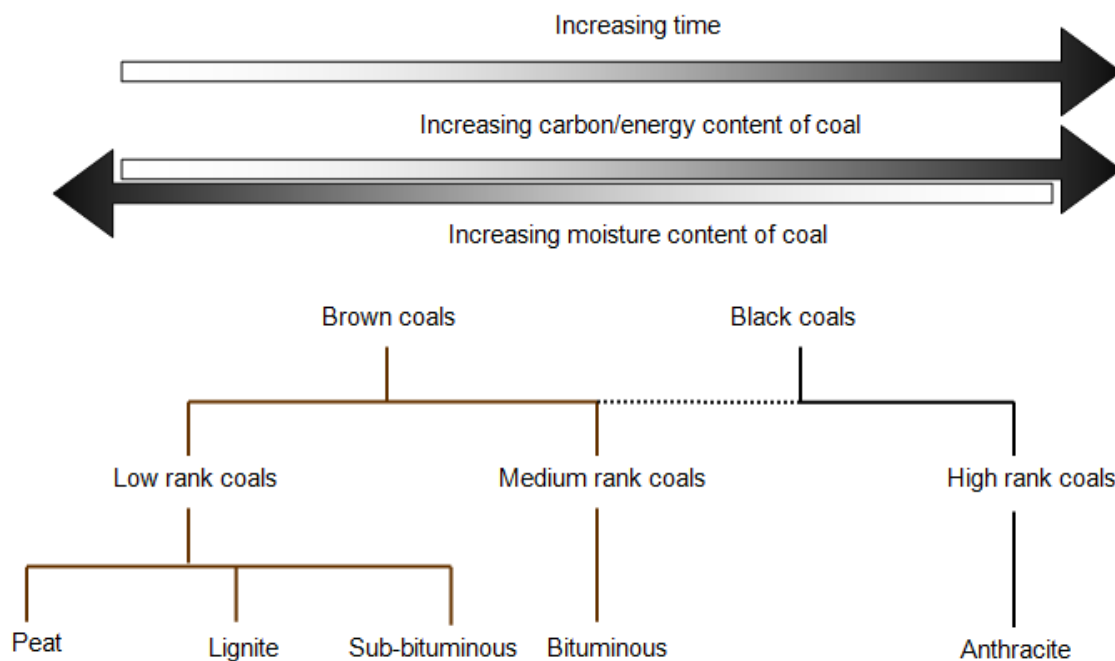


Figure 2.2-1: Stages of coal maturity adapted from WCI (2009)

During the first stage of coalification, lignite is formed. Initially, peat is formed swamp vegetation remains, which evolves into a soft brown coal over time, known as lignite (WCI, 2009). Falcon & Ham (1988) clarifies that with the influence of time, depth of interment as well as thermal exposure, the coal to starts to blacken and harden to form bituminous coal. Further progression in the coalification process yields the highest rank coal termed anthracite. As coal changes appearance throughout the specific coal ranks, the coal also

matures on a physical and molecular level. In essence, the molecular structure of the coal leans toward a purer organic material with a reduced overall porosity fraction (Falcon & Ham, 1988). Low rank coals with a higher degree of porosity brings about elevated levels of surface moisture, and a reduction of carbon (Petrick, 1969).

2.1.2 Lithotypes

At the coal interface, on a macroscopic level, bands of different textures and brightness are present. These distinct bands are referred to as lithotypes, classified as either vitrain, durain, clarain or fusain, which is a visual depiction of the coal (SACPS, 2015):

- **Vitrain:** Present in hard bituminous and anthracitic coals as thin layers of jet-black lustre and soft brittle grain.
- **Durain:** Catagorized into black and grey durain. Black durain is composed mostly of leaves and seed remains. Grey durain resembles black durain but is essentially a combination of vitrain and fusain coal.
- **Clarain:** Alternate bands of vitrain and black durain that presents itself as thin bands with a satin appearance.
- **Fusain:** Hard continuous bands that occurs as discrete flattened pieces of coal at various depths of a coal seam. This coal type represents oxidized plant matter that was formed during the first stages of coalification.

2.1.3 Coal macerals

Macerals are fractions of coalified plant remains conserved in coal and rock formations. Macerals is generally categorized according to three groups: vitrinite, inertinite and liptinite (Falcon & Ham, 1988).

Vitrinite macerals are fragments of matured plant cell walls. Falcon & Ham (1988) stated that the aromaticity and carbon content of vitrinite macerals is known to increase diagnostically with rank. The waxy and resinous remains of plant material form *liptinite macerals*. These plant fragments, which includes cuticles, spores, and resins, are impervious to weathering. In its advanced form, coalification is rarely seen in liptinite macerals, as these macerals start to dissolve in coals of medium volatility and are seldom present in coals of low volatility. Liptinite macerals are often referred to as “plant fossils” as they have a tendency to preserve its original vegetation state. *Inertinite* macerals are formed during the peat stage of coal formation from plant remains that underwent alteration, and degradation. Inertinite macerals endure minute structural changes with increasing rank. This

type of maceral has the highest reflectance level, making it more easily identifiable from other macerals (SANEDI, 2011).

Surface moisture on coal particles typically resides on the surface of macerals. Macerals can also be classified according to maceral state of reactivity. Vitrinite and liptinite are reactive macerals, whereas inertinite is branded unreactive (Falcon & Ham, 1988).

2.1.4 Coal characterization and mineralogy

Due to coal's inherently complex combination of organic and inorganic matter, accurate and complete classification is largely reliant on numerous chemicals, physical and thermal properties. These properties include the macerals present, moisture and mineral contents as well as the associated density and heating values. Collectively, these features have a large impact on the quality of coal and greatly influences its behaviour during processing. High quality coals are characterized by high calorific values, low-intermediate densities, low moisture levels as well as reduced portions of mineral matter content. The mineral matter content of coal can be defined as inert solid constituents that ultimately adds to the dilution of coal quality. The mineral matter contents do not ignite during combustion and subsequently forms the grey ash residue (ash yield) that is visible after combustion (SACPS, 2015). Mineral matter is classified into inherent and extraneous types. Inherent mineral matter consists of solid minerals that were intrinsically combined in the coal during seam formation. Extraneous mineral matter generally consists of layers of dirt in the coal seam in addition to sandstone, intermediary rocks and shale that forms part of the mined product (SACPS, 2015). Coal grade is established by the amount and type of mineral matter present and varies according to coal seam and rank. Various environmental and technological issues associated with coal preparation and utilization is related to ash yield (Schweinfurth, 2009). It is important to note that, especially for beneficiation plants, mineral matter, or ash yield, has a significant effect on coal density. The density of mineral matter is, by nature, higher than that of organic matter and as such, the type and amount of minerals present significantly influences coal density. Xie (2015) notes that for every 1% rise in coal ash yield; a 0.01% increase in coal density is expected. As a result, the behaviour of coal particles in beneficiation processes, such as dense medium separation, is greatly affected

2.1.5 Other properties of relevance

In addition to the ash yield or mineral matter content of a coal, moisture content, volatile matter and fixed carbon content plays an important role in the behaviour of coal. The

majority of coal mining operations make use of water, which often results in coal with high surface (free and capillary) moisture levels. Moisture that is retained in coal has a negative effect on the quality of a coal, much like ash content. Evaporation of moisture retained in the coal requires additional heat energy, which reduces the efficiency of combustion. Therefore, moisture plays an important role in the performance of coal in processing techniques. Another component that is of relevance to this study is volatile matter. Volatile matter consists of a combination of long and short chain hydrocarbons as well as sulphur and is released at high temperatures in the absence of oxygen. Volatile matter is responsible for the release of undesirable carbon-, sulphur- and nitrogen-oxide production associated with the process of combustion, making dealing therewith key (Schweinfurth, 2009). The fixed carbon content or the pure carbon yield is essentially the difference in mass of a sample of coal after the volatiles, moisture and mineral matter fractions have been liberated. These components of coal are quantified by ultimate and proximate analyses. Components are driven off individually during these analytic processes by high temperatures and determines each of the component amounts by difference in mass.

2.1.6 Density

Density of coal is of utmost relevance and importance for the study at hand. It is an indication of the geographic structure and nature of the coal and aids in defining the grade of coal. The mineral matter in coal, as previously mentioned, are associated with its density. Because of the diverse nature of coal, a large density profile is usually present throughout coalfields and coal seams. Density can be categorized according to, apparent relative density (ARD), true relative density (TRD), and bulk density (BD). TRD refers to the density of the coal, excluding the pores, and is typically used in quality research, by pycnometer analysis. Secondly, BD equates to the ratio of coal mass filling a container to the volume of the container. Finally, the ARD is the ore mass per unit volume, including any voids or pores. Ultimately, TRD typically presents the maximum density followed by ARD and then BD (Xie, 2015).

To summarize, coal is particularly heterogeneous in nature and, as a result, can be categorized according to different elements. The sections that follow discuss coal seams in South Africa with reference to coal formation and quality, including rank, grade, type, density and composition.

2.3 Significance of coal in South Africa

South Africa is heavily reliant on coal, with its national energy resource base currently dominated by coal. Accounting for $\pm 36\%$ of the global fuel consumption for electricity production, coal remains the most widely used primary fuel (DOE, 2019). Nationally, coal provides for about 77% of South Africa's primal energy requirements, which is unlikely to change in the next 20 years owing to a relative lack of suitable alternatives (DOE, 2019). The majority of national coal deposits is currently economically exploited which continues to strengthen South Africa's large coal-mining industry. In addition, roughly 28% of South Africa's coal is exported, primarily shipped through the coal terminal at Richards Bay, placing South Africa fourth among the world's largest coal exporting countries (WEC, 2022).

Approximately 51% of South African coal is mined underground, with the remainder being produced by opencast methods. The South African coal-mining industry is extremely concentrated, with 85% of saleable coal production being distributed between five companies (Minerals Council of South Africa, 2022). More than 20% of the run-of-mine coal produced locally is exported, 21% is consumed in the local market (excluding thermal coal) and the remaining unsaleable fraction is discarded. Coal preparation in South Africa, particularly for export purposes, results in $\pm 65\text{Mt}$ of coal discards being produced per annum (Minerals Council of South Africa, 2022).

By international standards, South Africa has comparatively shallow coal deposits with thick seams, which makes them more accessible, easier and cheaper to mine. At the present production rate, more than 50 years of coal supply should be left. However, as the energy demand is on a continual escalation rate, the coal production rate in South Africa is expected to remain a pressing situation for years to come. Figure 2.3.1 indicates the locations of the major coalfields in South Africa.

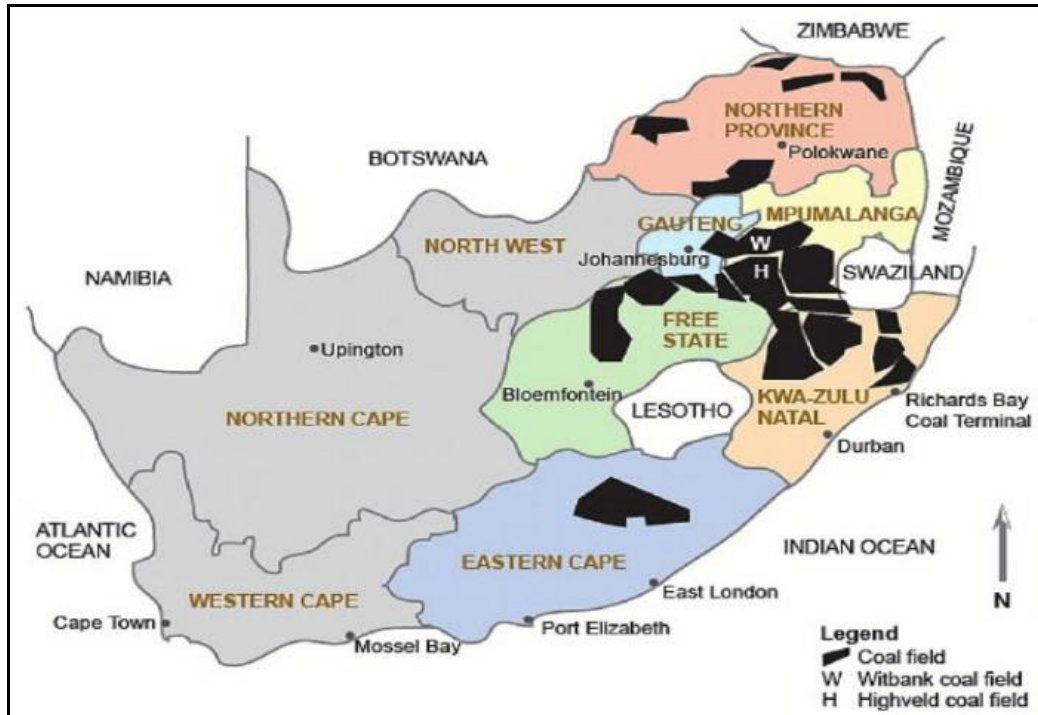


Figure 2.3-1: Main coalfields in South Africa, taken from SACPS (2015)

The coalfields in South Africa are shown in black on the map in Figure 2.3.1, with the Highveld and Witbank coalfields represented by H and W, respectively. Coal from the Witbank coalfield was selected for use during this investigative study and is characterized as semi-bituminous to bituminous rank, with reasonable calorific values and ash yields of $\pm 35\%_{\text{wt}}$ (Jeffrey, 2005).

2.3.1 Coal products and uses

Coal employed in various South African sectors namely, metallurgical processes, synthetic fuel preparation, electricity generation and international trade and commerce. These applications require a specific coal quality, with specific calorific value, ash yield, volatile matter, sulphur content and moisture levels. South Africa's coal grades are typically classified into four main categories, grade A, grade B, grade C and grade D. Some characteristics of the four grades of South African ROM coal are summarized in Table 2.3.1 (IPC Coal, 2016; SANEDI, 2011).

Table 2.3-1: General characteristics of South African ROM and coal products (IPC, 2016; SANEDI, 2011)

Characteristic	ROM	Grade A	Grade B	Grade C	Grade D
Sulphur (%)	<2.0	<0.6	<0.6	<1.0	<1.5
Calorific value (MJ/kg)	16-21	27-28.5	26-27.5	24-26	21-23
Ash (% _{wt})	20-40	12-14	14-16	16-18	20-25

It is apparent, with regards to the characteristics in Table 2.3.1, that the quality decreases from grade A to D, with the coal being of moderately low quality. Metallurgical coal, employed in the steel production sector for coking, as well as export coal, requires low sulphur and ash yields with a calorific value of >23MJ/kg (Eberhard, 2011). When considering the characteristics listed in Table 2.3.1, it is evident that beneficiation is necessary to achieve the required quality coal from the ROM coal. Thermal coal quality used for electricity production and synthetic fuel production is considerably lower. The ROM coal grade and product specification does not dictate the requirement and extent of beneficiation. Instead, economic feasibility related to operation and transportation influences the requirement and extent of beneficiation. Resultantly, various coal-mining operations yields a dual product that consists of beneficiated coal for export and metallurgical purposes, including a middling product fit for domestic consumption (SANEDI, 2011). Coal-mining operations in the Witbank area (producing coal used in this investigation) are in the number 1, 2, 4 and 5 seams. The coal produced from this seam is known to be of high-quality export, metallurgical, local steam and blended coke coal, respectively (SANEDI, 2011). The coal sample employed during this investigation originates from Witbank seam 4, which is a high-quality thermal coal (grade D), as this study is focussed on the South African thermal sector.

Table 2.3.2 provides the ESKOM coal specification and rejection ranges. These specifications, except for variations on ash and sulphur content, applies to all operational coal-fired power plants in South Africa (Steyn, 2009).

Table 2.3-2: ESKOM coal specification and rejection ranges (Steyn, 2009)

Parameter	Units	ESKOM	Rejection
Calorific Value (CV)	MJ/kg (AD)	≥21	20
Total Moisture	Maximum % (AR)	10	12
Ash	Maximum % (AR)	25-33	>35

Sulphur	Maximum % (AR)	1	2
Volatiles	Maximum % (AR)	≥ 20	< 20
Abrasiveness Index	Maximum	500	550

According to Steyn (2009), if coal is mined and supplied as ROM coal to Eskom, they generally pay at the bottom of the price band. When ROM coal goes through a specific beneficiation cycle, the price is raised.

2.4 Coal preparation and beneficiation

Beneficiation processes are important in eliminating various issues arising from the amount of extraneous and incombustible material in coal. Some of these issues include a decrease in calorific value; an increase in the volume of coal fines handled and transported; increased depreciation on equipment and the release of particulate emissions and harmful gases. Removal of the unwanted impurities from ROM coal to generate a uniform, suitable product is called coal preparation and beneficiation (SANEDI, 2011).

2.4.1 Preparation

Raw coal that is unprocessed (straight from the mine) is referred to as run-of-mine (ROM) coal. This coal mass is usually a combination of different fragment sizes and is known to contain impurities such as rock and dirt that are essentially unwanted. However, coal consumers require coal of a consistent quality. Refining of ROM coal to ensure a consistent quality product, that is more suited for specific end-uses, is referred to by the industry as coal preparation (Anglo American, 2016). The treatment that is required is subject to the properties of the coal as well as the intended use thereof. In some cases, simple crushing is sufficient whereas in other instances complex treatment processes to reduce impurities may be required. Impurities is removed from the ROM coal by first crushing and then separating the loot into the various size fractions. Larger material fractions are usually treated using dense medium separation whereas smaller size fractions may be treated in various ways, typically based on mass differentials, such as in centrifuges. The size designations, mass distributions (%_{total product}) and moisture contents (%_{weight}) of coal in South Africa is provided in Table 2.4.1.

Table 2.4-1: Classification and characteristics of coal size fractions (SANEDI, 2011)

Classification	Particle size (mm)	Mass distribution (%_{total product})	Moisture content (AR %_{wt})
Coarse	12-250	61	5

Small	1-12	28	9
Fine	0.1-1	7	13
Ultra-fine	0.1-0.15	4	27

Coal particles, presented in Table 2.4.1, between the sizes of 5.6mm to 13.2mm are deemed as part of the scope for the study. These sizes fall within the small coal classification, which is associated with 28% of the total coal mass distribution and low moisture levels of 9%_{wt}.

2.4.2 Coal beneficiation

Beneficiation processes are implemented in the preparation section of coal processing to minimize the quantity of minerals present in coal and produce coal that adheres to product specifications. Density separation techniques are useful to employ in coal processing due to the large density differential between the mineral matter content and the fixed carbon content of the coal particles (SANEDI, 2011). Density separation methods essentially separate minerals with a range of specific gravities, by employing particle relative movement in response to gravity and additional forces. Particles experience resistance to motion because of the movement of a viscous fluid (water or air) (Wills & Napier-Munn, 2008). Based on the particular medium, beneficiation techniques can be classified as dry (air as medium) or wet (water as medium) techniques; however, both techniques employ the same principles. In dense medium separation, ore or coal particles of varying density are added to a medium of fixed density and depending on individual particle density, the ore or coal particles will sink or float. The particles with a density higher than the density of the medium will sink and the particles with that are less dense will float (Wills & Napier-Munn, 2008). Coal with a high mineral matter content (ash forming minerals) present is denser; dense medium separation to remove ash-forming minerals is possible (SANEDI, 2011). Figure 2.4.1 illustrates the principle of dense medium separation.

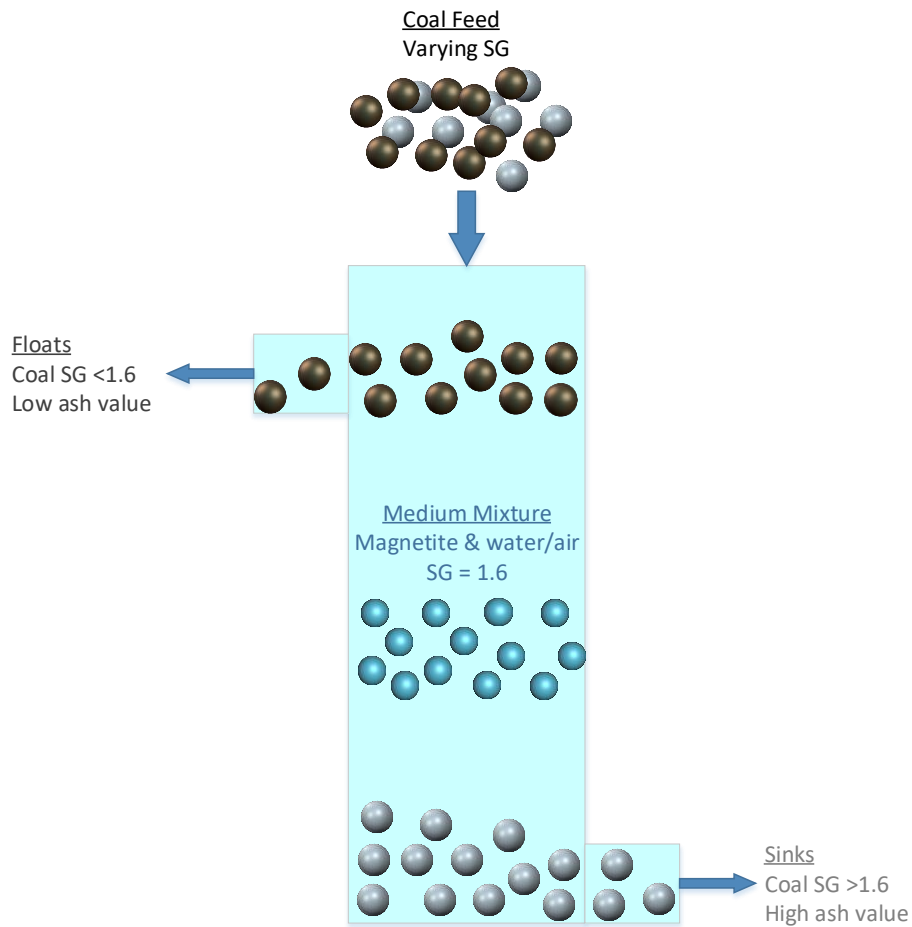


Figure 2.4-1: Principle of dense medium separation

A coal sample, with variable density, shown in Figure 2.4.1, is placed into a container/vessel containing a dense medium with a density of 1.6g/cm^3 (reference density). The coal that is denser than the reference density, sinks and the coal particles less dense than the reference density will float. Effective separation is only possible if there exists a clear differential between the desired ore and gangue densities. Moreover, the influence of particle size on gravity separation should be considered, as the efficiency is likely to reduce when particle movement is governed by surface friction instead of gravity. This phenomenon usually arises when finer particle fractions are at play (Wills & Napier-Munn, 2008). Wet dense medium separation techniques have frequently been employed, on a global scale, for the beneficiation of coal are based on wet beneficiation (Yang *et al.*, 2013). The coal feed is washed by means of spirals, cyclones, froth floatation and dense medium drums, all of which are known for extensive water requirements that yields a moisture-laden product with a lower heating value. Because of this, wet beneficiation processes usually call for better-developed infrastructure for water recovery and tailings disposal (Yang *et al.*, 2013).

As previously reported by De Korte (2004) and Reddick (2006), between 2002 to 2004, about more than 80% of the total coal mined in South Africa was beneficiated. However, according to Prevost (2010), this value has declined to 60% in 2007.

Irrespective of efficiency, wet coal beneficiation remains incredibly water intensive, involving large volumes of water that is from a reliable source and readily available. Securing such water sources to operate a beneficiation plant may be exhaustive in terms of both capital and operational expenses (Reddick, 2006). Figure 2.4.2 represents a coal washing circuit of a colliery in South Africa.

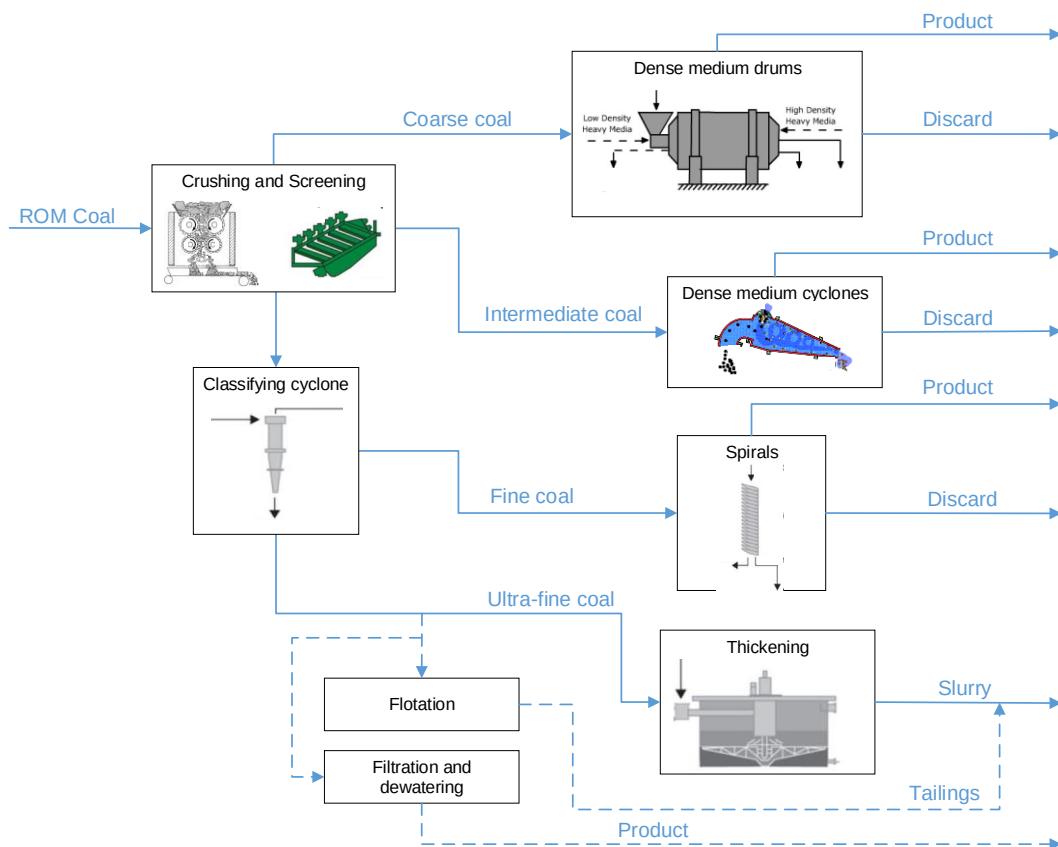


Figure 2.4-2: Typical colliery coal washing circuit adapted from SANEDI (2011)

After foreign material has been separated from the raw, ROM coal, it is crushed into smaller, more manageable, size fractions. The crushed coal is then separated and classified, by screening, according to the requirement and design of the client/colliery. The coarse coal, as shown in Figure 2.4.2, is then sent to be washed in dense medium drums to separate the desired low ash coal from the undesired high ash coals (Reddick, 2000). The intermediate coal fraction is directed to the dense medium cyclones, where it is washed and separated according to density. Small coal is separated into fine and ultra-fine fractions in a classifying cyclone, after which the fine coal fraction is beneficiated in spiral concentrators, and the

ultra-fine coal is sent to a flotation unit (Nicol, 1992). Alternatively, the ultra-fine coal can be sent to a filtration and dewatering section or discarding in slime dams. Water that is salvaged from the thickener is recycled to the beneficiation plant. The ultra-fines leave the bottom of the thickener in slurry with the process water that has remained in the thickener (Reddick, 2000). Finally, water from the slime dams is also recovered and recycled to the beneficiation plant.

2.4.3 Dry beneficiation techniques

Numerous dry coal beneficiation techniques have been established in the coal washing market for the past century. These dry beneficiation techniques include hand-picking, x-ray sorting, air tables and pneumatic separator, air jigging, specialized air jigging (FGX), electrostatic separation, fluidized dry dense medium separation (Chen & Wei, 2003; Dwari & Rao, 2007, Zhao *et al.*, 2015). Each of these techniques manipulate the variances between coal and gangue properties in order to generate separation between the desired material and undesired material. These properties include particle size, density, shape, electric and magnetic conductivity, radioactivity as well as frictional coefficient, to mention a few (Chen & Wei, 2003).

Dry coal processing has grown popular in recent years among various countries, especially in the USA and China (Honaker *et al.*, 2008; Chen & Wei, 2003). The reasoning behind this can be attributed to the reduction in capital and operational cost, elimination of costly water treatment procedures and providing coal with an increased calorific value and lower moisture content. However, since 1990, a reduction in this trend was experienced due to the increased moisture levels in ROM coal and increasingly strict dust control requirements. As a result, the demand for better coal qualities at acceptable separation efficiencies became crucial to the industry (Honaker *et al.*, 2008). Despite some countries losing interest, large-scale research of dry beneficiation techniques continued to some extent, as the benefits proved to be favour-some in countries experiencing drought and extreme climate fluctuations around the globe. Various fields of study were undertaken and it was found that studies employing pneumatic beneficiation methods, such as air jigging and ADMFB, proved to be suitable solutions (Chen & Wei, 2003). As with any beneficiation processes, a number of difficulties, including restrictions on feed particle size, reduces separation efficiency values, high airflow requirements and air pollution, accompanies pneumatic processes (Chen & Yang, 2003). Of these methods, air dense medium fluidization proves to be highly capable of producing a product of favourable quality, accompanied by the least number of difficulties. The Ecart Probable Moyen (E_p) values for various coal beneficiation techniques are provided

in Table 2.4.2. This parameter is used to effectively measure the extent to which separation is achieved by a beneficiation technique. A lower E_p value indicates better separation has occurred, with zero indicating ideal separation. Chikerema & Moys (2012) completed separation tests with various beneficiation methods for different particle size ranges which ranged from +9.5mm-16mm, +16mm-22mm, +22mm-31.5mm and +37mm-53mm.

Table 2.4-2: E_p 's of beneficiation techniques taken from Chikerema & Moys (2012).

<i>Method</i>		<i>Ecart Probable Moyen (E_p) range</i>
Dry	Air jigging	0.200 - 0.300
	FGX	0.150 - 0.300
	ADMFB	0.040 - 0.150
Wet	DMS	0.015 - 0.120

When comparing the values in Table 2.4.2, it is evident that ADMFB technology produce results that falls within closest range from wet dense medium separation methods (Chikerema & Moys, 2012). Therefore, the scope of this study is centred, primarily, on dry beneficiation of coal using air dense medium fluidization (ADMFB).

2.5 Fluidization

In principle, fluidization is the process by which solid particles in a packed bed adopts a fluid like behaviour by subjecting the particles to an upward flowing gas or liquid stream, creating a suspension of particles. This beneficiation technique generates numerous characteristics, some of which can prove beneficial to the user (Kunii & Lievenspiel, 2001). Air dense medium fluidized beds, that operate on the fluidization process, was selected as dry beneficiation technique for this study and is detailed in the subsequent sections.

2.5.1 Fluidization principle

Fluidization is based on Archimedes principle, which states:

“A body, whether fully or partially submerged in a fluid, is subjected to an upward buoyancy force equivalent to the weight of the fluid the body displaces.” Archimedes (c. 250 BCE).

Fluidization is categorized into three states/stages namely, fixed state, minimum fluidization and turbulent fluidization, depending on the behaviour of the particles in the system. During fixed state, the particles remain in their original positions and minimal to no movement is present. The upward flowing fluid permeates the empty spaces between the particles, leaving the particles unhindered. When the velocity of the upward flowing fluid is increased slightly, the bed expands, allowing the particles to lift from their original positions, and at a specific velocity, become suspended. This is the incipient point of fluidization and is termed

the minimum fluidization stage. The final stage is the stage where the particle in the bed is subjected to a high upward fluid flow, with the system entering a turbulent flow region where vigorous particle and fluid movement takes place (Kunii & Levenspiel, 2001).

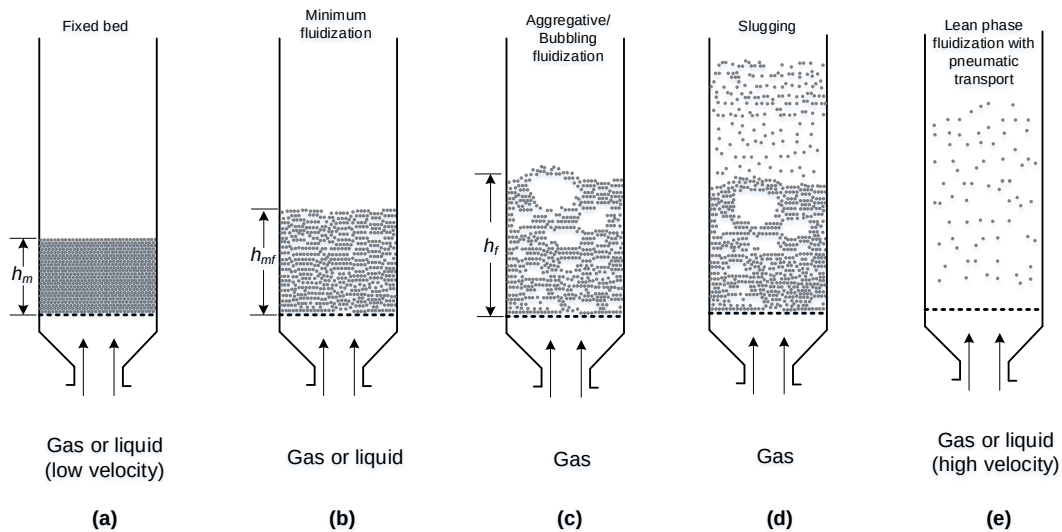


Figure 2.5-1: Stages of contacting during fluidization, adapted from Kunii & Levenspiel (2001)

At fluidization velocities above the minimum fluidization, irregularities and differences in particle behaviour for gas-solid and liquid-solid systems is evident. In liquid-solid systems, a consistent and progressive bed expansion is obtained, and flow instabilities are damped, remaining reasonably small. Gas-solid systems are usually challenged by bed instabilities with bubbles and channels forming when the gas velocity is operated beyond the minimum fluidization. As the gas flow rate increases, the particle in the bed starts to move and vibrate more vigorously, however bed expansion is limited to that observed at minimum fluidization, shown in Figure 2.5.1. Figure 2.5.1 illustrates the stages of fluidization. When the gas/fluid flow rate is increased even more, the terminal velocity of the particles will reach its critical point and eventually be exceeded, and the upper surface of the bed disappears. It is at this point that entrainment is observed, which, along with bubble formation is undesirable. When increasing the gas flow beyond this point, a fluidized bed where particles are removed along with the fluid from the system is created (Kunii & Levenspiel, 2001). Gas-solid fluidization systems may also encounter coalescence of gas bubbles, resulting in bubbles to multiply in size while rising throughout the bed. When the dimensions of the bed vessel are of sufficient depth and diameter, these bubbles may coalesce to the point where it covers the entire diameter of the vessel. This causes the particles within the bed to respond differently than the behaviour expected in a normal bed setting. Fine particles flow down smoothly along the vessel wall surrounding the rising gas void/bubble, resulting in axial slugging. With regards

to a coarse particle bed, the portion directly above the extended gas bubble is forced to rise which results in particles falling back through the bubble void, eventually rupturing and disintegrating the bubble. Continual replication of this process creates an unstable oscillatory motion that is not ideal for effective fluidization (Kunii & Levenspiel, 2001). An unusual, yet valuable, property of the gas-solid fluidized bed is its pseudo-fluid like behaviour, which became quite useful in dry beneficiation processes such as ADMFB. The system can be compared to boiling water with its fluid-like behaviour (Kunii & Levenspiel, 2001), which is explained at the hand of Figure 2.5.2.

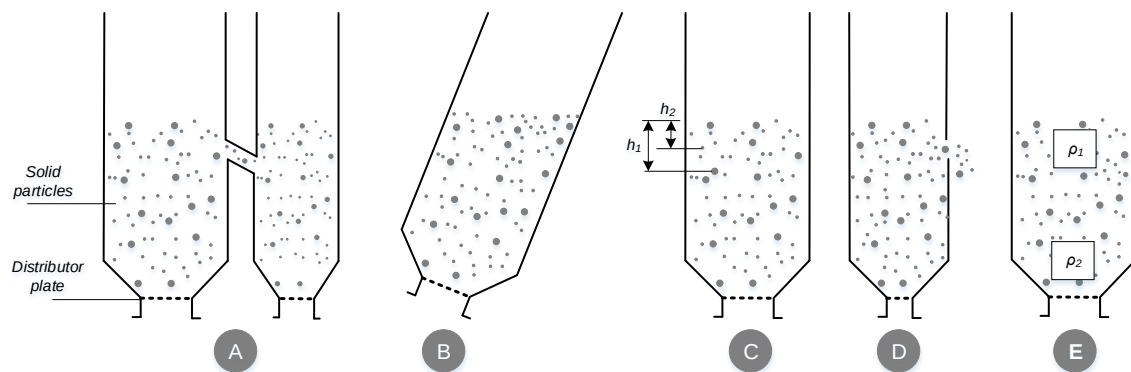


Figure 2.5-2: Pseudo fluid properties, adapted from Luo & Chen (2001); Mohanta *et al.* (2011)

It should be noted (not portrayed in Figure 2.5.2), that when a large, less dense particle is added to a stable fluidized system, the particle will rise to the top and float on the surface of the bed. Diagram A and B in Figure 2.5.2 shows that the particle levels of two fluidized beds that are connected would balance out and when the system is sloped, the upper level of the fluidized bed will remain horizontal and even out, respectively. Diagram C indicates that a pressure differential between two points in the fluidized bed system is equivalent to the static head between the same points. Diagram D shows that the solids inside the bed would eject from a hole in the side of the vessel due to the pressure differential. Moreover, diagram E indicates that, over time, the particles in the bed experience density segregation by which more dense particles moves to the bottom of the bed and the less dense particles migrates to the bed surface (Mohanta *et al.*, 2011). These properties are all generic liquid properties that presents themselves in a gas-solid fluidized bed system.

An adequate, strong fluidized bed, that has no turbulence or extensive bubble formation, is observed at a fluidization velocity point slightly above minimum fluidizing velocity. At the point of minimum fluidization, particles start to move around because of several forces acting on them, and of which, buoyancy and gravitation are the most dominant (Rhodes, 2008). The fluid like behaviour adopted by the particle suspension in the system gives rise to density segregation along the fluidized bed system (Sahu *et al.*, 2009). A bed of particles in

this state behaves in the same way as a Bingham fluid with rheological characteristics typically explained by the falling sphere principle (Wei & Chen, 2001). The external forces that act on a particle in a fluidized bed are described in Figure 2.5.3 (Sahu *et al.*, 2009)

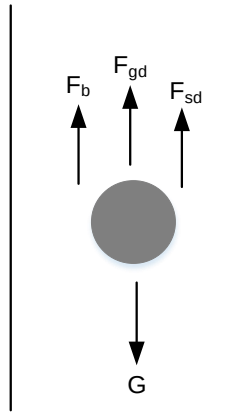


Figure 2.5-3: External forces acting on a particle during fluidization, adapted from Sahu *et al.* (2009)

The fluid drag force (F_{sd}) on the particle is usually negligibly small as the diameter of the dense medium particles used is much smaller than that of the particle suspended in the fluid. Therefore, there are three primary forces acting on a suspended particle namely, gravity (G), Buoyant force (F_b) and frictional drag force (F_{gd}). In practice, the assumption can be made that ideal separation can be reached by ideal particle distribution; however, in reality the possibility exists that light and dense particles can both be misplaced. For example, a heavy, denser particle that should descend floats at the surface of the bed and similarly, a light less dense particle is displaced to the bottom of the bed. This misplacement of particles is known as the displaced distribution effect (Chen & Wei, 2005; Sahu *et al.*, 2009). Numerous factors can contribute to displaced distribution effect; however movement and viscosity plays a predominant role. The displaced movement effect is most noticeable in the presence of an upward gas flow that is either too high or too low; therefore, the feed air velocity plays a significant role in the separation efficiency of the system (Chen & Wei, 2005). It is also apparent that particle diameter, bears pronounced effect on the displaced distribution theory. Research have shown that less dense particles, with a larger diameter, can sink while more dense particles, with reduced diameter, have the ability to float. Because of this, gas velocity remains a vital parameter in guaranteeing stable fluidization as well as effective segregation within the bed. Therefore, the velocity of the fluidization gas must be controlled and remain slightly above minimum fluidization point, however below bubbling velocity.

2.5.2 Minimum fluidizing velocity

The minimum fluidization velocity is essentially the point at which particles in packed bed state starts to fluidize. The particles adopt a fluid-like behaviour and are unrestricted, so they can move in the bed according to its density relative to the density of the entire fluidized bed (He *et al.*, 2016a; Zhao *et al.*, 2015). The minimum fluidization points or incipient point is therefore crucial to the stability of the fluidized bed and the separation efficiency of the fluidized bed operation. The drag force of a particle is directly proportional to the velocity required for fluidization as the drag force is responsible for solid particle suspension within the bed (Escudero & Heindel, 2011). Kunii & Levenspiel (1991) derived an equation commonly used to estimate the incipient fluidization or minimum fluidization point in a fluidized bed system. Some important parameters influencing the minimum fluidization velocity include particle properties, fluid properties and bed geometry (Escudero & Heindel, 2011).

Kunii and Levenspiel (1991) notes that the point of minimum fluidization is reached when the drag force of the upward flowing fluid is equivalent to the weight of the particles (Equation 2.1).

$$\text{Drag Force} = \text{Weight of Particle} \quad \text{Equation 2.1}$$

$$\Delta p_b A_t = W = A_t L_{mf} (1 - \varepsilon_{mf}) [(\rho_s - \rho_g) g g_c] \quad \text{Equation 2.2}$$

Equation 2.2 can be rearranged to represent the minimum fluidisation condition:

$$\frac{\Delta p_b}{L_{mf}} = (1 - \varepsilon_{mf}) (\rho_s - \rho_g) \frac{g}{g_c} \quad \text{Equation 2.3}$$

Minimum fluidising velocity or the superficial velocity at minimum fluidization conditions (U_{mf}) are, is given by

$$\frac{1.75}{\varepsilon_{mf}^3 \varphi_s} \left(\frac{d_p u_{mf} \rho_g}{\mu} \right)^2 + \frac{150(1 - \varepsilon_{mf})}{\varepsilon_{mf}^3 \varphi_s^2} \left(\frac{d_p u_{mf} \rho_g}{\mu} \right) = \left(\frac{d_p^3 \rho_g (\rho_s - \rho_g) g}{\mu^2} \right) \quad \text{Equation 2.4}$$

With

$$\text{Reynolds number } (Re) = \left(\frac{d_p u_{mf} \rho_g}{\mu} \right) \quad \text{Equation 2.5}$$

$$\text{Archimedes number } (Ar) = \left(\frac{d_p^3 \rho_g (\rho_s - \rho_g) g}{\mu^2} \right) \quad \text{Equation 2.6}$$

By substituting Equation 2.5 and 2.6 into Equation 2.4, Equation 2.4 can be rearranged to form Equation 2.7.

$$\frac{1.75}{\varepsilon_{mf}^3 \varphi_s} Re_{p_{mf}}^2 + \frac{150(1-\varepsilon_{mf})}{\varepsilon_{mf}^3 \varphi_s^2} Re_p = Ar \quad \text{Equation 2.7}$$

By simplifying Equation 2.7, the minimum fluidisation velocity (U_{mf}) of individual particles can be determined (Kunii & Levenspiel, 1991):

$$U_{mf} = \frac{d_p^2 (\rho_s - \rho_g) g}{150 \mu} \frac{\varepsilon_{mf}^3 \varphi_s^2}{1 - \varepsilon_{mf}} Re_{p_{mf}} < 20 \quad \text{Equation 2.8}$$

Equation 2.8 called the Ergun equation, and it shows that an increase in particle size (d_p) and particle density will call for a quadratic increase in minimum fluidization velocity. From Equation 2.8 it is also evident that the minimum fluidization velocity (U_{mf}) is proportional to the bed voidage (ε_{mf}). The location of a particle in the fluidized bed is largely dependent on the relative velocity between the particle and fluidizing medium. This plays an important role in the separation efficiency as it is largely reliant on the bed viscosity and medium velocity. Emphasis is placed accurately determining the minimum fluidization velocity as improper fluidization conditions can easily lead to the displacing effect, resulting in unsuccessful separation.

The symbols for the equations discussed in this section is provided in Table 2.5.1.

Table 2.5-1: Symbols for Fluidization Equations 2.1 to 2.8

Symbol	Description	Unit
Δp_b	Pressure difference	Pa
A_t	Cross-sectional bed area	m ²
W	Bed weight	N
H_{mf}	Bed height	m
ε_{mf}	Bed voidage	-
ρ_s	Particle density	kg/m ³
ρ_g	Fluid density	kg/m ³
g	Gravitational acceleration constant	m/s ²
U_{mf}	Minimum fluidization velocity	m/s
d_p	Particle diameter	m
ϕ_s	Particle sphericity	-
μ	Fluid viscosity	Pa.s

2.5.3 Pressure and velocity relations

superficial velocities ranging from $1.2U_{mf}$ to $2.0U_{mf}$. He *et al.* (2016a) was able to establish that some segregation did occur at $1.3U_{mf}$ due to slight bubble formation; however, by further increasing the gas velocity from $1.3U_{mf}$ to $1.8U_{mf}$, efficient separation was achieved as sufficient small bubble formation had occurred. Optimal separation was achieved at $1.6U_{mf}$. Further increasing the gas flow velocity to $1.8U_{mf}$ subjected the particles in the bed to severe back mixing and lead to displacement of particles that effectively reduced the separation efficiency. This is because the stability of the bed was compromised due to large bubble formation within the bed during operation (He *et al.*, 2016a). Zhao *et al.* (2016) conducted a similar study, where +1-3mm coal and 232 μ m magnetite was fed in a 4:1 volume ratio to a fluidized bed and the results obtained were somewhat different from that obtained by He *et al.* (2016a). The research conducted by Zhao *et al.* (2016) showed that back mixing of particles in the bed only started at gas velocities greater than $2.4U_{mf}$. This may be attributed to the differing coal and magnetite particle size ranges employed by the respective studies. From research, there no specific guidelines documented for gas velocity with reference to separation efficiency. It should be noted that the extent to which the operating velocity (U) should exceed U_{mf} is relative to particle size range, making this operation case specific.

2.5.4 Bubble formation and behavior

From the section 2.4.3, He *et al.* (2016a) and Zhao *et al.* (2016) notes the importance of accounting for both the operating gas velocity and minimum fluidization velocity during fluidized bed operation. The operating gas velocity and minimum fluidization velocity governs the process of bubble formation, which is critical for good separation in a gas-solid dense medium fluidized bed. Nevertheless, bubbles have benefits and drawbacks on effective and stable fluidization of particles in a bed. The following should be taken into account when operating a fluidized particle bed (Luo *et al.*, 2012; Sobrino, 2009):

- Bubble formation may lead to remixing. During initiation, mixing particles in the bed is required as it enhances density distribution and uniformity in the bed, which is generally driven by small bubble formation. However, large bubbles can cause remixing of particles that can effectively hinder particle stratification due to particle displacement and entrainment.
- Moreover, small bubbles moving through the bed brings about a “loosening effect” on the particles (particles are able to relocate more freely), thereby improving separation. Large bubbles are undesirable and negatively affects the bed as it creates an unstable bed by causing significant fluctuations in bed height. While smaller bubbles have a favourable effect on the fluidization process and encourages

well-developed fluidization, larger bubbles are known to spur a reduction in bed stability, effectively suppressing segregation (Luo *et al.*, 2012).

2.5.5 Classification of particles

Another factor that should be considered during fluidization is minimum fluidizing velocity that is essentially dependent upon particle properties, bed and particle geometry as well as the properties of the fluidizing fluid (Escudero & Heindel, 2011). Therefore, particles have been categorized according to four specific classes, namely Geldart A, B, C and D particles (Kunii & Levenspiel, 2001). The Geldart powder classification diagram is provided in Figure 2.5.5.

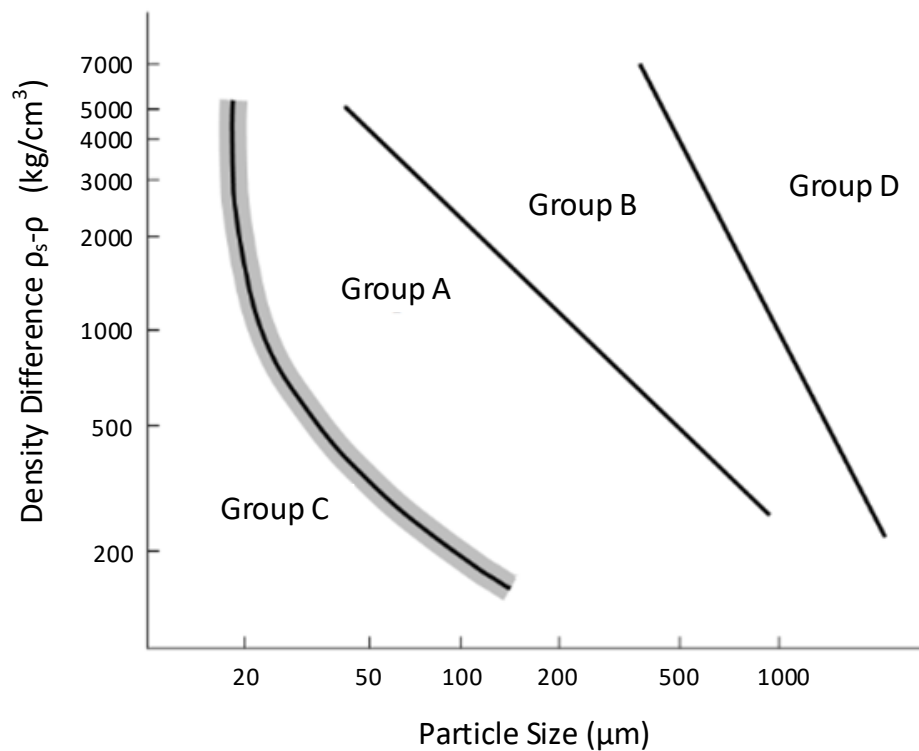


Figure 2.5-5: Geldart powder classification diagram taken from Richardson *et al.* (2008)

- **Group A**

These particles are fluidized with ease and bubble formation is relatively controllable at low gas velocities. Group A particles are, for the most part, characterized by particle densities of less than 1.4 SG and small average particle sizes.

- **Group B**

Group B particles is characterized by vigorous bubbling during fluidization because of the large density range of 1.4-4.0 SG as well as its small particle size (40-500 micron).

- **Group C**

Group C particles are very fine, making fluidization of the bed almost impossible. This is because the particle-gas interaction forces are weaker than the particle-particle interaction forces.

- **Group D**

Group D particles are very large and/or dense which disturbs the fluidization process due to large bubbles formation and eruption. Severe back-mixing and gas channelling appears in this group during fluidization (Yang *et al.*, 2015).

The feed coal employed in the scope of this investigation can be classed into the Geldart A/D types according to particle size range and is anticipated to fluidize easy without dense medium. The magnetite dense medium is Geldart B particles and because of their size, vigorous bubbling is expected which calls for close monitoring during operation.

2.6 Dry beneficiation of coal using ADMFB technology

The first dry coal beneficiation study conducted with an ADMFB can be traced back to 1926 when a paper was published on findings by Yancey and Frazer, after which the process was patented. The research was based on a small-scale fluidization process of +10-50mm coal and sand (SG of 1.45) as a dense medium (Houwelingen & de Jonge, 2004; Mohanta *et al.*, 2013). The process was however, not verified on either pilot or industrial scale and as a so the results obtained remained largely questionable. Dry beneficiation using air dense medium fluidized beds have since been explored on a universal basis for numerous applications (Zhao *et al.*, 2011). Separation of +6- 50mm coal particles by dry fluidization techniques has been investigated by various research institutions worldwide, yielding favourable results with E_p values close to 0.05 (Luo & Chen, 2001; Luo *et al.*, 2003; Wei *et al.*, 2003). The first ADMFB plant (pilot scale) was built by the Chinese University of Mining and Technology (CUMT) and was able to process about 320 000 tonnes of +6- 50 mm coal per annum, successfully separating the coal with an E_p of 0.05. Subsequently, large-scale commercial plants have been assembled with reported yearly throughputs of 700 000 tonnes (Chen & Wei, 2005). The DDM (dry dense medium) plant located in Shenhua, China is given in Figure 2.6.1.



Figure 2.6-1: DDM (dry dense medium) plant located in Shenhua, China (Kunii & Levenspiel, 1991).

Supplementary studies directed by the Chinese University of Mining and Technology (CUMT) concerning dry beneficiation of coal produced the following conclusions:

2.6.1.1 Air dense medium fluidized beds with the addition of vibration

Chen and Wei (2005) were able to reduce the ash yield of +0.5-6.0mm coal by 8.22 percentage points and bring about a cumulative mass yield of 80.2% by employing an ADMFB together with vibration. Moreover, an average E_p value of 0.065 was obtained for the experimental runs conducted.

- **Deep air dense medium fluidized beds**

Chen and Wei (2005) also discovered that by employing deeper beds, an even lower E_p value of 0.02 is obtainable when beneficiating -50mm coal. The high-quality separation achieved was particularly beneficial to areas that calls for removal of waste from feedstock such as China's large surface mines.

- **Triboelectric cleaning technology**

Triboelectric cleaning technology was employed to reduce the ash yield of ultra-fine coal with an average particle size of 43 microns. This technology returned exceptionally low product ash yields of about 2%_w. This value is, however, a result of coal liberation before beneficiation that added to removal ash forming minerals and subsequently lowered the ash yield (Chen & Wei, 2005).

- **Dual-density air dense medium fluidized beds**

Dual-density air dense medium fluidized beds is able to produce a threefold product; namely, a clean coal fraction, middling fraction and discard coal. Chen and Wei (2005) documented clean and discard coal densities of 1.52g/cm^3 and 1.87g/cm^3 with respective E_p values of 0.07 and 0.1.

2.6.2 Dense medium

The dense medium used in dense medium separation (DMS) processes is usually a dense particulate mineral suspended in a fluidizing agent such as air or water for dry and wet beneficiation techniques, respectively. A density profile is obtained over the depth of the bed. Because of the density differential between the clean coal, gangue and specific gravity of the dense medium, gravity separation takes place. Besides this, addition of dense medium to the system can assist in eliminating particle back mixing in the bed (Prusti *et al.*, 2015) and prevent vigorous bubbling (Fan *et al.*, 2001). Such as magnetite, the dense medium material selected in DMS operation usually exhibit strong magnetic properties. This is because recovering and cleaning of the dense medium by magnetic separation processes is easily achievable (Maoming *et al.*, 2003). Ultra-fine magnetite particles are known to fluidize poorly as the inter-particulate cohesive forces between the particles are particularly strong resulting in either plug flow or dead zones during operation (Luo *et al.*, 2002; Zhao *et al.*, 2010b). When the gas velocity is increased sufficiently, extreme bubbling may occur and eventually fine material will be directly blown out of the bed (De Korte, 2016; He *et al.*, 2016a). On the other hand, a dense medium that is too large will also exhibit problems during fluidization and the operation may call for increased gas velocities, resulting in a more energy intensive process. Moreover, the bed voidage would most likely be larger creating even larger bubbles (He *et al.*, 2016b). Another effect that should be taken into account is the difference in particle diameter between the coal and magnetite particles. The diameters of the coal and magnetite should not fall within a narrow range from one another as this will increase the strength of the inter-particulate surface forces and hinder fluidization. He *et al.* (2016a) investigated the effect of three different magnetite sizes on minimum fluidization, pressure drop and fluidized bed stability. The magnetite particles are classified as Geldart B powders with the following size categories: - 0.3+0.15mm, -0.15+0.074mm and -0.074mm. Figure 2.6.2 indicates how the pressure drop in the bed changes along with the change in superficial gas velocity for the magnetite particles.

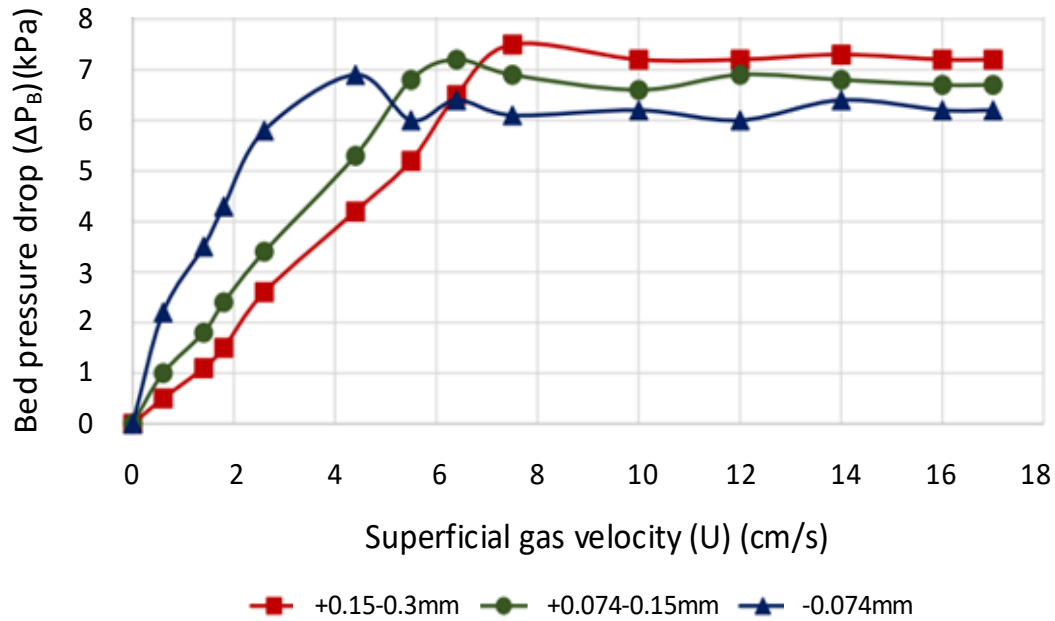


Figure 2.6-2: ΔP across a fluidized bed as a function of superficial velocity adapted from, He et al. (2016) (a)

Figure 2.6.2 shows that the -0.074mm medium yielded the lowest minimum fluidization point. It is also apparent that the minimum fluidization velocities decreased from - 0.3+0.15mm to - 0.074mm. The pressure drop plateaued once minimum fluidization was reached; however it is evident that the curve stability improved with increasing magnetite size. Furthermore, the conclusion was made that a bed with reduced bulk density, higher porosity and increased expansion was generated when finer magnetite was used – all which added to creating an unstable bed (He *et al.*, 2016a). The finer dense medium was more sensitive to step changes in the feed gas velocity. Essentially, it was concluded that -0.074mm magnetite produced an unstable fluidized bed and is not adequate to employ as a fluidizing medium. The -0.3+0.15mm magnetite performed well, delivering a more stable fluidized bed.

Magnetite in South Africa can be obtained as a by-product from the tailings of copper production but is very fine compared to the grade of magnetite used in the pilot plant operation in China (De Korte, 2016). Figure 2.6.3 portrays a size analysis for general South African magnetite compared to the magnetite found in China.

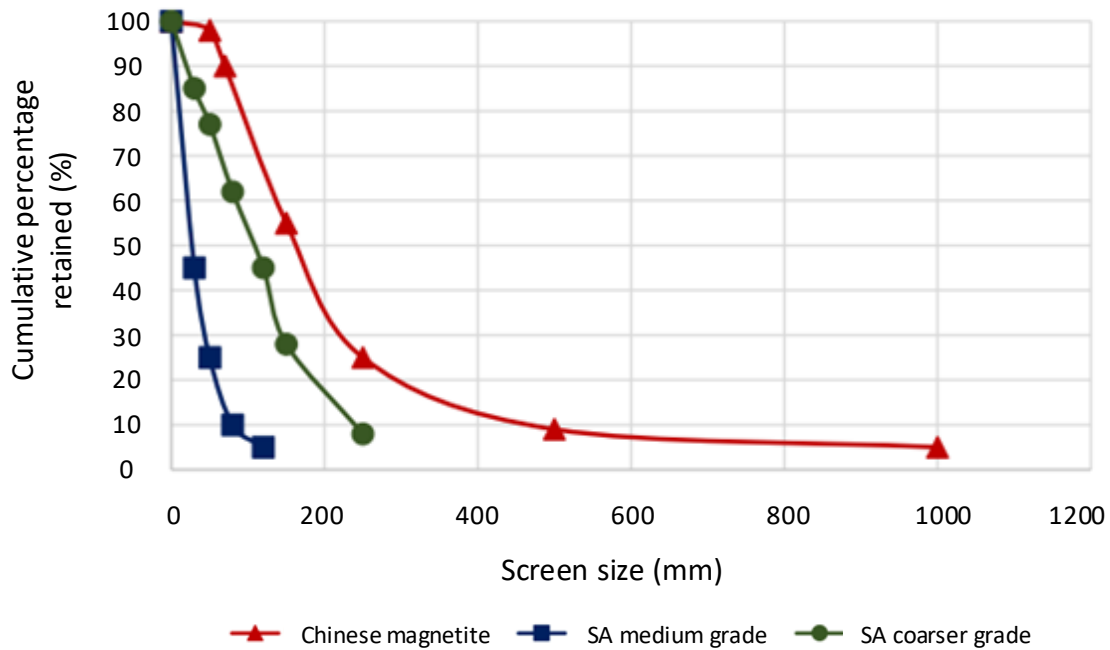


Figure 2.6-3: Comparison between magnetite sizes from three sources (adapted from de Korte (2016))

The curves in Figure 2.6.3 illustrates cumulative mass percentage retained against screen size for South African medium and coarse grade magnetite as well as magnetite used in China. The available magnetite in South Africa is very fine compared to the grade of magnetite employed in China, as can be seen from the results in Figure 2.6.3. The lack of coarse magnetite in South Africa is a serious obstacle that stands in the way of implementing ADMFB technology, a dire situation that is in need of action (De Korte, 2016). Separation of dense medium and coal after fluidization can be somewhat challenging and because of this, the consumption of dense medium can be quite high (Sahu *et al.*, 2009). Moreover, the process of recovering magnetite from the coal can be expensive (capital and operational) and can lead to product deterioration (Yang *et al.*, 2013). Particle attrition may cause fine coal accumulation in the dense medium, which affects the consistency of the medium density. The agglomerated coal needs to be removed completely in order to reset the medium density. Research by He *et al.* (2016c); Luo and Chen (2001) and Yang *et al.* (2013b); have shown some promising results when mixing the finer magnetite with slightly coarser coal. This mixture, also known as an autogenous medium, can provide an alternative solution for the constraint on the availability of coarse magnetite in South Africa. In addition, by using an autogenous medium, difficulties with recovering and cleaning dense medium can be more easily resolved.

2.6.3 Vibrated air dense medium fluidized beds

Stratification of coal particles is mainly a result of the presence of bubbles in the fluidized bed. Bubbles rising through the fluidized bed creates instability for a moment in that specific bed section, which induces a lower solid volume than the surrounding area. A particle in this unbalanced region will be able to travel without restrictions, and so light, less dense particles have the ability to move to the top of the bed (float) and denser particles sinks. This movement of particles in the bed, also known as stratification/segregation, can be enhanced by the addition of vibration. When vibration is added to the bed, a loosening effect is induced and bubble formation is improved that leads to improved segregation (Yang *et al.*, 2013a). Additionally, vibration creates a well dispersed and developed dense medium fluidized bed, as the gas to solid interactions are amplified (He *et al.*, 2015). Channelling is also decreased when vibration is added to the system, especially when operating with smaller particle size ranges (Moaming *et al.*, 2003). Macpherson and Galvin (2010) found that vibration could aid in reducing the point of minimum fluidizing; improve bed stability and reduce forces acting on the particles. The pressure drop of a dense medium fluidized bed with vibration was also found to be lower than that of a bed without vibration (He *et al.*, 2015). The results of the study conducted by Macpherson and Galvin (2010) are depicted in Figure 2.6.4.

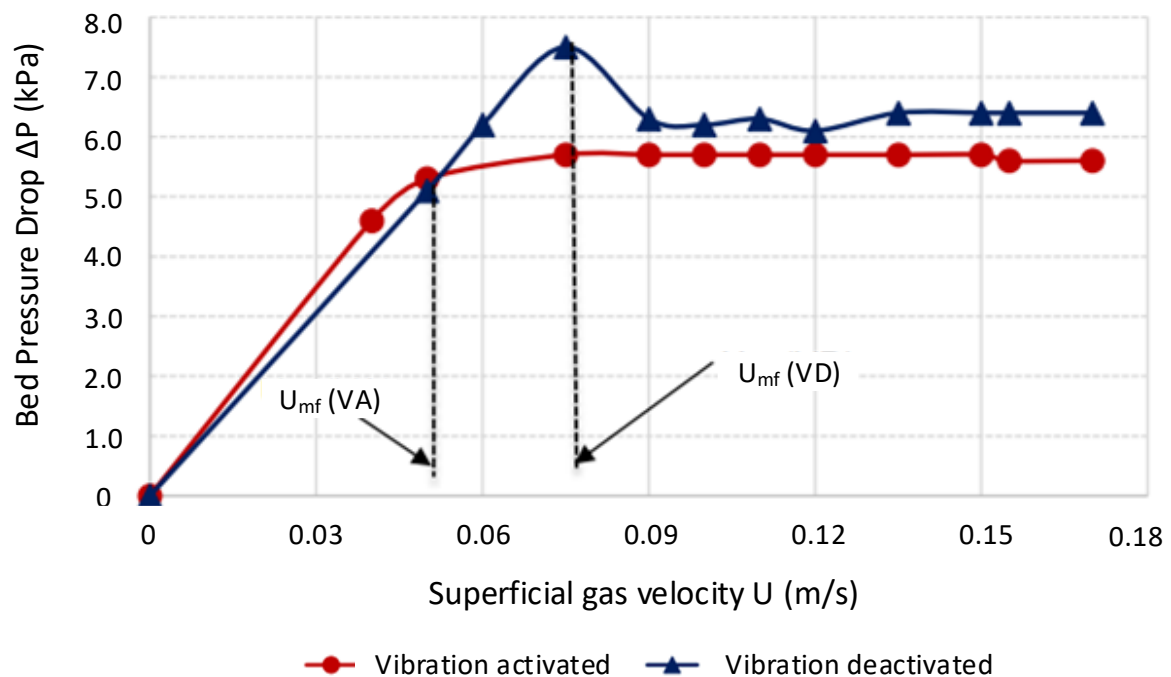


Figure 2.6-4: ΔP fluctuation of the bed with and without vibration (adapted from He *et al.* (2015))

The change in bed pressure according to superficial gas velocity are illustrated in Figure 2.6.4, for a bed with vibration and without vibration. $U_{mf} (VA)$ and $U_{mf} (VD)$, in Figure 2.6.4, indicates the points of minimum fluidisation for the bed with and without vibration,

respectively. Vibration provides a reduced, more stable pressure drop across the bed that reaches minimum fluidization velocity at lower superficial gas flows. The conditions and effects of adding vibration to an air dense medium fluidized bed operation have been the subject of numerous investigative studies. Yang *et al.* (2013a) studied the performance of -3+1mm and -6+3mm coal when vibration was added to the fluidized bed. Both particle size ranges delivered promising results when vibration was added to the system. The ash yield of the feed coal was reduced by 15%_{wt} from an average of 40%_{wt}, producing a gangue product containing an ash value of $\pm 70\%$ _{wt} (Yang *et al.*, 2013a). A study conducted by He *et al.*, (2015) on fine coal yielded good results with a reduction of $\pm 28\%$ _{wt} ash content to a coal product with ash content of $\pm 10\%$ _{wt}. These results were accompanied with high yields and E_p values between 0.065-0.055, subjected to particle size. Luo *et al.* (2008) was able to deliver coal with an ash yield of less than 10%_{wt} and E_p value of 0.07 by using a dense medium fluidized bed with vibration. Similar results were obtained by Dwari & Rao (2007) when fluidizing 125 μ m ultra-fine coal in a laboratory bench scale setup. The bench scale vibratory fluidized bed setup was able to generate a yield of $\pm 80.2\%$ and induce an ash decrease of 8.22% points.

Even though studies show favourable results, various parameters influence the operation of a vibratory fluidized bed, including frequency and amplitude. If the vibration frequency and amplitude is too small, the induced vibration effect will be too weak, and formation of dead zones and big bubbles may occur. Contrarily, if the frequency and amplitude is too large, circulation and back mixing of particles may occur in the bottom section of the bed (Yang *et al.*, 2013b). He *et al.* (2015) confirmed that the amplitude of vibration induces the largest effect on fluidization and segregation of the coal particles in the bed. Therefore, the frequency and amplitude should be suitably adjusted to maintain bed uniform bed stability and density during fluidized bed operation. It was established by He *et al.* (2015) that, for -6+1mm coal and -0.3+0.074mm magnetite, as dense medium, an amplitude and frequency of 1mm and 15Hz, respectively, was most effective. Based on the same principle, Yang *et al.* (2013b) found that frequencies of 18 to 33Hz and amplitudes of 2.6 to 3.8mm was optimal for operation, however it should be noted that different particle size ranges were employed. From the adverse results, the conclusion can be made that the ideal vibratory fluidization is largely coal and bed dependant.

2.6.4 Effect of moisture

Ineffective separation can take place when either or both the coal and dense medium particles contain high moisture levels. This is due to particle agglomeration and adherence to bed walls, thereby obstructing effective fluidization because of limited interaction between

the gas and solid phases. Research by Terblanche (2011) showed that coal particles with surface moisture levels higher than 7%_{wt} struggles to fluidize. He *et al.* (2015) recommended that the surface moisture of -6+1mm coal should be <2.5%_{wt}, for optimum results. Dewatering the coal that is fed to the ADMFB is required, as the majority of South African coals contains 5 and 6%_{wt} of surface moisture (De Korte, 2016). The moisture content of the fluidization air is also essential to the moisture levels of the product coal and should therefore be held to a minimum. The moisture present in the air stream could easily condense inside the bed and create improper fluidization as well as trouble with medium recovery (De Korte, 2016).

2.6.5 Particle size, shape and density

Particle size, density and shape are directly related to the location/position of the coal particles in the fluidized bed. Particles in the fluidized bed will float at different heights depending on the relative density difference between the particle itself and the bed (Prusti *et al.*, 2015). It was also found that for a particular particle shape, high dense coal with a small surface area would gradually move down along the height of the bed whereas coal particles that has a larger surface area would rapidly descend when density is increased. This response can be ascribed to different dead zones forming because of larger size and shape that may lead to an alteration in the density of the particle and ultimately the particle positioning along the bed height (Prusti *et al.*, 2015). Yang *et al.* (2013) remarked that the fluidized bed performance decreases with a reduction in particle diameter, which is a result of the reduction in settling velocity and increase in bubbling and remixing.

The fluidization performance is also affected by increasing particle size in relation to the bed diameter. When increasing the ratio of particle to bed diameter, the point of minimum fluidization is increased and a more prominent wall effect is observed (Rao *et al.*, 2010). Figure 2.6.5 provides curves that illustrates the change in particle position along the bed height according to coal density, for different particle shapes and sizes as established by Prusti *et al.* (2015).

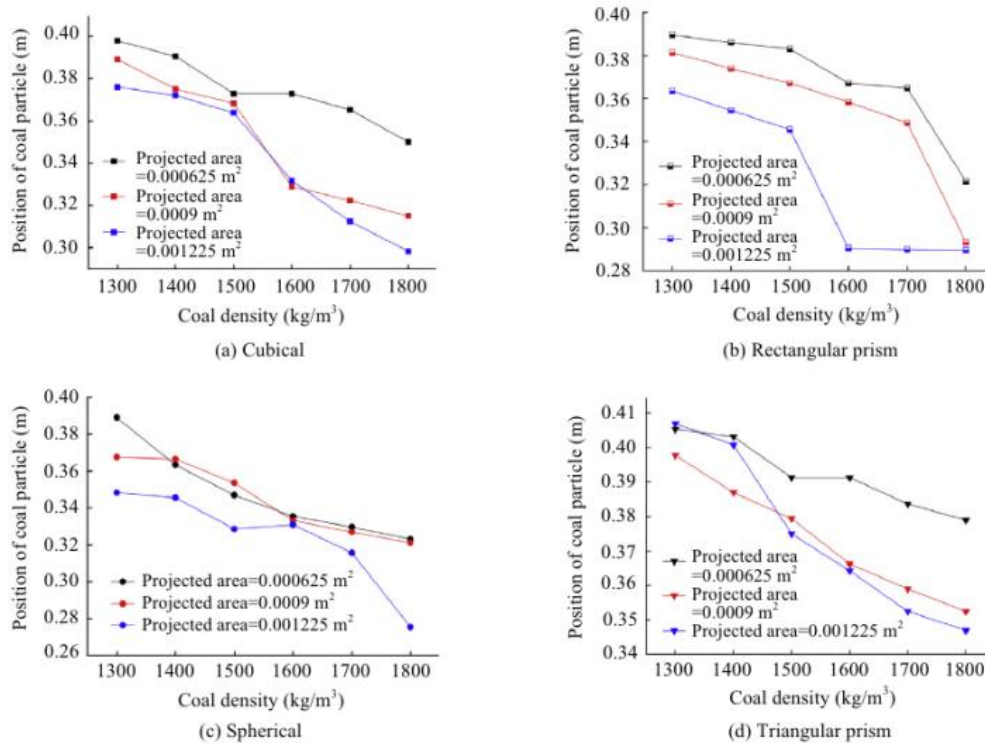


Figure 2.6-5: Effect of particle shape on the ADMFB separation (taken from Prusti *et al.* (2015))

From Figure 2.6.5, it is evident that particle shape effects the movement and behaviour of a single coal particle. Triangular shaped particles usually reach positions that are higher up along the bed height, regardless of density or size, followed by cubic, rectangular and spherical shapes. The position of the various shaped particles is due to the size of the dead zone areas. The largest dead zone area would generally form at the spherical shaped particles and the smaller dead zone area is formed at the triangular. Dead zones increase the density of a particle as medium and accumulated fines are deposited in the dead zone. This causes the particle to have an increased apparent weight, leading to a lower position along the bed height than anticipated (Prusti *et al.*, 2015).

2.6.6 Static bed height and column shape

The geometry and height of material in the fluidized bed is known to influence the point of minimum fluidization point of a particular bed. Rao *et al.* (2010) determined that an increase in minimum fluidization velocity is experienced when the column height to diameter ratio is increased for a column with fixed diameter. By increasing the column height to diameter ratio, the height of the column is increased, resulting in more prominent wall effects, which eventually leads to increased difficulty in particle fluidization. Fluidization of particles in a bed with a large static height to diameter ratio is therefore almost impossible. Contrarily, by

reducing the column height too low, most fine particles are removed from in between the coarse particles. This also results in hindering the fluidization process (Yang *et al.*, 2013b).

Sahu *et al.* (2013) found that the shape of the bed (cross-sectional) also has an influence on the minimum fluidization velocity. During this investigation, square, rectangular and circular fluidized beds were investigated to identify the bed geometry that would produce a more stable fluidization. The range between the point of minimum fluidization and point of minimum bubbling provides an indication of the stability of fluidization. These ranges were evaluated for circular, square and rectangular beds and an increase in fluidization stability was seen for circular to square and is the highest for rectangular beds. This is due to the high bed porosity as well as dead zones formed in the bed corners leading to an increased static bed height (Sahu *et al.*, 2013).

2.7 Development of a numerical model

In chemical industries, gas–solid fluidized bed reactors are well known and are employed in a wide range of applications. Various physical processes also employ fluidization as a technique for drying, separation and coating. The solid phase of fluidization, in practice, is usually composed up of two or more materials that differ in physical properties such as density, size and shape. Examples of fluidized bed systems include gasifiers and biomass combustors, in which the properties of the fluidization materials are remarkably different. Biomass particles are usually so large and permeable (light) that fluidization thereof is challenging and improbable. Therefore, an inert fluidization material or medium, such as alumina, magnetite or sand, is added to the system in order to assist with fluidization. However, due to the substantial size and density difference between the biomass material and bed medium particles, additional complexities and challenges arise. Proper segregation of the biomass particles and fluidized bed medium is one of these challenges. Improper segregation negatively impacts the performance of a fluidized bed performance in which chemical reactions occur. During processing of biomass, the large difference in relative density between the biomass and fluidized bed medium causes the biomass to travel to the top of the fluidized bed while almost none moves to the lower portion of the bed. This occurrence is referred to as “flotsam behaviour” and leads to reaction products being heterogeneously distributed in the bed, an uneven temperature profile across the fluidized bed, which results in a decline in product yield, tar formation as well as hot and cold spots throughout the bed (Fotovat *et al.*, 2015). It is important to note that segregation could be used favourably in physical treatments systems such as classifiers and fluidized bed separators (Sekito *et al.*, 2006; Yoshida *et al.*, 2010). Therefore, it is imperative to obtain comprehensive insight into particle segregation in fluidized beds.

A familiar experimental technique used to evaluate the degree of mixing and segregation in a fluidized bed that involves two diverse solid components is known as the “frozen bed” technique (Fotovat *et al.*, 2015, Hemati *et al.*, 1990). However, according to Fotovat *et al.* (2015), the frozen bed technique is incapable of providing the in-situ particle distribution under fluidization conditions. It is also known to be a demanding method in terms of the unpredictability of the solids in fluidized beds, making repetition of experimental runs essential in order to get consistent averages. Even though repetitions may provide reliable results, the results may still be distorted due to slumping transients in the bed for rapid mixing rates of the solids (Shen *et al.*, 2007). By utilizing advanced techniques for examining the hydrodynamic flow aspects of heterogeneous material in a fluidized bed, an accurate and representative set of results of the bed contents during the mixing state can be captured as well as insight into the underlying fluidization mechanisms can be availed (Fotovat *et al.*, 2015).

ADMFB has been well-known and widely used in different industries in various applications for numerous decades, however modelling of fluidized bed systems remains challenging due to the intricate underlying physics. Some of these challenges include the complexity of the hydrodynamic behaviour of the gas-solid interactions, transient behaviour and phase interactions of the systems. There is no known guideline compiled or defined for selection of appropriate model parameters (Gilbertson *et al.*, 1996). Gibilaro and Rowe (1974) and Naimer *et al.* (1982) are pioneers in modelling axial segregation of binate mixtures in fluidized beds. Various mechanisms driving segregation of diverse components were considered during the formation of the model. According to Fotovat *et al.* (2015), a study conducted by Zarza Baleato (1986) for biomass fluidization, reported that the Gibilaro and Rowe (GR) model can effectively predict the overall particle distribution in the bed at low concentrations of large biomass particles (Fotovat *et al.*, 2015). But Olivieri *et al.* (2004) noted that the experimental results may not always be reproduced properly by the model. It is difficult to correlate the parameters of the GR model with real values such as particle size and fluidization velocity (Olivieri *et al.*, 2004). Consequently, the use of the GR model is limited to indicating trends and comparing the effect of various associated mechanisms (Leaper *et al.*, 2007).

Simulations using computational fluid dynamics (CFD) may provide insight into the particle-scale phenomena that needs to be very well understood in order to effectively and successfully upscale the fluidized bed unit. Two CFD models namely, Eulerian–Lagrangian and Eulerian–Eulerian, are generally used for investigating fluidized bed systems. In the Eulerian–Lagrangian model, particles are modelled as discrete elements. Each individual particle’s Newtonian equations for motion are solved by inclusion of the influence of particle forces and collisions acting on the particles by the gas. By taking this approach, the fluid flow (continuum phase), can

be described by the local averaged Navier–Stokes equation. This approach is a combination of computational fluid dynamics and discrete particle method (CFD–DPM).

A characteristic of the Eulerian–Eulerian model is that two phases are treated as merged continua. Phase volume fractions are considered if the volume of a phase cannot be occupied by the other phases (Abbasi *et al.*, 2011). These volume fractions are continuous functions of time and space and the net of the two equals one. Conservation equations are derived for every phase by obtaining a collection of equations that have the same structure for all of the phases. The multi-fluid model known as “continuum–continuum”, is the most frequently used approach for projecting hydrodynamic behaviour of fluid–particle systems (Pain *et al.*, 2001).

The experimental and Euler–Euler 3-D computational studies on the dynamic flow behaviour of a gas–solid fluidized bed with varying particle size binary mixtures (Gao *et al.*, 2008). The results of the simulation were found to reasonably correspond to the experimental results with regards to bed density, bed expansion and local composition of the top section of the bed. From the experimental results it was found that smaller particles were displaced to the surface of the bed while larger particles of similar density tended to move to the bottom section of the bed. From the simulation portion of the investigation, it was seen that the small particles moved in a downward direction at the wall region and in an upward direction at the core.

Bai *et al.* (2013) investigated the segregation of particles in a fluidized bed with a cylindrical geometry. Different volume fractions of glass beads and ground walnut shell particles, both Geldart B type particles, were used. A 2-D multi-fluid model established on granular flow kinetic theory was selected to run the simulations. It was found that by increasing the biomass ratio at fluidization velocity ($U = U_{mf}$), an increase in the extent of segregation was experienced. However, the mixing of glass beads and biomass increased with an increase in gas velocity, irrespective of the component ratio in the mixture. From this it was concluded that the degree of segregation in different mixtures become similar at raised superficial gas velocities. Experimental as well as computational results showed small and light particles mixed poorly with large and heavy particles, although small and heavy particles proved to mix well with light and large particles.

A computational particle fluid dynamics (CPFD) model was used by Fotovat *et al.* (2014a) to simulate the bubbling characteristics of a bed involving mixtures of biomass and sand at various ratios. The results of the simulations, verified by the experimental test runs, were able to capture and illustrate the characteristic irregular bubble size as well as bubble velocity distributions. However, based on the biomass portion of the mixture, it is important to select a proper computational mesh size in order to achieve more precise results.

Fotovat *et al.* (2014a) conducted an experimental and numerical investigation of fluidized beds which involved spherical and cylindrical shaped biomass particles with sand as bed medium.

The investigation was focussed on the influence of the shape of the biomass particles on both its fluidization behaviour and segregation performance under predetermined bubbling conditions (Fotovát *et al.*, 2014a). Additionally, the ability of 3-D Eulerian–Eulerian CFD modelling to predict hydrodynamic features and characteristics of various previously studied systems is explored by comparing the experimental and simulation results. The simulation was able to replicate the experimental bubble size and velocity distribution profile at height equal to 175 mm; however, the chances of bubbles forming that were higher in velocity or larger in size (than the average values) was overestimated compared to the actual number of bubbles detected during the experimental runs. Like that observed for experiments, the simulation results predicted considerably less intense biomass particle velocity fluctuations in the radial compared to the axial direction. In comparison to the experimental results, the simulation results showed a somewhat more even biomass distribution in the axial direction. The discrepancies observed between the experimental results and that produced by simulation were mostly attributed to the inability of the simulation to allow for the impact of biomass particles on bubbling properties, especially where biomass accumulation is substantial.

2.7.1 Simulation model

NEPTUNE_CFD is the CFD software employed for the numerical simulations in this research study. The code is developed in the framework of the NEPTUNE project financially supported by the Commissariat for Atomic Energy and Alternative Energies (CEA), Electricity of France (EDF), The Institute for Radiation Protection and Nuclear Safety (IRSN) and AREVA-NP. A multifluid Eulerian method is followed by the code with a polydispersed fluid-particle capability applied by The Institute of Fluid Mechanics of Toulouse (IMFT) (Neau *et al.*, 2013).

The Eulerian n-fluid method employed is a hybrid approach (Morioka and Nakajima, 1987) where the transport equations are derivatives of phase ensemble averaging for the continuous phase and by implementation of kinetic theory of granular flows augmented by fluid and turbulent effects for the dispersed phase via a combined fluid–particle probability density function (PDF) approach. In the selected modeling approach, equations for mass, momentum and fluctuating kinetic energy are solved for every phase and joined by inter-phase transfer terms. The momentum transfer between particle and gas phases are modeled and represented by means of the drag law of Wen and Yu restricted by Ergun equation for the dense flows (Gobin *et al.*, 2003; Peirano *et al.*, 2002). The collisional particle stress tensor is a derivative of the frame of the kinetic theory of granular media (Boelle *et al.*,

1995). The fluid turbulence modeling is attained by the equations of $k - \varepsilon$ model extended to heavily loaded particle flows (particle-laden flows) which account for supplementary source terms as a result of the inter-phase interactions (Vermorel *et al.*, 2003). A combined transport equation system is solved on particle fluctuating kinetic energy and fluid–particle fluctuating covariance for the dispersed phase.

The impact of the particle-particle contact force in the densest zone of the flow is considered in the particle stress tensor by the added frictional stress tensor (Srivastava and Sundaresan, 2003). The interaction modeling of particle species method is established in the frame of simulation of reacting solid gas polydispersed reactive fluidized beds (Batrak *et al.*, 2005; Gourdel *et al.*, 1999).

2.7.2 Governing Equations

The Eulerian n-fluid method employed in this study is a hybrid method whereby phase ensemble averaging for the continuous phase is used to derive the transport equations and the kinetic theory of granular flows with fluid and turbulent effects for the dispersed phase was derived from the joint fluid–particle probability density function approach. In the next set of equations, subscript k or m_g denotes to the gas phase and k , m_p or m_g to the particle phase. $\alpha_m \rho_m$ ($m = p, q$) in the particle transport equation denote $n_m m_m$ where n_m is the number density of m particle centre and m_m is the mass of a single m particle: $\alpha_m = n_m m_m / \rho_m$ is an estimate of the local volume fraction of particle m . Therefore, the gas and particle volume fractions represented by α_g , α_p and α_q should equal 1 as shown by Equation 2.10.

$$\alpha_g + \alpha_p + \alpha_q = 1 \quad \text{Equation 2.10}$$

2.7.2.1 Mass transport equation

$$\frac{\partial}{\partial t} \alpha_k \rho_k + \frac{\partial}{\partial x_j} \alpha_k \rho_k U_{k,j} = 0 \quad \text{Equation 2.11}$$

where ρ_k is density of the k phase and $U_{k,j}$ is the i -component of its velocity.

2.7.2.2 Momentum transport equation

$$\alpha_k \rho_k \left[\frac{\partial U_{k,i}}{\partial t} + U_{k,j} \frac{\partial U_{k,i}}{\partial x_j} \right] = \alpha_k \frac{\partial P_g}{\partial x_i} + \alpha_k \rho_k \rho g_i + \sum_{m \neq k} I_{m \rightarrow k,i} - \frac{\partial \Sigma_{k,ij}}{\partial x_j} \quad \text{Equation 2.12}$$

Where P_g is the average gas pressure and g_i is the gravity component, i . $I_{g \rightarrow p}$ and $I_{g \rightarrow q}$ are the inter-phase momentum transfer in between gas and particle without gas pressure input and $I_{g \rightarrow q} = -I_{g \rightarrow p}$ refers to the inter-phase momentum transfer between particle species. $\sum I_{m \rightarrow k,i}$ refers to the effective stress tensor of phase k .

2.7.2.3 Interphase transfer modelling

Gas–particle average momentum transfer

Based on the particle to gas density ratio, Archimede's forces are the dominant forces between the gas phase and p -particles, therefore average momentum gas–particle transfer term is depicted by Equations 2.13 to 2.19.

$$I_{g \rightarrow p,i} = -\frac{\alpha_p \rho_p}{\tau_{gp}} V_{rp,i} \quad \text{Equation 2.13}$$

$$\frac{1}{\tau_{gp}} = \frac{3\rho_g \langle |\vec{v}_r| \rangle}{4\rho_p} C_d(Re_p) \quad \text{Equation 2.14}$$

$$C_d(Re_p) = \begin{cases} C_{d,WY} & \text{if } \alpha_g \geq 0.7 \\ \min[C_{d,WY}, C_{d,Er g}] & \text{if } \alpha_g < 0.7 \end{cases} \quad \text{Equation 2.15}$$

$$C_d(Re_p) = \begin{cases} \frac{24}{Re_p} (1 + 0.15 Re_p^{0.687}) \alpha_g^{-1.7} & Re_p < 1000 \\ 0.44 \alpha_g^{-1.7} & Re_p \geq 1000 \end{cases} \quad \text{Equation 2.16}$$

$$C_{d,Er g} = 200 \frac{(1-\alpha_g)}{Re_p} + \frac{7}{3}, \quad Re_p = \alpha_g \frac{\rho_g \langle |\vec{v}_r| \rangle_p d_p}{\mu_g} \quad \text{Equation 2.17}$$

$$V_{rp,i} = \langle v_{r,i} \rangle_p = U_{p,i} - U_{g,i} - V_{dp,i} \quad \text{Equation 2.18}$$

$$V_{dp,i} = -D_{gp,ij}^t \left[\frac{1}{\alpha_p} \frac{\partial \alpha_p}{\partial x_j} - \frac{1}{\alpha_g} \frac{\partial \alpha_g}{\partial x_j} \right] \text{ with } D_{gp,ij}^t = \frac{1}{3} \tau_{gp}^t q_{gp} \delta_{ij} \quad \text{Equation 2.19}$$

$V_{rp,i}$ is the average relative velocity and can be presented in terms of the average gas and p -particle velocities and a turbulent gas–particle drift velocity. $V_{dp,i}$ depicts the transference of the p -particles via turbulent gas eddies (Gobin *et al.*, 2003; Simonin *et al.*,

1993). $C_{d,WY}$ is Wen & Yu drag coefficient and $C_{d,Erg}$ is the Ergun drag coefficient, respectively.

q to p-particles average momentum transfer

From Gourdel *et al.* (1999), particle collisions lead to an average momentum transfer between classes and is shown by Equations 2.20 to 2.38.

$$I_{g \rightarrow p,i} = -\frac{m_p m_q}{m_p + m_q} \frac{1+e_c}{2} f_{pq}^c H_1(z) (U_{p,i} - U_{q,i}) \quad \text{Equation 2.20}$$

$$f_{pq}^c = g_0 4\pi n_p n_q d_{pq}^2 g_r, \quad d_{pq} = (d_p + d_q)/2 \quad \text{Equation 2.21}$$

$$n_p = \frac{\alpha_p}{\pi \frac{d_p^3}{6}}, \quad m_p = \rho_p \pi \frac{d_p^3}{6} \quad \text{Equation 2.22}$$

$$g_r = \sqrt{\frac{16}{\pi} \frac{2}{3} q_{pq}^2 + V_{pq,i} V_{pq,i}}, \quad V_{pq,i} = U_{p,i} - U_{q,i}, \quad q_{pq}^2 = \frac{1}{2} (q_p^2 + q_q^2) \quad \text{Equation 2.23}$$

$$g_0 = \left[1 - \frac{1-\alpha_g}{\alpha_{max}} \right]^{-2.5\alpha_{max}}, \text{ if } 1 - \alpha_g < \alpha_{max} \quad \text{Equation 2.24}$$

$$H_1(z) = \frac{8+3z}{6+3z} \quad \text{Equation 2.25}$$

2.7.2.4 Particle stress modeling

$$\Sigma_{p,ij} = \left[P_p + \lambda_p \frac{\partial U_{p,n}}{\partial n} \right] \delta_{ij} - \mu_p \left[\frac{\partial U_{p,i}}{\partial x_j} + \frac{\partial U_{p,j}}{\partial x_i} - \frac{2}{3} \frac{\partial U_{p,n}}{\partial x_n} \delta_{ij} \right] \quad \text{Equation 2.26}$$

$$\mu_p = \alpha_p \rho_p (v_p^{kin} + v_p^{col} + v_p^{fric}) \quad \text{Equation 2.27}$$

$$v_p^{kin} = \left[\frac{1}{3} q_{gp} \tau_{gp}^t + \frac{1}{2} \tau_{gp}^F \frac{2}{3} q_p^2 (1 + \hat{\alpha}_p g_0 \phi_c) \right] \times \left[1 + \frac{\tau_{gp}^F \sigma_c}{2 \hat{\tau}_p^c} \right]^{-1} \quad \text{Equation 2.28}$$

$$v_p^{col} = \frac{4}{5} \hat{\alpha}_p g_0 (1 + e_c) \left[v_p^{kin} + \hat{d}_p \sqrt{\frac{2}{3} \frac{q_p^2}{\pi}} \right] \quad \text{Equation 2.29}$$

$$v_p^{frict} = \frac{P_f \sin \phi}{2(1-\alpha_g)\rho_p \sqrt{I_{2D} + \left(\frac{2}{3}\right)\left(\frac{q_p^2}{d_p^3}\right)}}, P_f = Fr \frac{[(1-\alpha_g)-\alpha_{s,min}]^n}{[\alpha_{s,max}-(1-\alpha_g)]^q}, \quad Fr = 0.05, \quad \phi = \frac{\pi}{4},$$

$$I_{2D} = \left[\frac{\partial U_{p,i}}{\partial x_j} + \frac{\partial U_{p,j}}{\partial x_i} \right] \frac{\partial U_{p,i}}{\partial x_j} - \frac{2}{3} \frac{\partial U_{p,i}}{\partial x_j} \frac{\partial U_{p,j}}{\partial x_i}, \quad n = 2, q = 5, \quad \alpha_{s,min} = 0.6$$

Equation 2.30

$$\phi_c = \frac{2}{5}(1+e_c)(3e_c-1), \quad \sigma_c = \frac{1}{5}(1+e_c)(3-e_c) \quad \text{Equation 2.31}$$

$$P_p = \alpha_p \rho_p \left[1 + 2\hat{\alpha}_p g_0 (1+e_c) \right] \frac{2}{3} q_p^2 \quad \text{Equation 2.32}$$

$$\lambda_p = \alpha_p \rho_p \frac{4}{3} \hat{\alpha}_p g_0 (1+e_c) \hat{d}_p \sqrt{\frac{2}{3} \frac{q_p^2}{\pi}} \quad \text{Equation 2.33}$$

$$\hat{\alpha}_p = \sum_{m=p,q} \alpha_m \frac{2m_q}{m_p + m_m} \left[\frac{d_{pm}}{d_m} \right]^3 \quad \text{Equation 2.34}$$

$$\frac{1}{\hat{\tau}_p^c} = \sum_{m=p,q} \alpha_m \frac{2m_m}{m_p + m_m} \frac{1}{\tau_{pq}^c} \quad \text{Equation 2.35}$$

$$\hat{d}_p = \frac{1}{\hat{\alpha}_p} \sum_{m=p,q} \alpha_m \frac{d_{pm}^4}{d_m^3 m_p + m_m} \quad \text{Equation 2.36}$$

$$\tau_{gp}^t = \frac{\tau_t}{\sigma_t} [1 + C_\beta \xi_r^2]^{-1/2}, \quad \xi_r = \frac{|v_{rp}|}{\sqrt{\frac{2}{3}k}} \quad \text{Equation 2.37}$$

$$\tau_t = C_\mu \frac{3}{4} \frac{k}{\varepsilon}, \quad k = \langle u'_{g,i} u'_{g,i} \rangle / 2 \quad \text{Equation 2.38}$$

Where the turbulent Schmidt of gas phase is denoted by σ_t . C_β and C_μ are constants that are equivalent to 1.8 and 0.09, respectively. k and ε are the gas phase turbulent kinetic energy and the gas turbulence dissipation rate, respectively, represented with two transport equations considering the supplementary source terms because of the interphase interactions (Vermorel *et al.*, 2003).

2.7.2.5 Particle fluctuant kinetic energy transport equation

$$\alpha_p \rho_p \left[\frac{\partial q_p^2}{\partial \tau} + U_{p,j} \frac{\partial q_p^2}{\partial x_j} \right] = \frac{\partial}{\partial x_j} \left[\alpha_p \rho_p (K_p^{kin} + K_p^{col}) \frac{\partial q_p^2}{\partial x_j} \right] - \sum_{p,i,j} \frac{\partial U_{p,j}}{\partial x_j} - \frac{\alpha_p \rho_p}{\tau_{gp}^t} [2q_p^2 - q_{gp}] - \sum_{m=p,q} \varepsilon_{pm} + \sum_{m=p,q} X_{pm}$$

Equation 2.39

$$K_p^{kin} = \left[\frac{1}{3} q_{gp} \tau_{gp}^t + \frac{2}{3} q_p^2 \frac{5}{9} \tau_{gp}^F (1 + \hat{\alpha}_p g_0 \phi_c) \right] x \left[1 + \frac{5}{9} \tau_{gp}^F \frac{\xi_c}{\tau_{gp}^t} \right]^{-1} \quad \text{Equation 2.40}$$

$$\phi_c = \frac{3}{5} (1 + e_c)^2 (2e_c - 1), \quad \xi_c = \frac{(1+e_c)(49+33e_c)}{100} \quad \text{Equation 2.41}$$

$$K_p^{col} = \hat{\alpha}_p g_0 (1 + e_c) \left[\frac{6}{5} K_p^{kin} + \frac{4}{3} \hat{d}_p \sqrt{\frac{2}{3} \frac{q_p^2}{\pi}} \right] \quad \text{Equation 2.42}$$

$$\varepsilon_{pq} = m_p \left(\frac{2m_q}{m_p + m_q} \right)^2 \frac{1-e_c^2}{4} f_{pq}^c \frac{2}{3} (q_p^2 + q_q^2) \quad \text{Equation 2.43}$$

$$x_{pq} = \frac{m_q m_p}{m_p + m_q} \frac{1-e_c}{2} f_{pq}^c \left[\frac{m_q}{m_p + m_q} \frac{1-e_c}{2} (U_{p,i} - U_{q,i})^2 H_1(z) - \frac{8}{3} \frac{m_p q_p^2 - m_q q_q^2}{m_p - m_q} \right] \quad \text{Equation 2.44}$$

2.7.2.6 Fluid–particle velocity covariance transport equation

$$\alpha_p \rho_p \left[\frac{\partial q_{gp}}{\partial t} + U_{p,j} \frac{\partial q_{gp}}{\partial x_j} \right] = \frac{\partial}{\partial x_j} \left[\alpha_p \rho_p \frac{v_{gp}^t}{\sigma_k} \frac{\partial q_{gp}}{\partial x_j} \right] - \alpha_p \rho_p \langle u'_{g,j} u'_{p,jp} \rangle \frac{\partial U_{p,j}}{\partial x_j} - \alpha_p \rho_p \langle u'_{g,j} u'_{p,jp} \rangle \frac{\partial U_{p,i}}{\partial x_j} - \Pi_{q_{gp}} - \alpha_p \rho_p \varepsilon_{gp} \quad \text{Equation 2.45}$$

$$\langle u'_{g,i} u'_{p,jp} \rangle = v_{gp}^t \left[\frac{\partial U_{g,i}}{\partial x_j} + \frac{\partial U_{p,j}}{\partial x_i} \right] + \frac{1}{3} \left[q_{gp} + v_{gp}^t \frac{\partial U_{g,n}}{\partial x_n} + v_{gp}^t \frac{\partial U_{p,n}}{\partial x_n} \right] \delta_{ij} \quad \text{Equation 2.46}$$

$$v_{gp}^t = \frac{1}{3} q_{gp} \tau_{gp}^t \quad \text{Equation 2.47}$$

$$\varepsilon_{gp} = \frac{q_{gp}}{\tau_{gp}^t} \quad \text{Equation 2.48}$$

$$\Pi_{q_{gp}} = \frac{\alpha_p \rho_p}{\tau_{gp}^F} \left[\left(1 + \frac{\alpha_p \rho_p}{\alpha_g \rho_g} \right) q_{gp} - 2k - 2 \frac{\alpha_p \rho_p}{\alpha_g \rho_g} q_p^2 \right] \quad \text{Equation 2.49}$$

2.8 Previous Case Studies

For the past decade, the Institute of Minerals and Materials Technology of India have launched numerous investigations into dry coal beneficiation, including fluidized bed separation technology (Sahu *et al.*, 2009). A continuous processing pilot scale ADMFB with a throughput of 600 kg/h high-ash Indian coals was developed at this institute (Sahu *et al.*,

2011). Magnetite dense medium, of $7.26\mu\text{m}$ and $21.7\mu\text{m}$ sizes, were used to generate a non-bubbling, uniform gas-solid fluidization condition. The Geldart A group dense medium creates a larger bed expansion with a relatively lower bed density of 1.6g/cm^3 . High-ash, non-coking coal, with particle size range of $+6-25\text{mm}$, was successfully beneficiated with a cut point density of 1.68g/cm^3 and an E_p value of 0.12g/cm^3 , and an effective ash content reduction of between 8 - 4.5% and a product yield of up to 72%. The stability of different fluidized bed was investigated with reference to bed geometry (Sahu *et al.*, 2013). Several parameters including the particle expansion function, fluidization index, bed and distributor pressure drop as well as bubbling and minimum fluidization velocities were selected to be tested for characterization of the operational stability of the bed. A fluidized bed with a rectangular cross-sectional geometry provided a more stable bed than a bed of circular square geometry.

Mohanta *et al.* (2011) have studied dry beneficiation of coal using fluidized bed technology at the Indian Institute of Technology in India. Over the last couple of years, the effect of the feed coal particle size, on the bed stratification performance was evaluated with a lab scale ADMFB (Mohanta *et al.*, 2011). Mohanta *et al.* (2011) attempted to quantify and identify the optimum feed coal size range over which satisfactory separation can be obtained in the system. Coal samples from various local coal mines were tested to demonstrate the relationship between the feed coal size and the sharpness of the bed separation. The results obtained from the fluidized bed system showed that, regardless of the types of feed coals, all 15-50mm coal was satisfactorily beneficiated. Moreover, the separation performance of various feed parameters of cubic, flat, and sharp-edged prism shapes was evaluated in a batch wise ADMFB system. It was concluded that cubic shaped coal, with the smallest volume to surface area ratio, was separated better compared to sharp-pointed shapes, as the medium viscosity effects is reduced (Chikerna *et al.*, 2012). Lastly, the minimum fluidization velocities of various magnetite powders employed in the ADMFB system were established, and based on particle properties, a semi-empirical correlation was set up for determining the minimum fluidization velocities without information of the shape factors and bed voidage (Mohanta *et al.*, 2012).

Azimi *et al.* (2013) comprehensively studied the separation performance of low-ash coal using dry fluidized bed separation technology at the University of Alberta, Canada (Azimi *et al.*, 2013 (a); Azimi *et al.*, 2013 (b)). The influence of operating parameters such as residence time, bed height and superficial gas velocity on the fluidized bed performance was explored using response surface methodology. The effect of feed coal weight and size on gravity separation was also addressed individually. The results showed that the effectiveness of the operating parameters tested ranged from residence time > bed height >

superficial gas velocity. A beneficiated product with an ash value of about 11%_{wt} and a recovery rate of 95.6% was achieved at the optimum operating condition projected by the proposed mathematical method. A number of progressive techniques were employed to evaluate the beneficiation performance of the coal in an ADMFB system. One of these, Inductively Coupled Plasma-Mass Spectrometry and X-ray Fluorescence, was used to synthetically study the immigration behaviour of hazardous elemental components and main clay mineral components during the fluidization processing, respectively. Differential Thermogravimetry was used to analyse the variation in reactivity of the clean coal containing the peak temperature and maximum weight loss rate. In order to gain in depth understanding of the ADMFB separation technique, statistical analysis and CFD simulations were used to explore the fluidization hydrodynamics, flow patterns and beneficiation performance of the continuous ADMFB system (Azimi *et al.*, 2015; Azimi *et al.*, 2017).

Furthermore, studies at University of Alberta in Canada also studied the prospect of applying ADMFB technology on fine coal particles (Chong *et al.*, 2006). Different particle size ranges of magnetite, including 45 – 75 μm , 45 – 106 μm , and 150 – 300 μm , were selected as dense medium. Acceptable separation efficiencies can be achieved with a feed coal size of +1mm. More precisely, +3.35-5.66mm feed coal can be adequately beneficiated by employing +45-106 μm magnetite particles, with a corresponding error of $\pm 0.03\text{g/cm}^3$. It was also concluded that coal of +1-3.35mm could be successfully beneficiated in an ADMFB by employing magnetite particles of +45-75 μm , with a probable error of 0.10g/cm^3 .

The possibility of mercury rejection by using ADMFB as a dry beneficiation technique was also investigated (Mak *et al.*, 2008). It was discovered that the mercury can be rejected along with the mineral matter and report to the gangue fraction of the ADMFB product, and a definitive linear relationship between the mineral matter and mercury content in the gangue was identified. The mercury and mineral matter rejection of 58 and 60%, respectively, was achieved with an overall combustible loss of $\pm 13\%$.

In 2011, China University of Mining and Technology, Zhao *et al.* (2011) recently developed a modularized industrial ADMFB plant together with Tangshan Shenzhou Manufacturing Co. Ltd. for effective dry beneficiation of coal (Zhao *et al.*, 2011). Figure 2.8.1 shows the modularized industrial ADMFB plant.



Figure 2.8-1: World's first modularized dry coal beneficiation plant (Zhao *et al.*, 2016)

In this design, almost all the operational processes are integrated into one modularized ADMFB unit. Figure 2.8.2 shows an engineering flowsheet of the modularized ADMFB system.

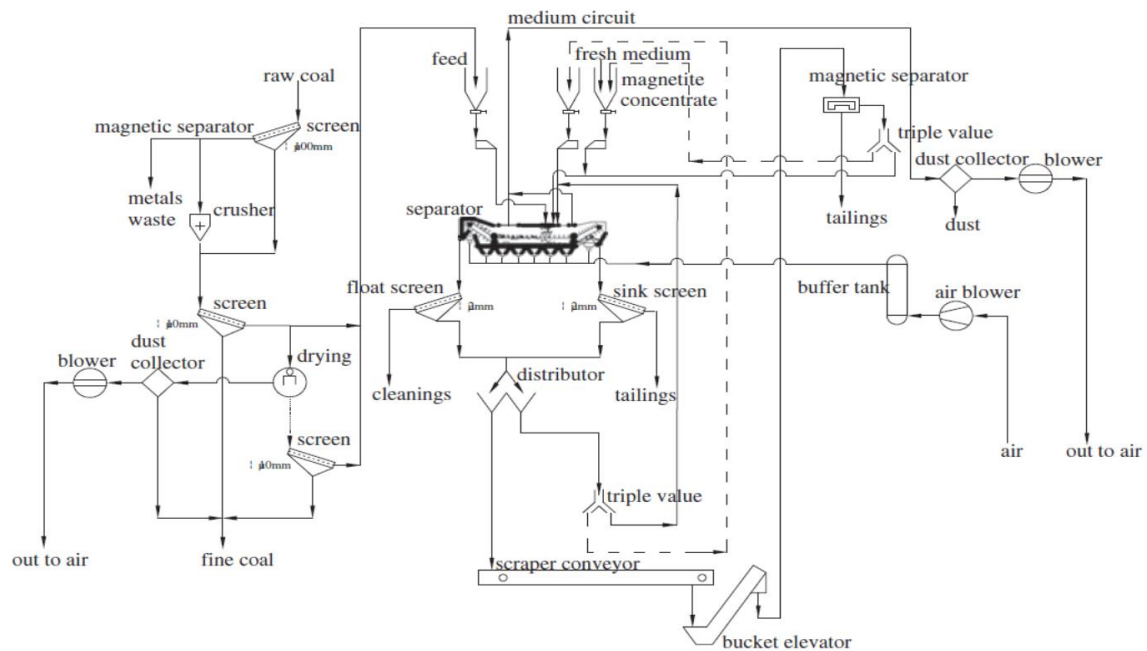


Figure 2.8-2: Engineering flowsheet of the modularized system (Zhao *et al.*, 2016)

The ADMFB unit includes pre-treatment of ROM coal, ADMFB separation, and product separation and cleaning, dense medium circulation, dust collection system, air supply system, etc. When comparing the modularized ADMFB system to a conventional ADMFB

plant, the CapEx and OpEx is reduced by 60% and 80%, respectively, which again is also far below the cost of a conventional wet beneficiation plant. A commercial, modularized ADMFB system was developed, implemented and placed into full operation at the Xinjiang Energy Co. Ltd. of China since 2013 (Zhang *et al.*, 2014; Zhao *et al.*, 2017). The dimensions of the new modularized ADMFB system were Length (L) x Width (W) x Height (H) = 25 m x 8 m x 11 m, and the coal beneficiation throughput is 40 to 60 t/h. ROM coal with a particle size range of +10-100 mm was effectively beneficiated with an error of $\pm 0.055 \text{ g/cm}^3$. The clean coal ash content was reduced from about 24%_{wt} to 3.5%_{wt} with an 86% ash rejection, which can be redirected to the activated carbon circuit. Moreover, the dense medium consumption and OpEx was found to be <5 kg and \$2/t of product coal, respectively. Hence, the modularized ADMFB system has the benefits of reduced CapEx and OpEx, good separation efficiency and being environmentally friendly, which offers a unique and desired solution for dry coal beneficiation.

2.9 Conclusion

Coal remains an important energy-resource base worldwide as it is considered abundant, affordable and reliable (WEC, 2022), meeting up to 83% of South Africa's maximum electricity generation in 2018 (DOE, 2019). The coal-based energy demand in South Africa's is unlikely to change in the next 2 decades because of the lack of appropriate alternative energy sources.

Wet beneficiation processes are known to yield good separation efficiencies however, it requires significant quantities of water. Moreover, difficulties in handling, transport and storage as well as increased capital, operating and transportation cost result from excess water retained in coal (Choung *et al.*, 2006). Furthermore, in water scarce coal producing and processing countries, including South Africa, a suitable water supply is restricted, and wet coal beneficiation techniques is therefore not ideal (Chen & Yang, 2003; Zhao *et al.*, 2015).

The industry is currently investigating dry methods of coal beneficiation in the hopes of finding an alternative that can produce acceptable separation efficiencies. Dry beneficiation techniques, such as ADMFB system, can prove to be an effective solution to eradicate ash-forming mineral fractions from ROM coal to improve coal quality and alleviate application issues as well as reduce carbon footprint.

ADMFB has been well-known and widely used in different industries in various applications for numerous decades, however modelling of fluidized bed systems remains challenging due

to the intricate underlying physics. Simulations using computational fluid dynamics (CFD) may provide insight into the particle-scale phenomena that needs to be very well understood in order to effectively and successfully upscale the fluidized bed unit. NEPTUNE_CFD is the CFD software employed for the numerical simulations in this research study. A multifluid Eulerian method is followed by the code with a polydispersed fluid-particle capability applied by The Institute of Fluid Mechanics of Toulouse (IMFT) (Neau *et al.*, 2013). The Eulerian n-fluid method employed is a hybrid approach (Morioka and Nakajima, 1987) where the transport equations are derivatives of phase ensemble averaging for the continuous phase and by implementation of kinetic theory of granular flows augmented by fluid and turbulent effects for the dispersed phase via a combined fluid–particle probability density function (PDF) approach.

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CHAPTER 3

Chapter 3 contains the paper titled “Dry beneficiation of small South African coal using an air dense medium fluidization bed”. In this paper, air dense medium fluidized bed (ADMFB) is evaluated as a suitable dry beneficiation technique to upgrade **+5.6mm-13.2mm** South African coal. The separation efficiency of the ADMFB was assessed by determining the effect, if any, of particle size range, medium to coal feed ratios and the addition of vibration on the separation ability of the bed. The coal was upgraded to varying degrees, depending on the parameters of the applied methods.

Dry beneficiation of small South African coal using an air dense medium fluidization bed (Paper 1)

This section provides information regarding contributions made by the various authors towards completion of the paper. General information regarding the published paper along with additional information of the journal by which the paper was published.

Contribution by authors

Misses E.S. Diedericks conducted a coal beneficiation research study with the guidance supervision of Prof. M. le Roux and Prof. Q.P. Campbell. The concept and scope of work for this research were largely formulated by Prof. M. le Roux and Prof. Q.P. Campbell. The air dense medium fluidized bed experimental setup was designed by Miss N Hughes and constructed by the North-West University technical support staff, with guidance from Miss N. Hughes. Misses E.S. Diedericks was responsible for conducting the experimental work, collecting data and processing of the collected data. The results from experimental runs were discussed and assessed during multiple consultations between Misses E.S. Diedericks, Prof. M. le Roux and Prof. Q.P. Campbell. After compilation review cycles, the paper was published in a peer-reviewed journal, International Journal of Coal Preparation and Utilization. The paper was written by Misses E.S. Diedericks and reviewed and edited by Prof. M. le Roux as well as Prof. Q.P. Campbell.

General information

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3.1 Dry beneficiation of small South African coal using an air dense medium fluidization bed

ES Peters-Diedericks, M le Roux, QP Campbell & N Hughes*

School of Chemical and Minerals Engineering, North-West University, Potchefstroom, South Africa.

**marco.leroux@nwu.ac.za*

Abstract

This paper investigates the air dense medium fluidized bed (ADMFB) as a suitable dry beneficiation technique to upgrade +5.6mm-13.2mm South African coal. The separation efficiency of the ADMFB was assessed by determining the effect, if any, of particle size range, medium to coal feed ratios and the addition of vibration on the separation ability of the bed. The coal was upgraded to varying degrees, depending on the parameters of the applied methods. For example, when considering the bottom bed layer (one quarter of the total bed height) as discard, the cumulative ash yield of the clean coal fraction ranged between 13.0% to 17.6% with a corresponding yield of 82.21% to 83.93%, respectively.

A variation in feed coal PSD did not significantly influence the response of the bed pertaining to its ability to beneficiate the coal. This was quantified by the similarity in percentage ash yield for the feed, top and bottom layers. In addition, it was established that by feeding a coal and magnetite (+300 μ m) mixture, in various mass ratios, did not significantly influence the separation efficiency of the bed, nor did the application of external vibration to the bed aid in the beneficiation of the coal. Although these factors did not influence the performance of the ADMFB, it was found that the minimum fluidization velocity and bed stability was largely improved by the introduction of these variables.

Keywords: Small South African coal, dry coal beneficiation, dense medium, magnetite, air dense medium fluidization, minimum fluidization velocity.

Introduction

Coal remains an important energy resource worldwide because of its abundance and affordability. Consequently, coal will be applied as fuel source to meet about 29% of the world's energy demands for the next 30 years (World Energy Council, 2017). The energy resource base of South Africa is currently dominated by coal, with coal fired power stations

providing more than 75% of the country's power supply (DOE, 2017). The Department of Mineral Resources (2014) noted that coal contributed a significant portion to South Africa's export market as about 28% of the ROM coal was shipped to China, India and Europe.

The qualities of economically mineable coal deposits are decreasing (Jeffrey *et al.*, 2014). Added to this, the general scarcity of water in coal producing regions are also becoming an important production and environmental consideration for coal producing countries. In fact, over the last decade, South Africa has been labelled a water-stressed country (Rand Water, 2017). Rand Water (2017) recorded an average annual rainfall of 492mm for South Africa during 2016, which is roughly half of the global average. The standardized precipitation index (SPI) of South Africa is provided in *Figure 1* for the period April 2015 to January 2017.

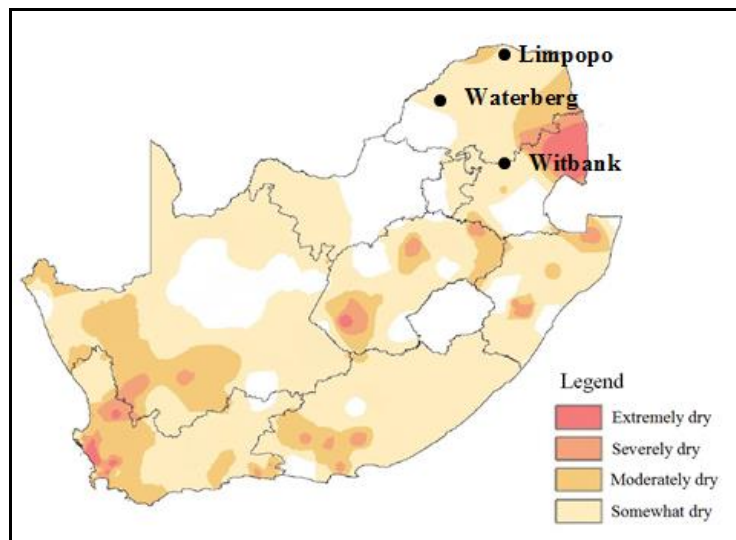


Figure 3.1-1: Standard precipitation index for South Africa adapted from SAWS (2017)

The areas that are especially concerning to the coal industry of South Africa, shown in *Figure 3.1-1*, are the Limpopo, Waterberg and Witbank coalfields, which lie within the more arid parts of the country. As such, the focus of coal processing industries in this country, as well as worldwide, is shifting away from wet processing to efficient dry processing, in an attempt to secure an energy sustainable future.

Background

Coal is a fossil fuel that is primarily made up of carbonized plant remains and numerous inorganic deposits. The heterogeneous nature of coal often warrants removal of assorted impurities before consumption. By removing the impurities, a higher quality coal is obtained that yields better energy returns as well as emissions that adhere to environmental legislations (Chen & Yang, 2003). Conventional processes used to remove impurities from

coal employ a wide range of mechanization and water treatment applications. Separation of the desired and undesired materials in the coal takes place during beneficiation (World Coal Institute, 2009). Currently, wet beneficiation processes produce high quality separation and acceptable results, however, considerable quantities of process water are required. This intensive water usage leads to problems associated with handling, storage and transportation as well as increased capital overheads and operational expenditures (Choung *et al.*, 2006). In addition to this, water-scarce coal-producing countries, such as Mongolia and South Africa, have insufficient or unreliable process water that makes wet beneficiation practices not ideal (Chen & Yang, 2003; Zhao *et al.*, 2015).

Current dry beneficiation techniques such as FGX separators, X-Ray sorters, air jigging, pneumatic oscillating tables and air dense medium fluidized beds (ADMFB), have been investigated and the reported performance were close to those obtained for wet beneficiation techniques (Chen & Wei, 2003; Zhao *et al.*, 2015). Air dense medium fluidization has received a considerable amount of attention as an alternative dry beneficiation technique and has been implemented in China (Mohanta *et al.*, 2011). The principle of the technique is that coal particles are stratified according to density, with the high-quality coal (lower density particles) reporting to the overflow, and the higher dense waste particles to the underflow (Jingfeng *et al.*, 2016). By the early 2000's, numerous successes have been recorded in various countries for beneficiating coal using ADMFB's technology. It is evident from *Table 3.1-1* that the separation efficiency (E_p) is highly dependable on the minimum particle size and formation origin of the feed coals.

Table 3.1-1: Separation results of various coal size fractions

Country	PSD	E_p (g/cm³)	Medium	Reference
China	>50mm	0.02	Magnetite	Chen & Yang (2003)
China	-50+6mm	0.05-0.07	Magnetite	Zhao <i>et al.</i> (2011)
India	-25+13mm	0.06	Magnetite	Mohanta <i>et al.</i> (2011)
India	-13+4.75mm	0.19	Magnetite	Mohanta <i>et al.</i> (2011)
Canada	-1+0.42mm	0.10	Magnetite	Choung <i>et al.</i> (2006)

Experimental

MATERIALS USED

Coal

Coal from the number 4 seam in the Witbank area in South Africa was used for this study. This is one of the largest South African coalfields, but it is expected to be exhausted by 2040 (Jeffrey, 2005). Coal from seam 4 is bituminous thermal grade coal, and is mined for the export market, with a small portion sold into the local market (Jeffrey, 2005).

Sample preparation consisted of drying, crushing and screening to provide the feed to the ADMFB. From research, it was established that coal fluidization is best achieved for particle surface moisture levels below 5%_{wt} (Sahu *et al.*, 2009; Terblanche, 2013). The coal (as received) was air-dried by laying it out on a flat surface, at ambient air conditions, for 24 hours. Hereafter it was crushed and screened into the particle size ranges shown in *Table 3.1-2*. A representative sample of the material in each particle size range was taken by the cone-and-quarter method. The proximate analysis and calorific values for each PSD fraction were determined and it was found that the results were comparable across the different size ranges (*Table 3.1-2*).

Table 3.1-2: Average initial proximate and calorific values for each coal particle size range

Description	PSD-01 (+5.6mm-6.7mm)	PSD-02 (+6.7mm-8.0mm)	PSD-03 (+8.0mm-9.5mm)	PSD-04 (+9.5mm-11.2mm)	PSD-05 (+11.2mm-13.2mm)	Standards
Moisture content (% _{wt})	1.07	0.61	1.07	1.04	0.95	ACT-TPM-010 based on SANS 5925: 2007
Volatiles (% _{wt})	21.73	22.89	21.43	22.06	22.35	ACT-TPM-012 based on ISO 562: 2010
Ash yield (% _{wt})	29.10	30.80	28.00	28.40	28.10	ACT-TPM-011 based on ISO 1171: 2010
Fixed carbon (% _{wt})	48.10	45.70	49.50	48.50	48.60	By difference
Gross calorific value (MJ/kg)	22.3	21.7	22.8	22.5	22.6	ACT-TPM-014 based on ISO 1928: 2009

A density profile of each particle size range was obtained by doing a float-sink analyses using zinc chloride solutions, for a range of 1.5 to 1.8g/cm³ (SANS 7936:2010). From this, the separability index (D_x) was calculated to quantify the amount of near dense material between the limits of the specified relative density values (Van der Walt, 1984). This refers

to the mass percentage of material within a ± 0.1 range around the proposed cut point density (x). A D_x value exceeding 25 is indicative of a more challenging separation, whereas a low D_x (below 25) index refers to an easier separating coal (Van der Walt, 1984). Table 3.1-3 shows the degree of difficulty for each size fraction at a number of proposed cut point densities. In general, this coal can be considered a difficult coal to beneficiate, which is typical of a Southern Hemisphere coal.

Table 3.1-3: Degree of difficulty of washing of the coal samples

PSD (mm)	$D_{1.5}$	$D_{1.6}$	$D_{1.7}$	$D_{1.8}$
PSD-01 (+5.6-6.7)	37	30	22	28
PSD-02 (+6.7-8.0)	34	30	23	29
PSD-03 (+8.0-9.5)	37	27	19	30
PSD-04 (+9.5-11.2)	38	28	21	28
PSD-05 (+11.2-13.2)	35	29	22	33

Dense medium

Magnetite powder was used as a dense medium in this study. It was prepared by drying it to $<5\%_{wt}$ and screening to +300 μ m with a d_{50} size of 490 μ m. The true density of the magnetite was found by helium pycnometry (SANS 1014:1985) to be 4.8g/cm³. The magnetic susceptibility of the magnetite was 20mm and the particle shape, as determined by SEM photography, was pyramidal.

EQUIPMENT USED

Experiments were conducted in a batch operated air dense medium fluidized bed (ADMFB) unit that consisted of three sections namely the *airflow system*, *fluidized bed* and *dust control* section, as shown in Figure 3.1-2. The *airflow system* consists of an air blower, airflow sensor (pitot tube) as well as an air-distribution compartment coupled to an air distribution plate. The apertures in this plate had an inner diameter of 0.004m and a pitch of 0.006m. Eight rectangular frames (0.3m x 0.3m x 0.05m) manufactured from clear PVC were stacked on top of the distribution plate and clamped together to form a fluidizing unit having a height of 0.4m. Each layer could be detached and removed separately from the unit to aid in obtaining a sample cut of material at a certain height. The unit was extended at the top by a further 0.3m to allow sufficient space for the particles to fluidize. The *dust control* section of the ADMFB unit is composed of a tapered trapezoidal extension (20° hopper angle), covered

by a 100 μ m stainless steel wired mesh. This limited the carry-over of ultra-fine particles by a reduction in air velocity at the top of the unit. A 0.18kW oscillating vibration motor was mounted onto the outside structure of the unit, which could induce vibration to the bed. The frequency and amplitude range provided by the vibration motor is 15Hz and 0.7-1.0mm, respectively.

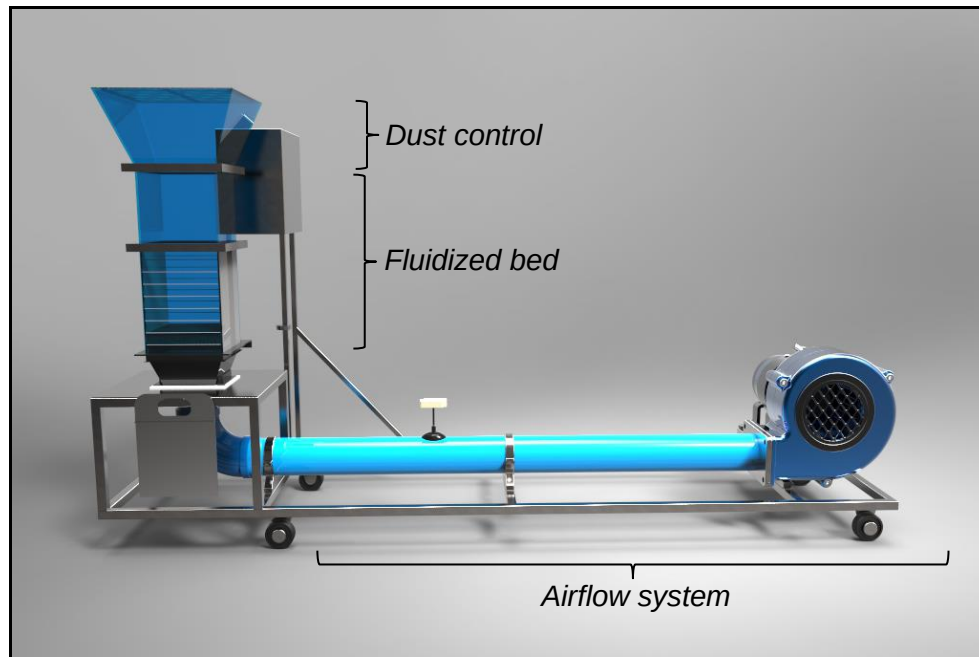


Figure 3.1-2: Three-dimensional illustration of the ADMFB

A mixture of coal and magnetite, in predetermined mass ratios, was placed inside the *fluidized bed* section. The load was fluidized with air at a fluidization velocity (U^*) of 1.0-1.1 times the predetermined minimum fluidization velocity (U_{mf}) of each load for a duration of 10 minutes. After fluidization, the frame sections were dismantled from the top (layer 8) to the bottom (layer 1). The contents of each layer were removed and coned and quartered until a correctly sized representative sample could be obtained for ash yield, calorific value and density analysis. The results obtained from these analyses were represented in terms of bed height and indicated the stratification of coal along the height of the ADMFB.

Results and discussion

Control run

Due to the vast amount of data obtained during this study, only some representative experimental runs will be shown and discussed in detail. PSD-05 (DM 0:1), vibration deactivated (VD) was selected for explanatory purposes due the clarity of the general trend

observed for this setup. The ratio of dense medium to coal (DM 0:1) indicates that no magnetite was added to the system and therefore only pure coal was used. *Figure 3.1-3* shows the ash yield of the different bed layers, from bottom to top, as well as the feed ash value (represented by the vertical line).

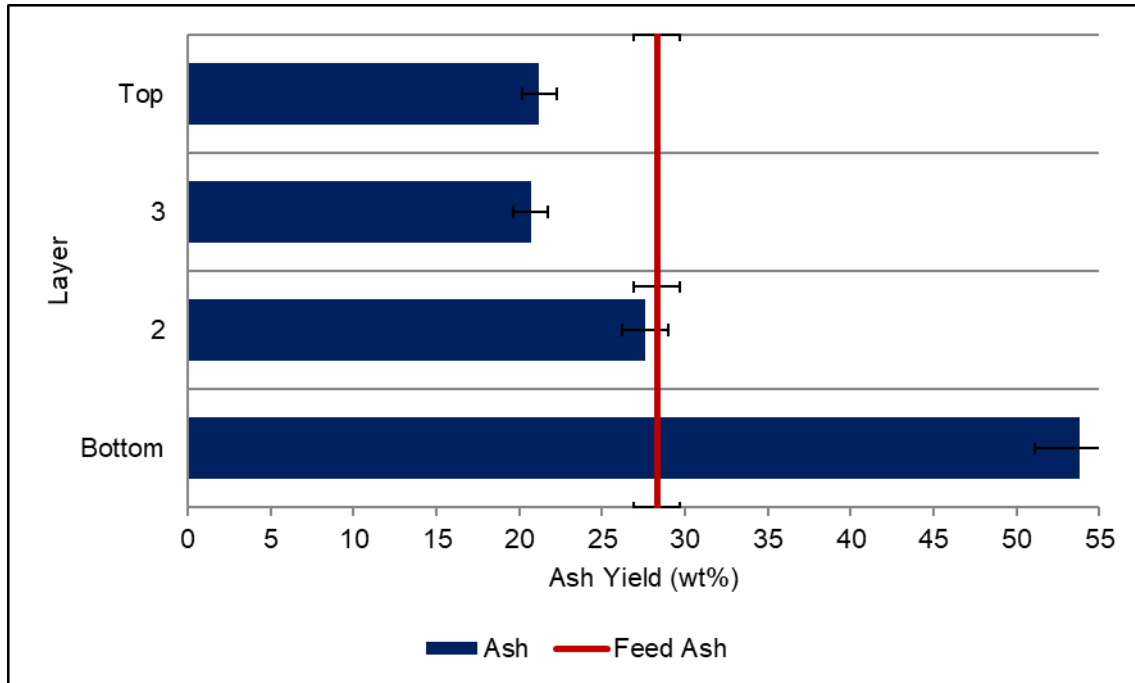


Figure 3.1-3: Ash values of control run according to bed layers (PSD-05; DM 0:1; VD)

The feed coal, shown in *Figure 3.1-3*, had ash yield and CV of 28.3%wt and 21.8MJ/kg respectively. After fluidization, the ash yield of the bottom layer almost doubled to 53.8% with a minimum ash yield in top two layers close to 21%, which is indicative of a de-stoning action occurring within the bed. The corresponding CV value obtained for the top layer was 24.9MJ/kg. It was further possible to identify a middling section which yielded ash yield and CV values very close to that of the original feed coal. Coal stratification across the four layers was therefore clearly visible, even without the using a dense medium to fluidize the coal.

Variables consideration

Numerous variables were investigated during this study and the influence of particle size range, dense medium to coal ratio and vibration on the fluidization operation was determined. The response on the ash yield, CV and density of the coal reporting to the different layers in the bed was measured and reported for every change in each of the mentioned variables.

Effect of particle size range

The effect that changes in the feed coal particle size has on the minimum (U_{mf}) and operating fluidization (U) velocities of the bed is briefly discussed in this section. More specifically, PSD-05, PSD-03 and PSD-01 were selected and the respective bed masses, U_{mf} and U values are summarized in *Table 3.1-4*.

Table 3.1-4: Minimum (U_{mf}) and operating (U) fluidizing conditions for PSD-05, PSD-03 and PSD-01

PSD (mm)	Bed mass (kg)	U_{mf} (m/s)	U (m/s)
PSD-05 (+11.2-13.2)	15.06	15.1	16.1
PSD-03 (+8.0-9.5)	15.35	14.7	14.8
PSD-01 (+5.6-6.7)	15.20	9.2	10.4

It is apparent from the results in *Table 3.1-4* that an increase in U_{mf} is prompted by an increase in PSD of the feed coal. The reason for this is that a greater upward force is required to overcome the net weight of the larger individual particles in order to ensure full suspension of each particle; a larger minimum fluidization velocity is therefore required. For pure coal beds, an operating velocity in close range (1.1 to 1.3) to the minimum fluidization velocity delivered a stable suspended bed with minimal observed turbulence and back mixing. The response of the bed for PSD-05, PSD-03 and PSD-01 feed coal, operating it without dense medium (DM 0:1) or the addition of vibration (VD), are summarized in *Table 3.1-5*.

Table 3.1-5: Ash, CV and density value comparison of PSD-05, PSD-03 and PSD-01

PSD (mm)	Layer	Ash Yield (% _w)	CV (MJ/kg)	Density (g/cm ³)
PSD-05 (+11.2-13.2)	<i>Feed</i>	28.3	21.8	1.61
	Top	21.6	25.0	1.56
	Bottom	36.2	19.3	1.72
PSD-03 (+8.0-9.5)	<i>Feed</i>	30.6	21.3	1.66
	Top	24.0	24.0	1.54
	Bottom	41.6	17.1	1.84
PSD-01 (+5.6-6.7)	<i>Feed</i>	33.4	20.4	1.73
	Top	23.1	24.3	1.59
	Bottom	43.2	16.5	1.89

Comparable results were acquired for the bottom and top layers for PSD-05, PSD-03 and PSD-01 in *Table 3.1-5*. The degree of separation between the top and bottom layers remained similar for all feed PSDs, although it does show that the exact ash yield values of the layers are dependent on the characteristics of the feed coal. The ash values in the

bottom layer increases in accordance with the feed ash values, which may be indicative of an increased fraction of dense particles present in the feed relocating to the bottom layer.

Performance curves were drawn up to illustrate the coal yield obtained for the cumulative ash percentage in the product. The results of the performance curves of PSD-01, PSD-03 and PSD-05 are compared to the float-sink test results for the coal sample, represented by the solid line in *Figure 3.1-4*. The float-sink performance curve is defined as the yield versus the cumulative ash percentage and serves as a benchmark for the best possible performance obtainable for the given coal samples.

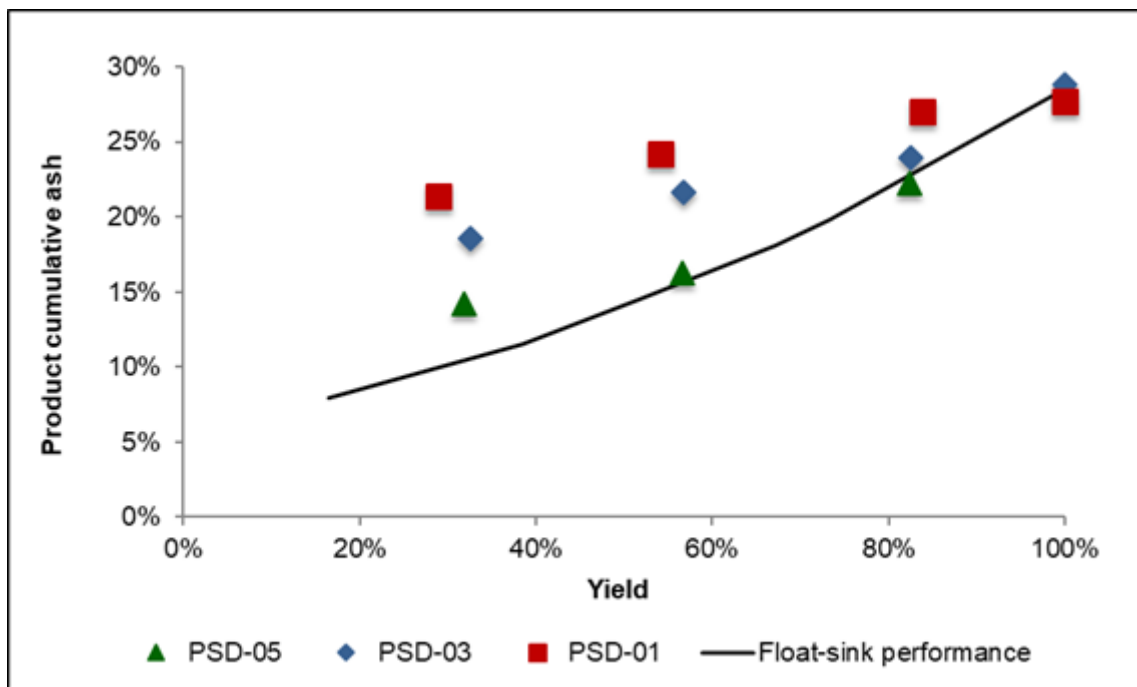


Figure 3.1-4: Comparison of bed performance curves for various PSD's (DM 0:1; VD)

The curves in *Figure 3.1-4* confirmed that density separation of particles in the ADMFB are easier when larger particles are fed to the bed. For the most part, PSD-05 produced results that narrowly simulated the float-sink curve, which is indicative of good separation performances. For a chosen product ash of 22%, yields of 30%, 65% and 85% were achieved for PSD-01, PSD-03 and PSD-05 respectively. Yang *et al.* (2015) found that a decrease in PSD returned poor separation results, which is confirmed by the performance of PSD-01 observed in *Figure 3.1-4*.

Effect of dense medium

The response of the bed for varying dense medium to coal ratios was measured to determine the influence of the magnetite on the bed's performance. The experimental runs

that did not use dense medium were loaded with coal only and referred to as DM 0:1. For the experimental runs using dense medium in various ratios, the dense medium and coal were weighed and mixed before being placed in the bed. For a DM 1:1, 7.5kg magnetite and 7.5kg coal was mixed whereas for DM 2:1, 5kg magnetite and 10kg coal was mixed. The mixing of the sample prior to loading is necessary to ensure that the coal and magnetite are uniformly distributed across the bed. The coal ash yield reporting to the different layers after fluidization, for the experimental runs of PSD-05, DM 0:1, 1:1 and 2:1 is comparatively summarized in *Table 3.1-6*.

Table 3.1-6: Comparison of dense medium ratio of +8.0mm-9.5mm coal

Ash Yield (%_{wt}) (Vibration deactivated)			
Layer	DM 0:1	DM 1:1	DM 2:1
<i>Feed</i>	30.6	26.9	27.9
Top	20.3	24.0	24.0
3	24.0	27.0	-
2	31.1	28.8	25.6
Bottom	54.3	34.0	32.1
Ash Yield (%_{wt}) (Vibration activated)			
Layer	DM 0:1	DM 1:1	DM 2:1
<i>Feed</i>	28.1	26.6	24.2
Top	24.0	23.2	23.7
3	24.0	-	-
2	26.3	24.5	23.5
Bottom	41.7	27.2	29.7

Table 3.1-6 shows marginal differences in the coal ash yield reporting to the top layer of the bed after fluidization. However, noticeable differences in ash yield values for the bed's bottom layers were measured after completions of the different runs. The pure coal bed (DM 0:1), produced a 54.3%_{wt} ash yield in the bottom layer of the bed, which was the highest of the beds compared in *Table 3.1-6*. A similar trend was produced by all the particle sizes investigated, even when vibration was added to the system. When dense medium was added to the bed, a more uniform cut density was created within the bed, leading to an equal density distribution in particularly the bottom section of the bed. Due to the difficulty in separating the coal, this causes the coal in the bottom half of the bed to be comparable in density.

Furthermore, it is important to note that the coal to dense medium ratio influences the minimum fluidization velocity of a bed. A fluidized bed with dense medium feed creates a suspension of magnetite particles while the coal particles simply float or sink according to the difference in relative density. A pure magnetite bed (DM 1:0) with a mass of 15kg was examined and the minimum fluidization velocity (U_{mf}) thereof was found to be 8.0 m/s, giving a bed density of 1.31 g/m³. Therefore, if the static height of the dense medium beds is kept relatively constant, no drastic variances in the minimum incipient point should be observed. Table 3.1-7 contains the minimum fluidization velocity (U_{mf}) values of PSD-05 and PSD-01 coal as well as the operating velocities (U), both without the addition of vibration.

Table 3.1-7: U_{mf} (m/s) and U (m/s) of PSD-05 (+11.2mm-13.2mm) and PSD-01 (+5.6mm-6.7mm) coal

DM ratio	Bed mass (kg)	Bed height (cm)	U_{mf} (m/s)	U (m/s)
+5.6mm-6.7mm				
0:1	15.20	20	9.2	10.4
1:1	15.00	15	7.2	8.3
2:1	15.00	15	5.2	6.9
+11.2mm-13.2mm				
0:1	15.06	20	15.1	16.1
1:1	15.00	15	8.6	9.8
2:1	15.00	15	9.2	10.5

The information provided in Table 3.1-7 shows that by introducing dense medium to the system, a reduction in bed height was observed, for both PSD-01 and PSD-05, and as a result a reduction in minimum fluidization velocity. The bed density is essentially increased as dense medium is added to the bed, however as a constant feed mass is maintained, the volume of the bed decreases and along with this the bed height. In conclusion, it is apparent that the minimum fluidization velocity (U_{mf}) for beds containing magnetite falls within a narrow range from the 8.0m/s of the pure magnetite bed.

The influence of dense medium is also expressed in Figure 3.1-5 in terms of performance curves for PSD-01 (+5.6mm-6.7mm). It is evident from the results in Figure 3.1-5 that the experimental runs conducted in the absence of dense medium (DM 0:1) compared the best with conventional float-sink data. Deviations from this line are observed for beds containing magnetite as a dense medium.

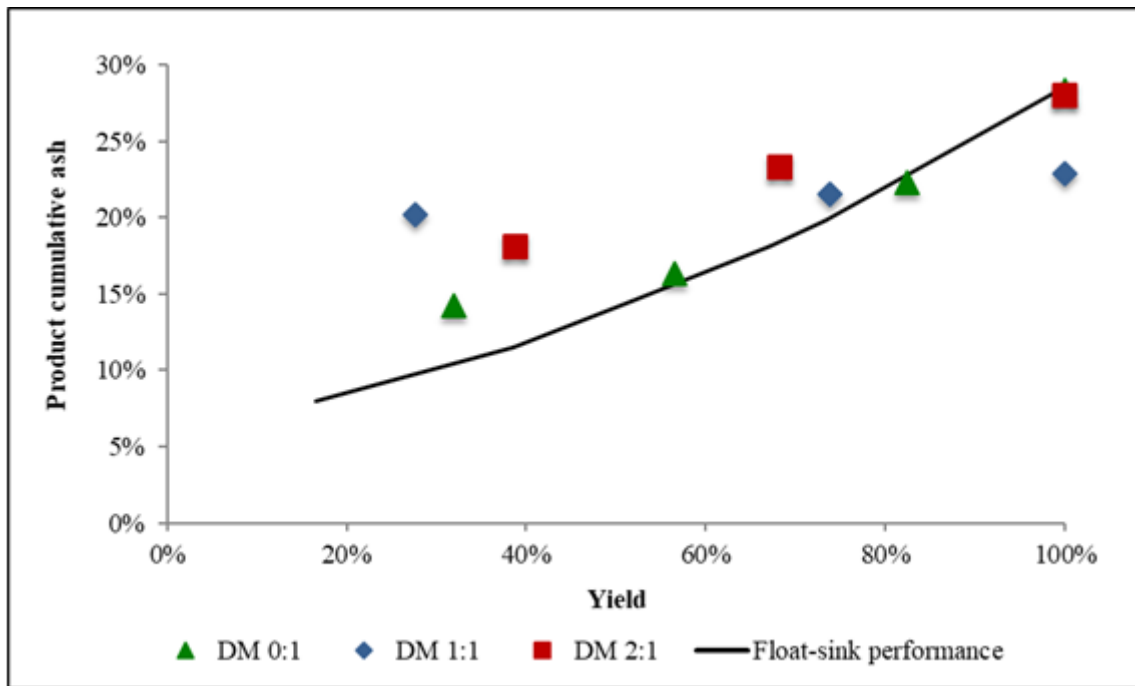


Figure 3.1-5: Comparison of bed performance curves for various DM (PSD-01; VD)

When taking into account the performance curves for DM 0:1 to DM 2:1, some deviances from the float-sink performance curve in *Figure 3.1-5* was noticed. The DM 0:1 curve in *Figure 3.1-5* generated lower ash yields in comparison to the DM 1:1 and 2:1 bed. A similar trend was observed for numerous experimental runs for varying ash differentials. This may be because of the poor stratification produced by the DM 1:1 and 2:1 throughout the bed, which resulted in a lesser separation performance. A bed operated without dense medium can therefore deliver a similar, and even better, density separation than the one's operated conventionally by adding magnetite.

Effect of vibration

The effect of vibration was tested by adding a horizontal vibration motion to the ADMFB. By adding vibration, the minimum fluidization velocity was lowered, and fluidization stability was improved due of sufficient particle-particle movement and bubble formation (He *et al.*, 2015). The influence of vibration on the fluidization velocities of PSD-01 (+5.6mm-6.7mm) and PSD-05 (+11.2mm-13.2mm) coal studied is provided in *Table 3.1-8*.

Table 3.1-8: Comparison of minimum and operating fluidizing conditions for varying vibration

Vibration	Fluidization (m/s)	DM 0:1	DM 1:1	DM 2:1
+5.6mm-6.7mm				
Deactivated (VD)	U_{mf}	9.2	6.2	4.7
	U	10.4	8.3	6.9
Activated (VA)	U_{mf}	8.0	5.4	7.8
	U	9.2	6.7	9.4
+11.2mm-13.2mm				
Deactivated (VD)	U_{mf}	15.1	7.8	6.7
	U	16.1	9.8	10.5
Activated (VA)	U_{mf}	14.7	6.1	6.2
	U	15.3	7.5	7.7

A definite decrease in minimum fluidization velocity is marked for both particle size ranges shown in Table 3.1-8, when initiating vibration. Figure 3.1-6 illustrates the response of the different bed layers in terms of coal ash yield when vibration is activated (VA) and deactivated (VD) respectively, for the PSD-05 (+11.2mm-13.2mm) (DM 1:0).

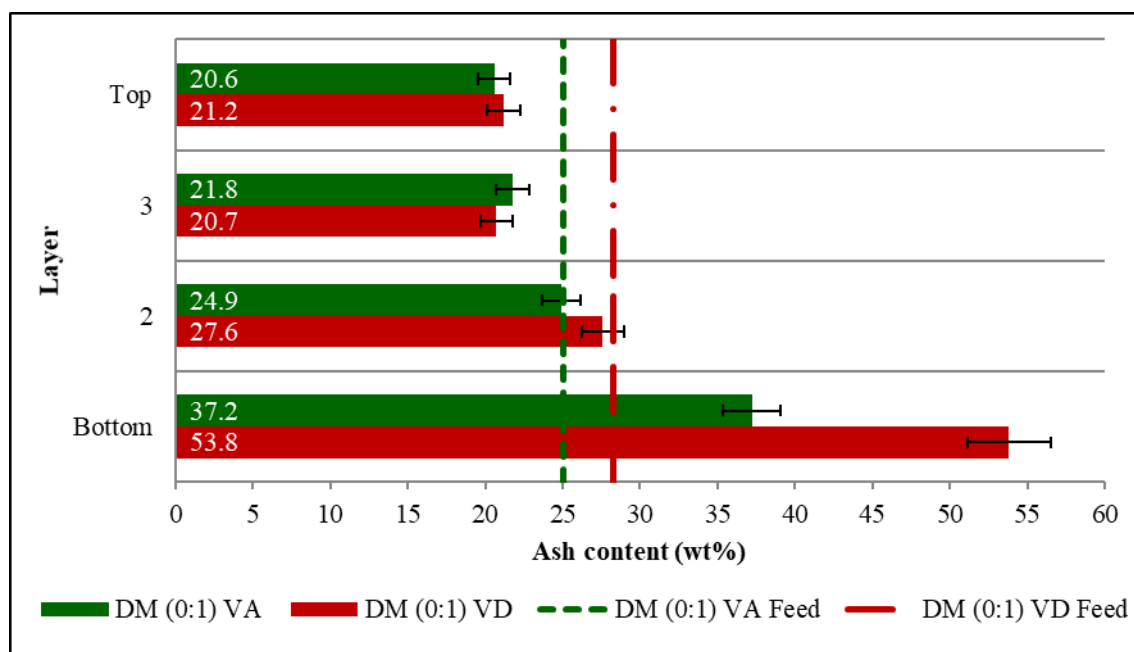


Figure 3.1-6: Comparison of vibration (PSD-05; DM 0:1)

Figure 3.1-6 shows stable bed performance and comparable ash yields over the top three layers when operated with and without the addition of vibration. There is however a significant difference in the ash yield of the coal reporting to the bottom layer for both the tested operational variables. The large difference in ash yield values observed in the bottom

layer of the bed may be a result of the manner in which the inter-particulate voids are blocked and the movement of particles through the bed layers is hindered, specifically referring to when vibration is added. The additional vibratory movement inside the bed causes the finer particles to percolate downwards after the airflow to the bed was turned off, resulting in a more tightly packed bed at the bottom and hence blocking larger, denser particles from moving down. The results do however not conform to findings reported in literature and will mostly be eliminated with continuous operations. Literature reports that the addition of vibration to the ADMFB during operation leads to a more stable fluidized bed, which improves stratification of coal particles amongst the bed height (Yang *et al.*, 2013a and He *et al.*, 2015).

Figure 3.1-7 portrays the performance curves calculated from the results of PSD-05, running with vibration activated (VA) and vibration deactivated (VD), in the absence of dense medium. It is evident from the performance curves in Figure 3.1-7 that the run operated without vibration (VD) performed remarkably better than with vibration. Similarities were observed when magnetite was added to the bed as dense medium, irrespective of the magnetite to coal ratio. The batch wise operation of this particular bed must always be kept in mind when reading these curves. Notable though is the high performance of the bed in the absence of any dense media or additional vibration. It proves that operation of the bed at lower costs results in comparable and even better results than during assisted operations.

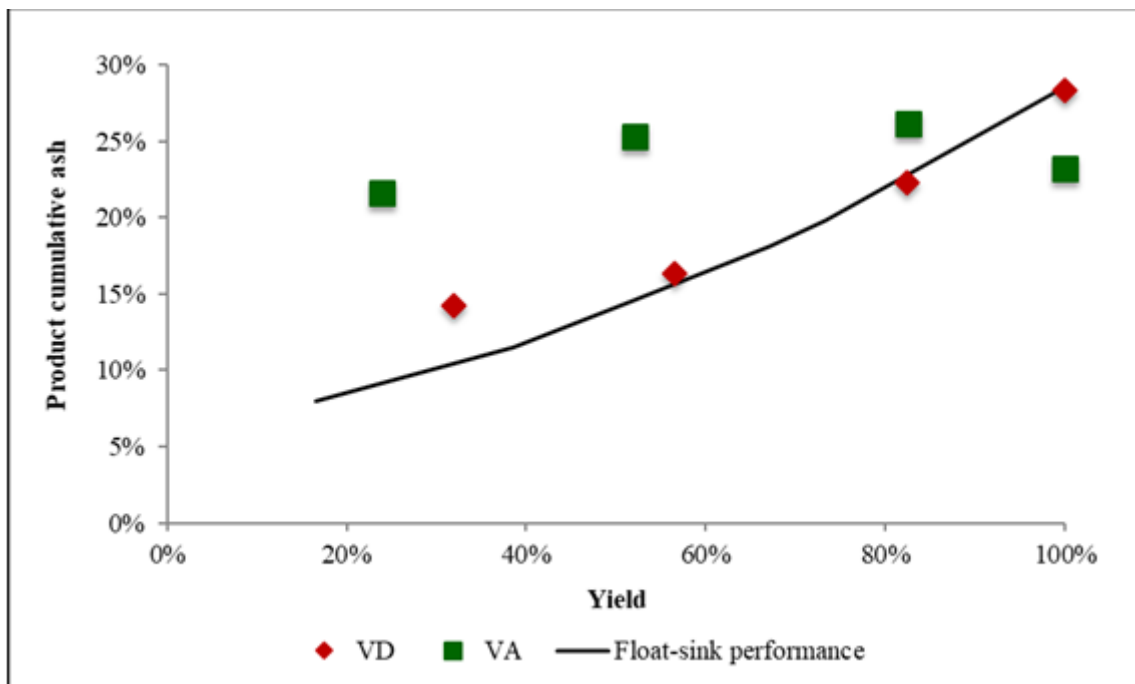


Figure 3.1-7: Comparison of bed performance curves for vibration deactivated (VD) and vibration activated (VA) (PSD-05; DM 0:1)

Conclusion

The aim of this paper was to study beneficiation of small South African coal by employing an air dense medium fluidized bed. It was found that the specific ADMFB tested was able to separate the coal successfully according to density. The denser coal particles were displaced to the bottom and the less dense coal (better quality) moved to the top layers of the bed. The best overall cumulative products obtained by discarding the bottom bed layer (cut point height of 50mm from the top of the bed) are provided in *Table 3.1-9*. These results were all obtained for ADMFB experimental runs that employed no dense medium (DM 0:1) and no addition of vibration (VD).

Table 3.1-9: Best cumulative results obtained from study

PSD (mm)	Ash yield (% _{wt})	CV (MJ/kg)	Density (g/cm ³)	Mass yield (% _{wt})
PSD-05 (+11.2-13.2)	14.2	24.1	1.56	31.9
PSD-04 (+9.5-11.2)	13.0	25.0	1.54	27.6
PSD-03 (+8.0-9.5)	14.9	23.7	1.56	34.2
PSD-02 (+6.7-8.0)	13.7	24.0	1.57	27.9
PSD-01 (+5.6-6.7)	17.6	22.7	1.64	28.9

The ash and mass yield results in *Table 3.1-9* proves that the ADMFB is able to produce low cumulative ash yields at a reasonably low mass production.

It was established that particle size range has an influence on the stratification obtained within the ADMFB as larger particle sizes produced a higher ash rejection compared to the smaller particle size ranges. Similar ash values were achieved for all respective bed layers of the particle size ranges investigated. A reduction in particle size range did however have an appreciable effect on the minimum fluidization velocity required by the bed, in that a decrease in minimum fluidization velocity was experienced. As discussed, this reduction is a result of the reduced lower buoyant force required to suspend the particles.

The results of the beds without medium were more promising, while the coal bed with magnetite yielded mediocre ash rejection results. The results of the top layers of DM 2:1 and DM 1:1 coal bed fall within a narrow range, with slightly lower ash values observed for the DM 0:1 coal bed. Moreover, high ash values were obtained in the bottom layers of the DM 0:1 coal bed, indicating that the DM 0:1 bed presented better de-stoning capabilities. Furthermore, by supplementing the coal bed with magnetite, a drastic reduction in minimum

fluidization velocity was observed. In a coal-magnetite bed, the magnetite is fluidized (in suspension) and the coal particles float or sink in accordance with density.

No significant improvement in the results in the various bed layers were observed when adding vibration to the system, however the ash values in the bottom bed layer was increased. This may be a result of the respective feed ash values.

In conclusion, it was established that the separation performance of the ADMFB with vibration (VA) was governed by the stratification in the two bottom layers. Vibration, in addition to increased quantities of magnetite, created an even more densely packed bed, which hindered the movement of coal particles.

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CHAPTER 4

Chapter 4 contains the paper titled “Dry beneficiation of +0.5mm-5.6mm South African coal using an air dense medium fluidization bed”. In this paper, air dense medium fluidized bed (ADMFB) is evaluated as a suitable dry beneficiation technique to upgrade **+0.5mm-5.6mm** South African coal. The separation efficiency of the ADMFB was assessed by determining the effect, if any, of particle size range, medium to coal feed ratios and the addition of vibration on the separation ability of the bed. The coal was upgraded to varying degrees, depending on the parameters of the applied methods.

Beneficiation of +0.5mm-5.6mm South African coal using an air dense medium fluidized bed (Paper 2)

This section provides information regarding contributions made by the various authors towards completion of the paper. General information regarding the paper submitted for publishing along with additional information of the journal to which the paper was submitted for publishing.

Contribution by authors

Misses E.S. Diedericks conducted a coal beneficiation research study with the guidance supervision of Prof. M. le Roux and Prof. Q.P. Campbell. The concept and scope of work for this research were largely formulated by Prof. M. le Roux and Prof. Q.P. Campbell. The air dense medium fluidized bed experimental setup was designed by Miss N Hughes and constructed by the North-West University technical support staff, with guidance from Miss N. Hughes. Misses E.S. Diedericks was responsible for conducting the experimental work, collecting data and processing of the collected data. The results from experimental runs were discussed and assessed during multiple consultations between Misses E.S. Diedericks, Prof. M. le Roux and Prof. Q.P. Campbell. After compilation review cycles, the paper was submitted for publishing in a peer-reviewed journal, Journal of the Southern African Institute of Mining and Metallurgy (SAIMM). The paper was written by Misses E.S. Diedericks and reviewed and edited by Prof. M. le Roux as well as Prof. Q.P. Campbell.

General information

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4.1 Dry beneficiation of +0.5mm-5.6mm South African coal using an air dense medium fluidization bed

ES Peters, M le Roux, QP Campbell*

School of Chemical and Minerals Engineering, North-West University, Potchefstroom, South Africa.

***marco.leroux@nwu.ac.za**

Abstract

This paper presents the separation performance of solid phase bed material, at various particle size ranges (-5.6mm+0.5mm South African coal particles), in an air dense medium fluidized bed (ADMFB).

The coal particles were separated into +0.5mm-1mm, +1mm-2mm, +2mm-2.8mm, +2.8mm-4mm, +4mm-4.75mm and +4.75mm-5.6mm PSDs (particle size distributions) and fed to the fluidized bed in these fractions. Along with the six coal particle size ranges, three dense media to coal ratios and the addition of vibration was tested to identify the best operating conditions. The separation performance of the bed was measured by ash content (%wt) and calorific value (CV) as well as minimum fluidization velocity (U_{mf}). Acceptable results were obtained for larger particle size ranges down to and including +2.0mm-2.8mm coal particles, after which the separation performance decreased significantly. Density stratification was irregular and not obvious for coal particles below 2.0mm and maintaining a consistent fluidization state also proved to be challenging, especially when dense medium was added. The coal particles separated vertically along the bed height because of differences in particle and bed density, while particle size proved to have a notable influence on the degree of separation. An air fluidization velocity of between 1.1 to $1.4U_{mf}$ was shown as the best performing velocity, which yielded the maximum ash differential between the top and bottom layers of the bed for all the particle size ranges tested.

For +2.0mm-5.6mm coal particles, low cumulative ash yields were obtained at high mass yields; however, the ash yields increased for -2mm coal. Vibration and dense medium have, in some cases, enhanced the separation efficiency of the ADMFB. The -2.0mm particles experienced stronger particle-particle interactions as well as elevated levels of bubbling and back mixing than that of the +2.0mm particles, which explains the poor performance of the small particle sizes.

Keywords: ADMFB, coal beneficiation, dense medium, magnetite, air dense medium fluidization, minimum fluidization velocity, ash yield.

Introduction

Coal preparation plants most commonly employ wet separation processes to clean coal (Luo *et al.*, 2006). However, water reserves are unevenly distributed worldwide, with approximately two-thirds of the coal reserves located in the arid regions (Luo *et al.*, 2006). South Africa is an established coal-producing country that is subjected to periods of draught, and it is therefore challenging to source reliable process water. Companies that supply coal need to ensure that the coal meets the specification for the thermal-, metallurgical- or export markets. To ensure that feedstocks adhere to these specifications, coal are washed using mainly wet dense medium processes. These processes are known to produce low ash products at high yields; however, it is water intensive and ends up with a wet product that is costly and difficult to dewater (Strydom *et al.*, 2016).

South Africa is a water-scarce country and as a result, government enforced water regulations have become more stringent. Slurry and coal fines associated with coal processing plants, are both difficult and expensive to dispose of and manage (De Korte, 2013). As a result, research has shifted to developing dry-based coal beneficiation processes that operate at high separations efficiencies (Luo *et al.*, 2006). Air dense medium fluidized beds (ADMFB) have grown popular in recent years and are known for producing comparatively high separation efficiencies with the added benefit of eliminating the use of water. An ADMFB is a dense gas-solid system that is characterized by a high bed density and micro-bubbles percolating from the bottom of the bed to the top (Wang *et al.*, 2015). The feed material is stratified according to density along the height of the bed resulting in the less dense material (clean coal) reporting to the top and more dense material (gangue) sinking to the bottom (Zhao *et al.*, 2010).

According to a study done by Ghosh (2013), air-based density separators, such as the FGX compound dry coal separator, is known for effectively treating +6mm-75mm ROM (run-of-mine). Moreover, a considerable decline in separation performance was experienced for -6mm coal particles. It was found that -6mm coal particles does not segregate well in these air-based density separators with particles generally reporting to the product layer regardless of particle density. As a result, the demand for modifying dry density separators to treat -6mm coal has increased significantly (Ghosh, 2013).

Therefore, this study entails investigating and reporting on the beneficiation performance of +0.5mm-5.6mm South African coal in an ADMFB.

Background

Coal deposits in South Africa have been mined commercially since the late 1800's (Falcon & Ham, 1988; Peatfield, 2003). Over the past 15 years, the coal production in South African has hovered between 250 to 260 Mt/annum, peaking at 261 Mt in 2014 (CoM, 2016). Collectively, the Witbank, Highveld and Waterberg coalfields holds more than 70% of South Africa's coal reserves (Jeffrey, 2005). Typically, the ash yield of run-of-mine coal from these fields in South Africa ranges between 20-30%_{wt}; which is expected to increase with the depletion of high-quality seams (Dikgwatlhe, 2018). To negate the impurities, beneficiation of high ash coals is required. Both dry and wet beneficiation methods separate particles based on the difference in density between the high value particles and the impurities. This is achieved by developing a stable bed of specific density, using a medium like magnetite that will stratify all particles fed to the bed and hence separate it at a controlled density cut point.

The movement of particles inside the fluidized bed is based on its response to gravity and the resistance to buoyance caused by the upwards movement of an induced fluid (Wills & Napier-Munn, 2006). The particles with a relative density lower than that of the pseudo fluid will float and those that are heavier will sink (Wills & Napier-Munn, 2006). During recent years, an extensive number of research studies for developing and industrializing high separation efficiency dry beneficiation technology with minimal engineering and economic limitations have been ongoing. Substantial progress has been made with the implementation of an air dense medium fluidizing bed (ADMFB). During ADMFB operation, a steady, uniform air-solid suspension is established as separation medium (pseudo fluid). The density and volume fraction of the solids suspended in the air is used to control the density of this pseudo fluid. Results from various investigations on ADMFB are provided in *Table 4.1-1* and it is apparent that the separation efficiency decreased according to a reduction in particle size. In particular, the study conducted by Luo et al. (2003) shows a significant increase in E_p value (error of separation, called E_{cart} probable or E_p) when the particle size is decreased, indicating the ADMFB separation performance suffers with particle size reduction.

Table 4.1-1: Summary of experimental investigations on ADMFB, adapted from Mohanta *et al.* (2011)

Investigators	Bed type and geometry (cm)	Gas velocity (cm/s)	Medium size (μm)	ρ_{medium} (g/cm^3)	H_{bed} (cm)	Coal particle size (mm)	Separation time (s)	Ep
Mak <i>et al.</i> (2008)	Cylindrical (20)	5.3-6.8	208	5.1	-	22.6-5.66	480	-
Luo <i>et al.</i> (2003)	Rectangular vibrated (200x8x20)	1.65	58.2	-	7.5	6-0.5	-	0.07
Choung <i>et al.</i> (2006)	Cylindrical (4)	23.5	300-150	5.2	2.6	5.66-3.35	120	0.03
Luo <i>et al.</i> (2008)	Rectangular vibrated (10x10x22)	1	74-43	4.35	7.5	6-0.5	20	0.07
Beeckmans <i>et al.</i> (1982)	Counter current cascade, rectangular (548x19x60.9)	21.4	170		18.5	3-1	-	0.21
Luo <i>et al.</i> (2003)	Magnetically stabilized Cylindrical (10)	-	74-43	4.2	-	6-3 3-1 1-0.5 0.5-0.3	-	0.05 0.06 0.09 0.22
Kozanoglu <i>et al.</i> (1993)	Cylindrical (15.2)	4.14	150-124	-	7	0.29-0.17	60	

Mohanta *et al.*, (2011) conducted a comprehensive experimental study on the applicability of ADMFB for cleaning high-ash Indian thermal coal and it was concluded that the performance of the ADMFB separator was poor for the -13mm feed coal. This is because the upward drag force that is required by a coal particle is reduced as the particle size is decreased; therefore, the probability of a low-density, coarse particle to pass through the dense medium and reporting the gangue fraction is smaller than that of a finer particle of similar ash content (Mohanta *et al.*, 2011).

Diedericks *et al.* (2020) studied the beneficiation of +5.6mm-13.2mm South African coal and was able to separate the coal successfully according to density using a verticle ADMFB. By discarding the bottom bed layer (cut point height of 50mm from the top of the bed), the best cumulative product was obtained and are provided in Table 4.1-2. These results were obtained for runs that had no dense medium (DM 0:1) and no added vibration (VD). PSD refers to the various particle size distributions investigated in the study.

Table 4.1-2: Best cumulative results obtained from study

PSD (mm)	Ash yield (% _{wt})	CV (MJ/kg)	Density (g/cm ³)	Mass yield (% _{wt})
+11.2-13.2	14.2	24.1	1.56	31.9
+9.5-11.2	13.0	25.0	1.54	27.6
+8.0-9.5	14.9	23.7	1.56	34.2
+6.7-8.0	13.7	24.0	1.57	27.9
+5.6-6.7	17.6	22.7	1.64	28.9

It was found that particle size range affected the stratification achieved in the ADMFB, as larger particles produced a higher ash rejection compared to the smaller particle size ranges. A reduction in particle size range had a significant effect on the minimum fluidization velocity required by the bed, in that a lower minimum fluidization velocity was experienced. The lower velocity was because of the reduced buoyant force required for particle suspension (Diedericks *et al.*, 2020).

Experimental

Coal used

Coal from the number four seam in the Witbank coal fields in Mpumalanga, South Africa, was used during this investigation. The structure of the Witbank coalfield is uncomplicated, making it one of the most significant coal-mining regions in South Africa (Jeffrey *et al.*, 2015). However, conventional collieries in this and related areas are expected to reach exhaustion in the near future. Bituminous thermal grade coal for both the local and export market is mined from seam four and therefore need to adhere to set specifications (Jeffrey, 2005).

The feed coal was used as received and prepared by air drying, crushing and screening it. According to a study conducted by Terblanche (2013), the particle surface moisture of coal required for effective fluidization should be about 5%_{wt}. The dried coal was spilt into the six particle size ranges listed in *Table 4.1-3* to obtain a representative sample for each particle size range. The proximate analysis and calorific value for each size interval were determined and the results are provided in *Table 4.1-3*.

Table 4.1-3: Average initial proximate and calorific values for each coal particle size range

Description	PSD-01 (+0.5mm-1.0mm)	PSD-02 (+1.0mm-2.0mm)	PSD-03 (+2.0mm-2.8mm)	PSD-04 (+2.8mm-4.0mm)	PSD-05 (+4.0mm-4.75mm)	PSD-06 (+4.75mm-5.6mm)	Standards
Moisture (% _{wt})	2.35	1.93	2.13	2.46	2.30	1.67	ACT-TPM-010 based on SANS 5925: 2007
Volatiles (% _{wt})	24.17	23.62	23.13	23.91	23.04	21.60	ACT-TPM-012 based on ISO 562: 2010
Ash yield (% _{wt})	27.46	29.16	30.72	30.79	29.99	32.59	ACT-TPM-011 based on ISO 1171: 2010
Fixed carbon (% _{wt})	46.02	45.29	44.02	42.85	43.66	44.14	By difference
CV (MJ/kg)	25.17	25.15	24.60	25.45	22.08	24.28	ACT-TPM-014 based on ISO 1928: 2009

A density profile of each particle size range was obtained by using the traditional method of densimetric analyses. During the process, zinc chloride was prepared at different densities and a sample was moved between these liquids to establish a density profile of that sample (SANS 7936:2010). For this study, the densities ranged from 1.3 to 1.8 g/cm³ for the +1.0mm-5.6mm coal. In addition to the zinc chloride densimetric analyses, an organic solution was used to complete the densimetric analyses on the +0.5mm-1.0mm coal. White spirits and tetrachloroethylene were prepared at densities ranging from 1.30 to 1.60 g/cm³ and solutions of tetrachloroethylene and tetrabromoethane ranging from 1.70 and 1.80 g/cm³ were prepared. Organic liquids are substantially more expensive compared to zinc chloride and demineralized water. From an economic perspective, it was decided to use the expensive organic liquids for +0.5mm-1.0mm coal particles only, as this size range only requires a small amount.

Figure 4.1-1 shows the cumulative float curves at the various prepared densities for each coal size interval.

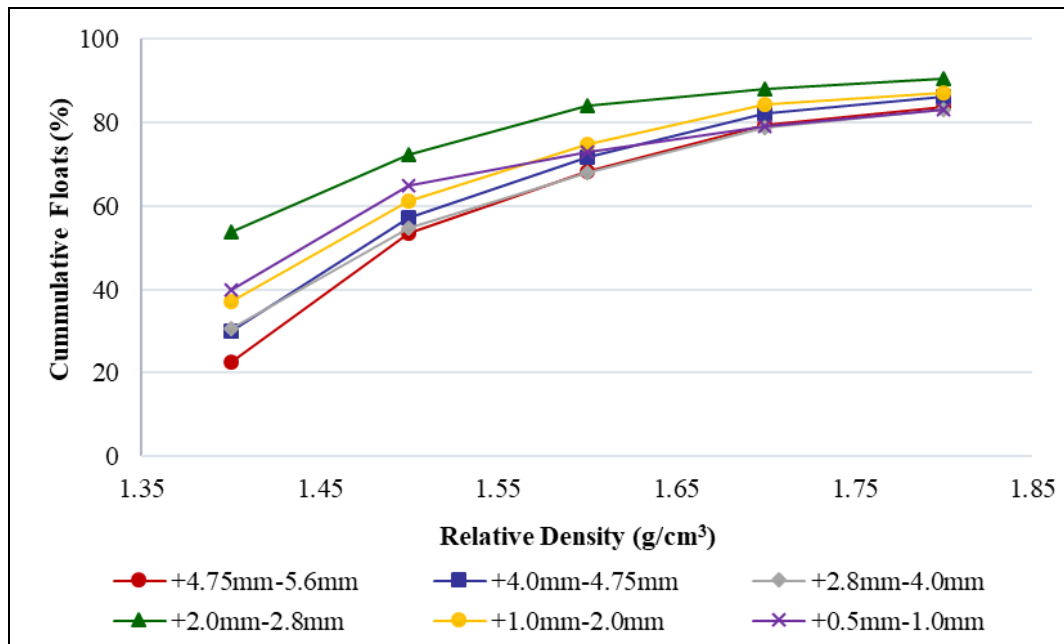


Figure 4.1-1: Density profiles for +0.5mm-5.6mm coal

Dense medium

Fine magnetite was selected as a dense medium for this study. The magnetite was prepared by air-drying it for 24-48 hours to obtain a moisture content below 5%_{wt} and then screened to >300 μ m (d_{50} of 490 μ m). The true density of the magnetite was found by helium pycnometry (SANS 1014:1985) to be 4.8 g/cm³. The particle shape of the magnetite, as determined by SEM photography, was pyramidal.

Experimental setup

An inhouse constructed air dense medium fluidized bed was operated in batch mode for the experimental runs. A three-dimensional rendering of the setup is shown in *Figure 4.1-2*. Three conjoined sections can be identified from the unit and will hereafter be referred to as (1) an *airflow system*, (2) *fluidized bed* and (3) *dust control* section. The airflow system was composed of an air blower, airflow sensor and an air-distribution compartment. Air from the blower was distributed evenly across the bed by a distribution device placed at the bottom, consisting of a packed bed of ceramic balls inserted between two distribution plates and a 100 μ m wire mesh. Eight rectangular frames (0.3m x 0.3m x 0.05m), constructed from clear blue PVC, was stacked on top of each other to form the 0.4m high *fluidized bed* section. A ninth fluidized bed layer (0.3m x 0.3m PVC layer), was added to the top of the preceding layers to allow enough space for the particles to fluidize. The bed was designed for the layers to be removable, allowing for variability in the cut-point along the height of the bed. At the top, a tapered trapezoidal extension (20° hopper angle) was added for *dust control*

purposes. The trapezoidal extension was covered by a 100µm stainless steel wire mesh to keep the particles inside the bed during operation.

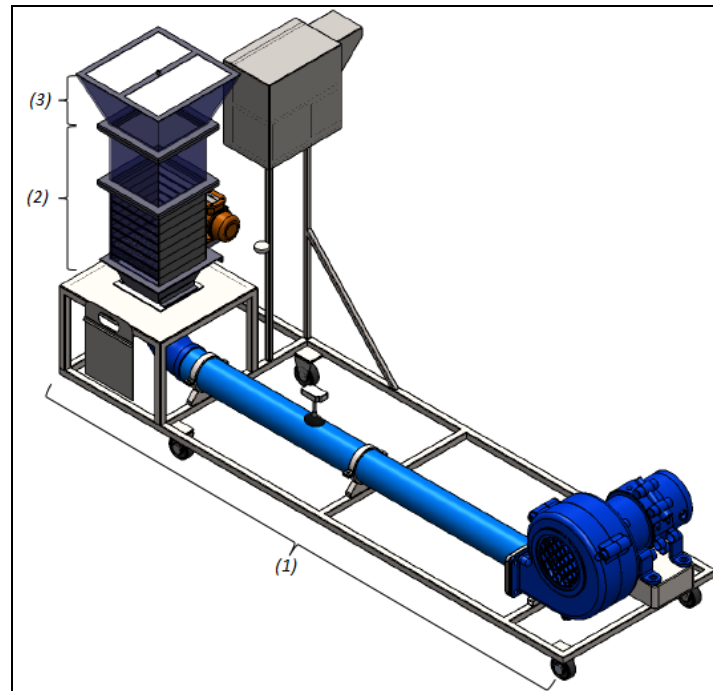


Figure 4.1-2: ADMFB with (1) airflow system, (2) fluidized bed system and (3) dust control section

The 20° incline angle of the trapezoidal extension provides an increase in total surface area of 0.35m². The ultra-fine particles that are formed from either mild attrition or breaking up of agglomerates report to a considerably higher point of the ADMFB compared to the larger particles. By moving into the trapezoidal extension area, the force by which an individual ultra-fine particle is fluidized as well as the velocity at which it travels is reduced. This effectively causes the ultra-fine particle to drop; however, as the mixture of air and particles flowing upward interferes with this process, the particles settle along the inside slope of the tapered extension. The stainless-steel wire mesh covering the top of the extension ensures any stray particles or particles that are unaffected by the surface area increase is retained in the system.

A 0.18kW oscillating vibration motor was affixed against the external structure of the ADMFB unit, which was used to induce vibration to the bed. The frequency and amplitude range of the vibration motor was 15Hz and 0.7-1.0mm, respectively. The fluidized bed section was loaded with a predefined amount of magnetite to coal. The load was then fluidized at 1.1 times the predetermined minimum fluidization velocity (U_{mf}), known as the operating velocity (U^*). After 10 minutes of fluidization, the air flow was switched off, allowing the bed to settle. Samples were taken from the bed by isolating each layer and removing all the particles

inside it. The content of each layer was coned and quartered until a representative sample for ash yield, calorific value and density analysis was obtained.

Results and discussion

Stratification of coal was quantified by analyzing the ash yield and calorific value for each sampled layer of the bed and comparing that with the feed coal. The change in these values along the bed height as compared to the feed, is an indication of the bed's ability to upgrade coal.

Influence of variables

Minimum fluidization velocity (U_{mf})

Distinguishing the state of transition from fixed bed to fluidizing regime is vital to understanding fluidization phenomena in ADMFB technology. This state is comprehensively investigated and reported on in literature for both theoretical and experimental approaches.

The minimum fluidization point of fluidization points for this study was theoretically obtained from first principles by using the Ergun equation. These results are however based on a single particle that is suspended in a fluid:

- +4.75mm-5.6mm – 6.27 to 16.56m/s
- +4.0mm-4.75mm – 4.44 to 11.91m/s
- +2.8mm-4.0mm – 2.18 to 8.45mm/s
- +2.0mm-2.8mm – 1.11 to 4.14mm/s
- +1.0mm-2.0mm – 0.28 to 2.11mm/s
- +0.5mm-1.0mm – 0.07 to 0.53mm/s

The minimum fluidization velocity ranges calculated may differ from the minimum fluidization velocity values determined experimentally. This is because of solid particles agglomeration, channeling, and electrostatic forces (generated by particle-particle friction or fluid-wall friction in the presence of elevated fluidization velocities) (Yang (2003), Kunii & Levenspiel (1991)). Minimum fluidization velocity by means of experimental approach can be determined by monitoring the total bed pressure drop (ΔP) vs. superficial velocity (u_0) as shown in Figure 4.1-3.

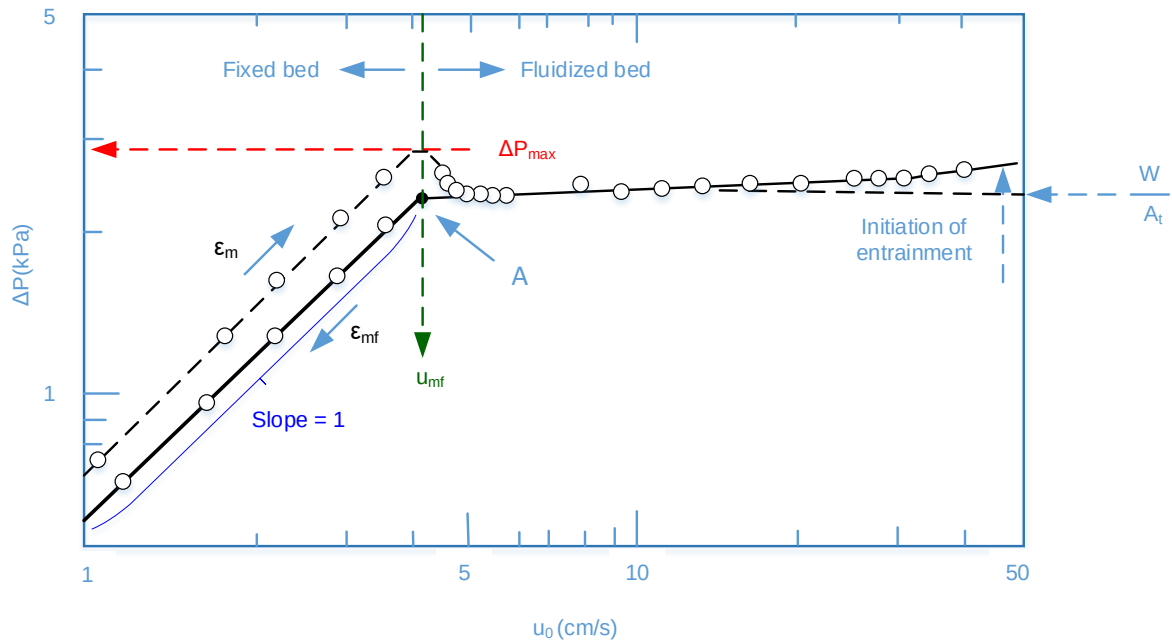


Figure 4.1-3: Pressure vs. velocity diagram, adapted from Kunii & Levenspiel (1991)

Similar graphs were drawn up during this study to determine the minimum fluidization velocity of the various PSDs. These can be determined experimentally by increasing the superficial air velocity incrementally until a point is reached where the static pressure of the bed is visibly lower than that of the bed pressure. Figure 4.1-4 shows the minimum fluidization points of +0.5mm-1mm (PSD-01), +1mm-2mm (PSD-02), +2mm-2.8mm (PSD-03), +2.8mm-4mm (PSD-04), +4mm-4.75mm (PSD-05) and +4.75mm-5.6mm (PSD-06) pure coal experimental runs.

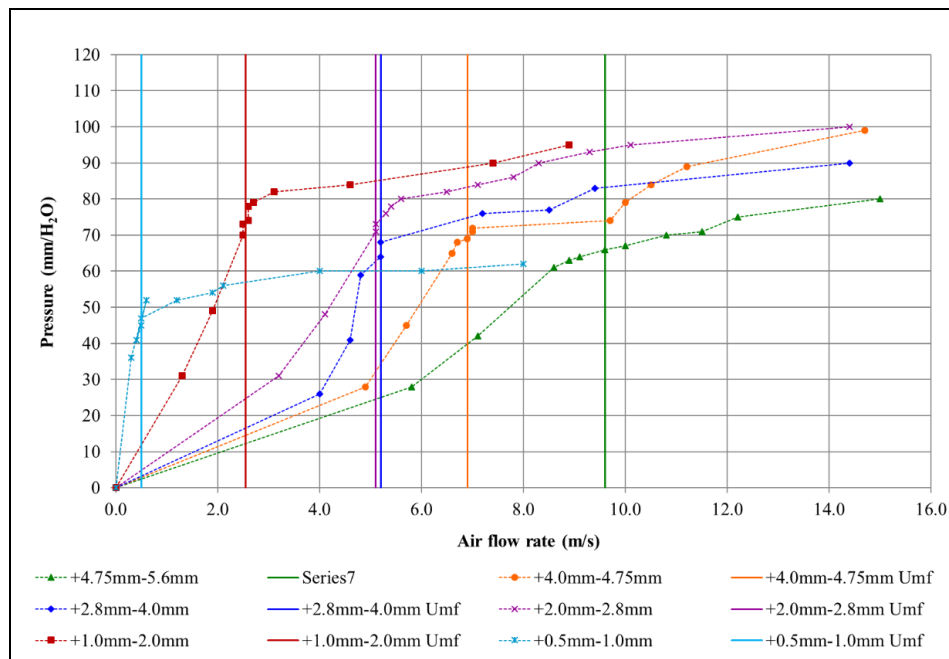


Figure 4.1-4: Experimental determination of minimum fluidization points for pure coal beds

Figure 4.1-5 shows a composite graph of curves showing various minimum fluidization points for +1mm-2mm (PSD-02), +2.8mm-4mm (PSD-04), and +4.75mm-5.6mm (PSD-06) experimental runs. These curves include those for pure coal runs, coal and dense medium (magnetite) runs as well as coal, dense medium (magnetite) and vibration added runs.

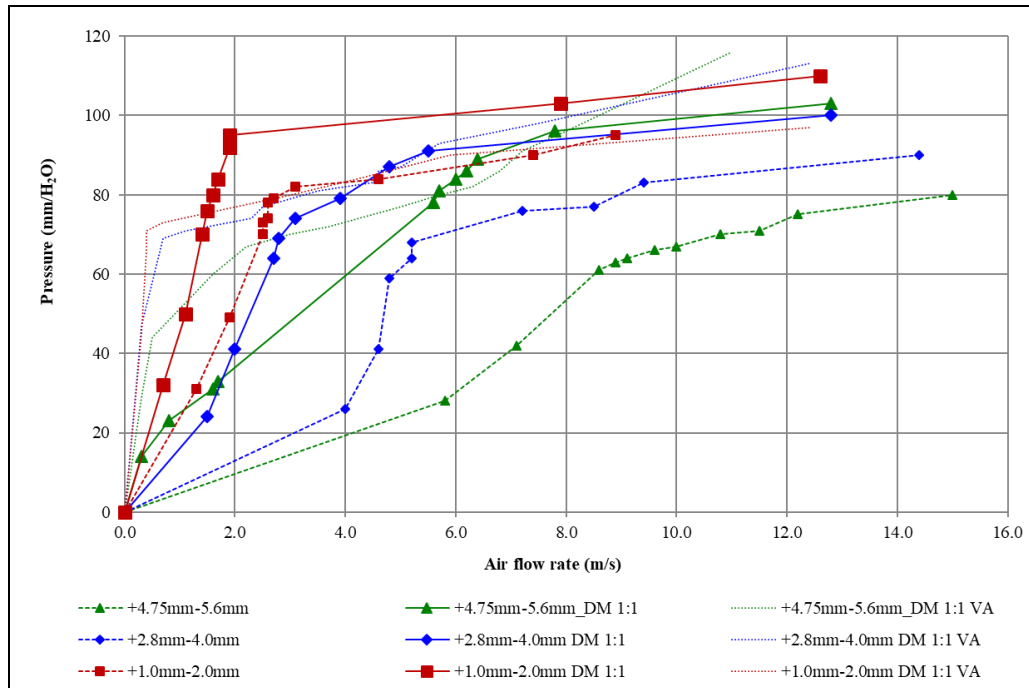


Figure 4.1-5: Minimum fluidization points determined for various experimental runs

The slope of the graph is increased with addition of dense medium and increased even further when vibration is added. The trend is reflected by all the particle sizes investigated during this study. This is due to the lubrication effect of the fine magnetite particles between the coarse coal particles which induces a smoother fluidization. This lubrication effect enhances movement and airflow percolation through the bed, which drives the system to reach minimum fluidization in a reduced amount of time (Richardson, 2002; Yang, 2003; Howard, 1989; Tang, 2009; He et al., 2003). Jin *et al.* (2004) investigated the influence of adding vibration to a 148 mm vertical cylindrical bed while studying the fluidization performance and segregation characteristics of +0.1mm-6 mm coal (density range between 1.3 and 2.66 t/m³). From the study, it was concluded that vibration aids in fluidization of medium by increasing the incidence of gas-solid contact and preventing growth of bubbles and coalescence, which in turn produces a uniform and stable bed (Jin *et al.*, 2004).

Particle size

The effect of particle size on the performance of the ADMFB is shown by the performance curves of the bed depicted in Figure 4.1-6. The curves show the cumulative ash yield for the three different size ranges of +4.75mm-5.6mm, +2.8mm-4.0mm and +1.0mm-2.0mm,

operating the bed with (VA) or without (VD) the vibration motor running. This was done in the absence of a dense medium to determine the behaviour of the pure coal particles in relation to one another. The solid line in *Figure 4.1-6* represents the general float-sink performance curve, which will serve as a benchmark for the best possible performance obtainable for the coal used in this study.

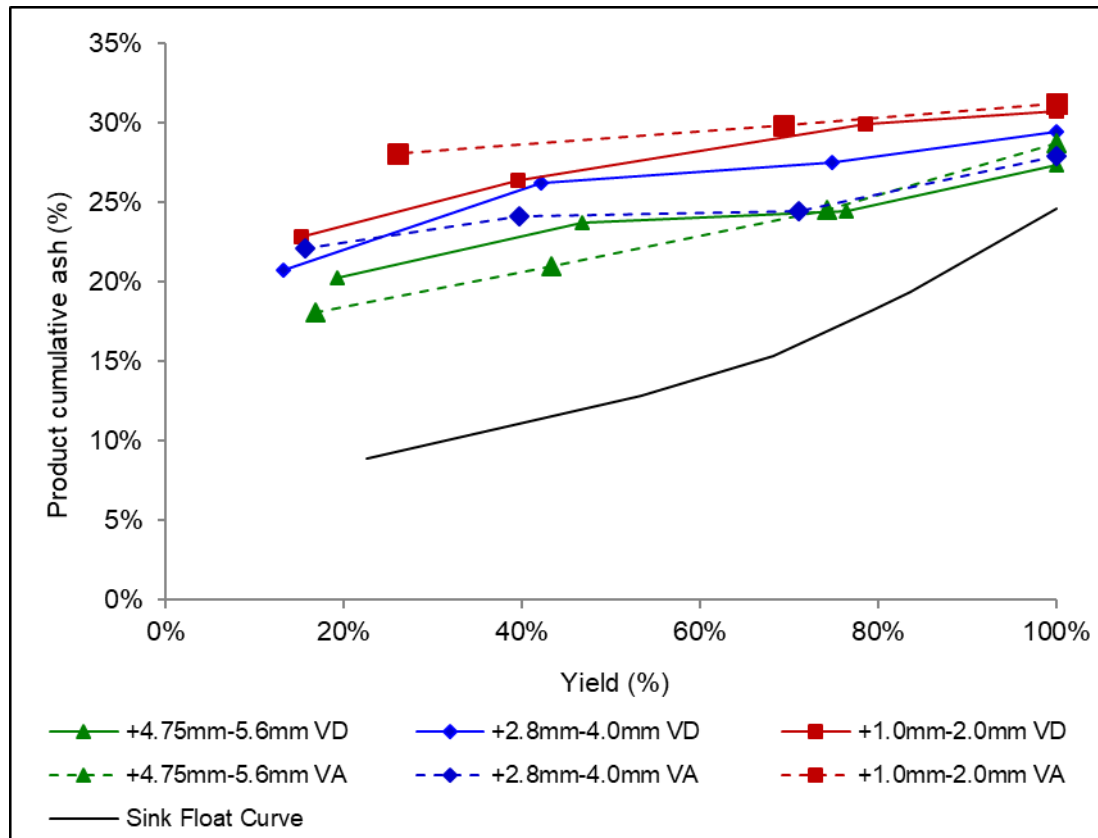


Figure 4.1-6: Comparison of bed performance curves for various PSD's (DM 0:1; VD and VA)

The results obtained for PSD-06 (+4.75mm-5.6mm), PSD-04 (+2.8mm-4.0mm) and PSD-02 (+1.0mm-2.0mm) followed the same general trend of the float-sink curve; however, it is evident that the separation performance of the bed becomes poor when the particle size range decreases. Although the product cumulative ash of PSD-06 and PSD-04 varied considerably from the float-sink performance curve, a cumulative ash differential of 24% to 18.1% and 24% to 22.1% was obtained at a yield of 55%. At the same yield, the trend of PSD-02 varied considerably from the float-sink curve and evidently, no ash yield reduction was observed. The separation performance mechanism of the ADMFB is density related, but the influence of particle size range is such that in some cases it competes with the extent of segregation based on density to the point that it overpowers the former. For larger sized particles, segregation occurs more vigilantly whereas the extent of segregation decreases for the smaller particles, typically those less than 2.8mm. This observation was confirmed by

Yang *et al.* (2015), who also found a low degree of segregation for -3mm sized particles in an ADMFB using Chinese coal.

Dense medium

The influence of addition of various feed mass ratios of dense medium to coal on the separation performance of the bed was investigated by comparing the coal ash yield values of the top and bottom bed layers to that of the feed coal after fluidization. The experimental runs without dense medium comprised of coal only and are referred to as DM 0:1. For a DM 1:1, 7.5kg magnetite and 7.5kg coal was mixed and fed to the bed, whereas for DM 2:1, 10kg magnetite and 5kg coal was mixed. The coal and magnetite sample were mixed thoroughly and fed to the bed. The feed mixture was fluidized at $1.1 U_{mf}$ for a period of 10 minutes, where after the airflow was switched off and the bed allowed to settle prior to sampling the different layers. The results for performance curves of +1.0mm-2.0mm (PSD-02) and +4.0mm-4.75mm (PSD-05), was compared to the theoretical float-sink performance curve (solid line) as shown in *Figure 4.1-7*.

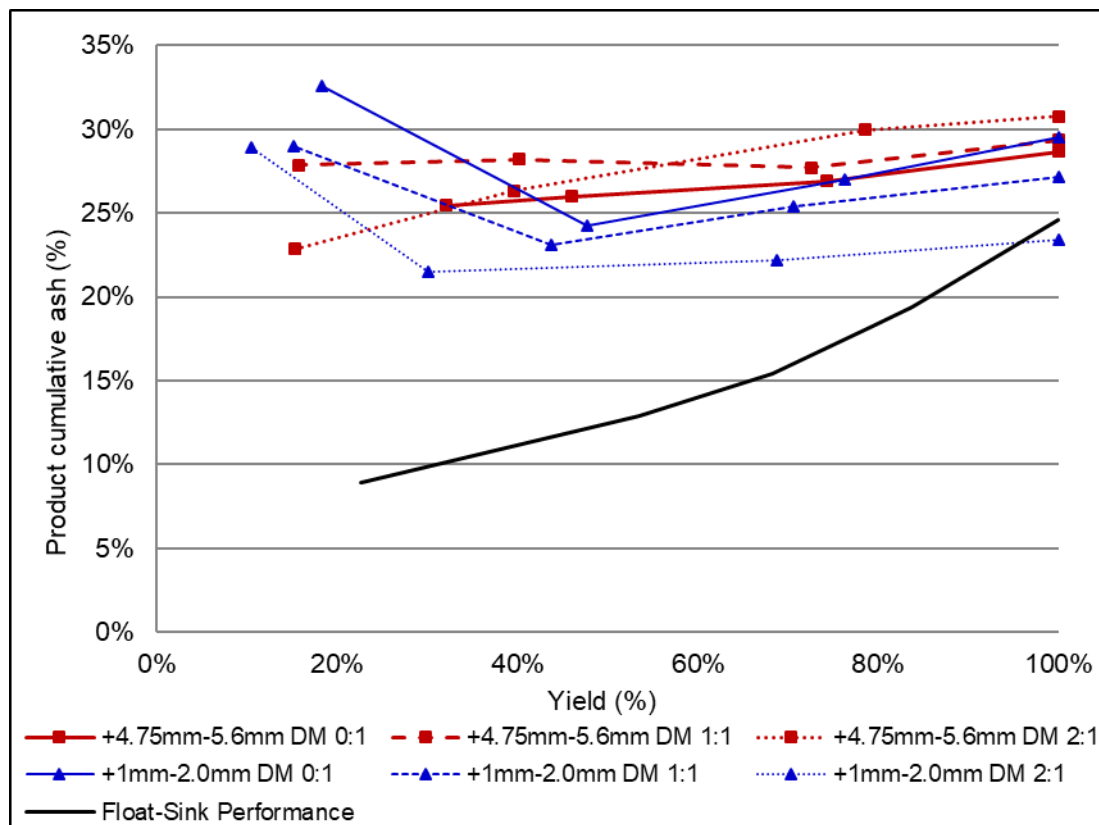


Figure 4.1-7: Comparison of bed performance for +1.0mm-2.0mm and +4.0mm-4.75mm without vibration

From *Figure 4.1-7* it is evident that the addition of dense medium yielded no significant improvement in the performance of the bed. The previously observed size influence was

again clear by comparing the extreme sizes for each of the setups. Theoretically, the small dense medium particles should behave as a lubricant that, by moving in-between the coal particles, it assists the coal in moving around more freely in the bed. In actual practice, the movement of dense medium particles in the ADMFB occurs due to a disturbance created by the air penetrating the bed (Mohanta *et al.*, 2015) and hence transporting the dense medium particles to form the pseudo fluid.

The curves displayed in *Figure 4.1-7* does not follow or fall within range of the float-sink performance curve, which is indicative of poor separation. By running the bed in batch mode whereby both the coal and dense medium is fed to the bed prior to fluidization, the dense media is not allowed to form a uniform suspension and therefore cannot direct coal to move through the bed based on its density. This results in the mixing of particles which hinders the segregation thereof.

Operating velocity

The operating air velocity of the ADMFB in relation to the minimum fluidization velocity was investigated by analysing the level of segregation of the coal particles subjected to various fluidization factors. These variable factors included the coal PSDs, coal to dense medium ratios, the addition and absence of vibration to the bed as well as the velocity of the fluidizing air. *Table 4.1-4* provides the ash yield results obtained for the feed coal, the top and bottom sampled layers as well as the ash difference between the feed coal and top layer and that of the top and bottom layer against the Fluidization factor. This factor is defined as the fraction of the minimum fluidization velocity (U_{mf}) at which the bed is operated. The results for PSD-03 (+2.0mm-2.8mm) are shown in *Table 4.1-4*.

Table 4.1-4: Influence fluidization factor on ADMFB performance

Ash Yield (%_wt) DM 0:1	Bed Layer	Fluidization Factor (U^*)				
		1.05	1.1	1.2	1.3	1.4
	<i>Feed</i>	28.87	27.55	27.22	27.47	27.88
	<i>Top</i>	23.97	22.89	22.19	24.86	25.38
	<i>Bottom</i>	32.38	34.55	35.97	35.73	32.20
	<i>Feed – Top</i>	4.9	4.66	5.03	2.61	2.5
	<i>Bottom - Top</i>	8.41	11.66	13.78	10.87	6.82

For all the PSD's investigated, it was established that the velocity of the fluidization air has a definite effect on the stratification of coal within the bed. The results in *Table 4.1-4* showed

that an increase in the fluidization factor prompted an increase in the ash yield values of the bottom layer and led to an increased ash differential across the bed (*Bottom-Top*). Increasing the fluidization factor beyond this ideal point resulted in vigorous fluidization/back mixing of particles within the bed, which affected the ash yields of the layers.

He *et al.* (2016(a)) marked that a point of maximal segregation can be identified at the moment when the performance of the ADMFB would produce the best segregation depending on the coal particle size, dense medium, and vibration as well as other fluidization factors. Maximal segregation efficiency was obtained, for all the PSDs investigated by He *et al.* (2016(a)), at a fluidization factor of 1.6. It was decided to investigate the maximal segregation for the finer coal fractions of +1.0mm-2.0mm (PSD-02) and +2.0mm-2.8mm (PSD-03). *Figure 4.1-8* contains the results from the experimental runs compared to the results reported by He *et al.* (2016(a)), including the maximal point of segregation.

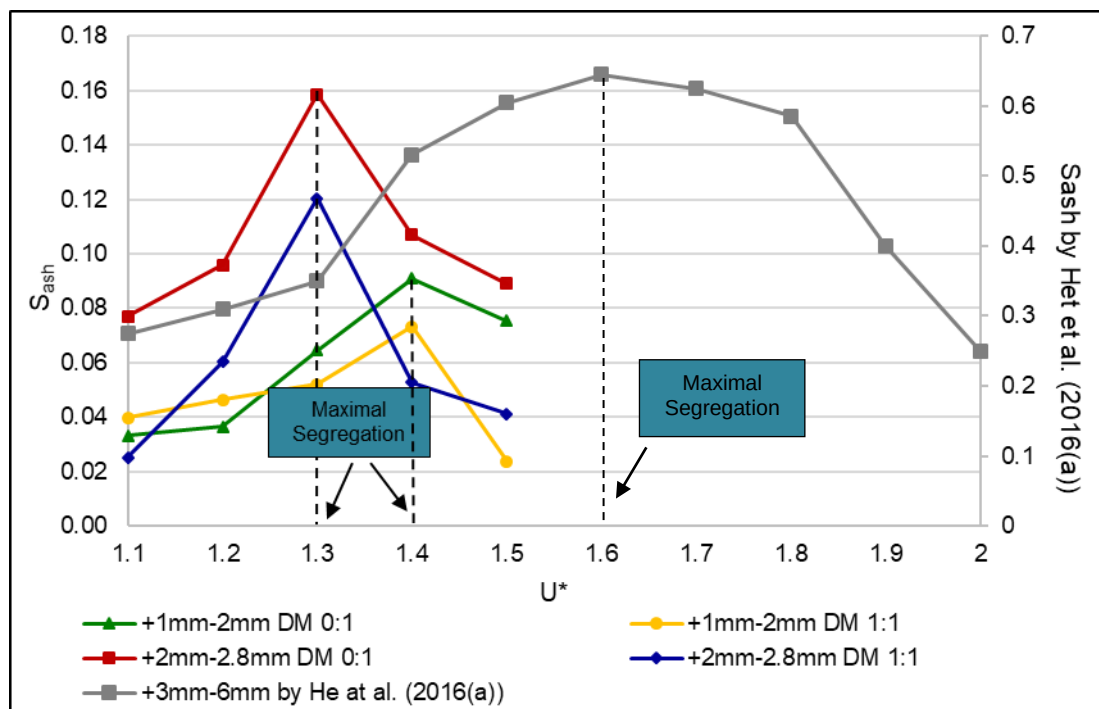


Figure 4.1-8: Degree of segregation for PSD-02 and PSD-03 compared to literature (He *et al.*, 2016(a)).

The maximal segregation point indicated in *Figure 8* represents the best ADMFB operating air velocity. The best operating velocity for the coal tested by He *et al.* (2016(a)) was found to be $1.6 U_{mf}$. He *et al.* (2016(a)) showed that segregation occurred, with maximal segregation efficiency (S_{ash}) increasing from 0.65 to 0.75 for +3mm-6mm and +20mm-25mm, respectively. A similar trend was observed for the results obtained by this study, with the maximum segregation efficiency increasing from 0.09 to 0.15 for +1.0mm-2.0mm (PSD-

02) and +2.0mm-2.8mm (PSD-03), respectively. Compared to the results of He *et al.* (2016(a)), the airflow velocity required for maximal segregation is significantly lower. This indicated that the segregation efficiency is reduced with a reduction in particle size.

At $1.1U_{mf}$ in *Figure 4.1-8*, more-dense and less-dense coals were slightly mixed, and therefore the density segregation of the coal particles was minimal. Initially, the air velocity was not sufficient to overcome the packed bed resistance caused by the coal and magnetite; however, the resistance was overcome when the air velocity was increased to $1.2U_{mf}$ and particle segregation started. When the airflow was increased, the segregation efficiency of the coal particles was improved as small bubble formation was enhanced, allowing additional space for coal particles to float or sink. Maximal segregation was obtained around at $1.3U_{mf}$ and $1.4U_{mf}$ for +2.0mm-2.8mm (PSD-03) and +1.0mm-2.0mm (PSD-02), respectively. However, when the airflow was increased beyond the point of maximal segregation, back mixing was initiated resulting in small bubbles joining to produce larger, undesirable bubbles. The increased velocity leads to bubbles bursting, resulting in back mixing of particles. This back mixing causes the finer particles to percolate to the bottom of the bed creating mixed bed density, hindering the stratification along the bed height. Upon further increasing the air velocity, back mixing in the bed was intensified.

Vibration frequency and amplitude

The separation performance of the ADMFB was expected to improve when vibration was added to the system because the particles are loosened, creating a more stable fluidized bed (Sahu *et al.*, 2009). He *et al.* (2015) states that the effect of vibration on an ADMFB depends on the geometry of the bed. In order to select the appropriate frequency and amplitude values for operating of the bed, a study conducted by Yang *et al.* (2015) was consulted. The study states that frequencies ranging from 18 to 33Hz and amplitudes ranging from 2.6 to 3.8mm is ideal for the fluidization of +1mm-6mm coal particles. The upper and lower bounds of the vibration frequency and amplitude was identified by testing the combined influence of the amplitude and frequency.

A 0.18 kW oscillatory vibration motor is affixed to the bottom of the bed. The motor is a 2-pole BM 200/3 model OMB vibration motor that regulates the vibration induced into the system by adjusting the frequency. The maximum capability that can be achieved by the motor is 3000 revolutions per minute (RPM) at 50 Hz and an amplitude 10mm. The amplitudes investigated were 30%, 40% and 50% of the motor's maximum amplitude. The lower bound frequency was identified as 20 Hz was, while the upper bound was found to be at a frequency of 22 Hz.

Figure 4.1-9 displays the results of two experimental runs in which the frequency is varied and the ash yields are compared.

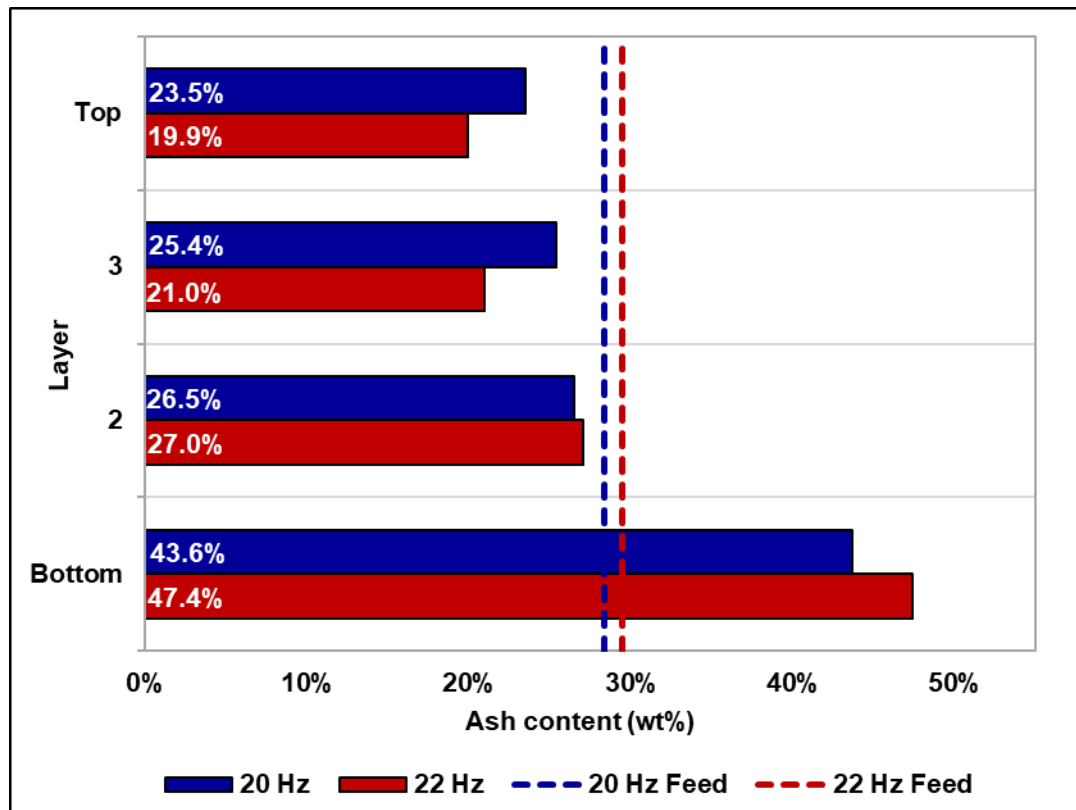


Figure 4.1-9: Ash analyses for +4.75-5.6mm particles and 30% amplitude and 20 Hz

The difference in ash values between the feed and top layer of the 22 Hz run, shown in Figure 4.1-9, is about 10%_{wt}, whereas the top and bottom layers yielded a differential of 27.5%_{wt}. The difference in ash values between the top and bottom layers of the bed improved when the frequency of the bed is increased, which is verified by the findings of Sahu *et al.* (2009). The two bottom layers for both runs shown in Figure 4.1-9 indicates that the stratification and amount of dense material reporting to these layers are dependent on the feed coal values. A study conducted by Diedericks *et al.* (2020) also showed that the separation performance when adding vibration to the ADMFB was dependent on the particle stratification in the two bottom layers of the bed. No major improvement in the results of the various bed layers was seen when vibration was added to the system, with the exception of an increase in the ash values in the bottom bed layer. However, this may be due to the respective feed ash values of the coal (Diedericks *et al.*, 2020).

To assess the influence of vibration amplitude added to the system, the vibration frequency was maintained at 21.97 Hz and the amplitude was varied between 30%, 40% and 50% of the motor's maximum capabilities (motor weights). It should be noted that the percentage

amplitude refers to the capability of the vibratory motor. The influence of 30% and 50% amplitude is shown by the ash yields obtained for +4.75mm-5.6mm coal in *Figure 4.1-10*.

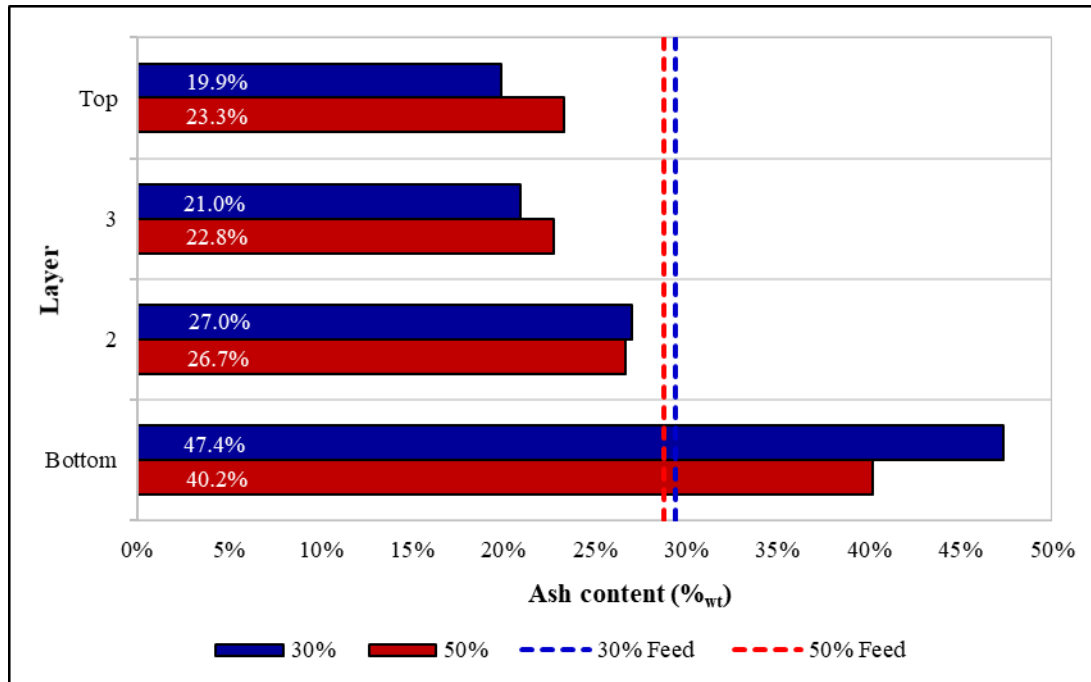


Figure 4.1-10: Ash analyses for +4.75-5.6mm particles at 21.97 Hz

The difference is apparent in the differentials of the beds shown in *Figure 4.1-10*; where the ash yield at 30% amplitude yields a 27.5%_{wt} differential and the 50% amplitude yields a 16.9%_{wt} differential (from top to bottom layer). This trend is observed for all the experimental runs tested and indicates that the separation performance of the bed decreases with an increase in the vibration amplitude. This is in accordance with the results found by He *et al.* (2015), who remarked that the formation of uniform micro-bubbles is hindered by elevated vibration amplitudes, which causes the separation performance to dwindle.

Furthermore, it was concluded from the experimental results that the minimum fluidization velocity (U_{mf}) was lowered when vibration was added to the system for all the particle size ranges. Studies conducted by McLaren *et al.* (2021) and He *et al.* (2015) made a similar conclusion that adding vibration to gas-solid fluidized beds enables fluidization of particles that are cohesive, reduce the minimum fluidization velocity (U_{mf}) of the particles as well as increase the fluidized bed stability.

Conclusion

The separation performance of solid phase bed material, at various particle size ranges (-5.6mm+0.5mm coal particles), in an air dense medium fluidized bed, were evaluated during this study. To obtain the best operating conditions, the experiments were conducted under similar ambient conditions in the presence of varying parameters. Six coal particle size ranges, three dense media to coal ratios and the addition of vibration was tested to identify the best operating conditions. Adequate results were obtained for larger particle size ranges down to and including +2.0mm-2.8mm coal particles, after which the separation performance decreased significantly. Density stratification was irregular and not obvious for coal particles below 2.0mm and maintaining a consistent fluidization state also proved to be challenging, especially when dense medium was added.

The coal particles separated vertically along the bed height because of differences in particle and bed density, while particle size proved to have a notable influence on the degree of separation. An air fluidization velocity of between 1.1 to $1.4U_{mf}$ was shown as the best performing velocity, which yielded the maximum ash differential between the top and bottom layers of the bed for all the particle size ranges tested. A summary of the lowest ash yield and highest mass yields obtained for +0.5mm-5.6mm is presented in *Table 4.1-5*.

Table 4.1-5: Cumulative ash yield and mass yield at a 50mm cut-point height (discarding only the bottom layer)

Particle size	Dense medium	Vibration	Feed Ash (% _{wt})	Cumulative Ash yield (% _{wt})	Cumulative Mass yield (% _{wt})
+4.75mm-5.6mm	0:1	Yes	34.8	18.1	74.3
+4.0mm-4.75mm	2:1	No	28.9	21.5	68.7
+2.8mm-4.0mm	1:1	No	31.8	20.7	74.8
+2.0mm-2.8mm	2:1	No	30.9	20.9	77.2
+1.0mm-2.0mm	0:1	No	29.9	25.5	74.3
+0.5mm-1mm	0:1	No	27.5	27.3	75.1

For +2.0mm-5.6mm coal particle sizes shown in *Table 4.1-5*, low cumulative ash yields were obtained at high mass yields, however the ash yields increased for coal smaller than 2mm. Vibration and dense medium have, in some cases, enhanced the separation efficiency of the ADMFB. The -2.0mm particles experienced stronger particle-particle interactions as well as

elevated levels of bubbling and back mixing than that of the +2.0mm particles, which explains the poor performance of the small particle sizes.

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CHAPTER 5

Chapter 5 contains the paper titled “Dry beneficiation of small South African coal using an air dense medium fluidization bed”. In this paper, air dense medium fluidized bed (ADMFB) is evaluated as a suitable dry beneficiation technique to upgrade **+0.5mm-13.2mm** South African coal. The separation efficiency of the ADMFB was assessed by determining the effect, if any, of particle size range, medium to coal feed ratios and the addition of vibration on the separation ability of the bed. The coal was upgraded to varying degrees, depending on the parameters of the applied methods.

Dry beneficiation of small South African coal using an air dense medium fluidization bed (Paper 3)

This section provides information regarding contributions made by the various authors towards completion of the paper. Bibliographic information regarding the published paper along with additional information of the congress proceedings in which the paper was published.

Contribution by authors

Misses E.S. Diedericks conducted a coal beneficiation research study with the guidance supervision of Prof. M. le Roux and Prof. Q.P. Campbell. The concept and scope of work for this research were largely formulated by Prof. M. le Roux and Prof. Q.P. Campbell. The air dense medium fluidized bed experimental setup was designed by Miss N Hughes and constructed by the North-West University technical support staff, with guidance from Miss N. Hughes. Misses E.S. Diedericks was responsible for conducting the experimental work, collecting data and processing of the collected data. The results from experimental runs were discussed and assessed during multiple consultations between Misses E.S. Diedericks, Prof. M. le Roux and Prof. Q.P. Campbell. After compilation review cycles, the paper was published in a peer-reviewed journal, International Journal of Coal Preparation and Utilization. The paper was written by Misses E.S. Diedericks and reviewed and edited by Prof. M. le Roux as well as Prof. Q.P. Campbell. The content of this paper was presented by Misses Diedericks at the XIX International Coal Preparation Congress (ICPC). The conference was held in New Delhi, India, from the 13th to the 15th of November 2019. Thereafter the work was published in the congress proceedings.

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5.1 Dry beneficiation of small South African coal using an air dense medium fluidization bed

ES Peters-Diedericks, M le Roux, QP Campbell & N Hughes*

School of Chemical and Minerals Engineering, North-West University, Potchefstroom, South Africa.

**marco.leroux@nwu.ac.za*

Abstract

This study investigated the air dense medium fluidized bed (ADMFB) as a suitable dry beneficiation technique to upgrade small (+0.5mm-13.2mm) South African coal. The separation efficiency of the bed was assessed by determining the influence, if any, of particle size range, dense medium (magnetite) to coal feed ratios and the addition of vibration to the bed. During batch operation, most of the high ash coal collected in the bottom layer of the bed, yielding coal with lower ash values towards the top layers. Product yields of 75% and 82% were found to contain ash of 13% and 22% respectively. The addition of +300 μ m magnetite ($d_{50} = 490\mu$ m) did not have a significant influence on the separation ability of +2.8mm coal particles, nor did added vibration to the system (apart from lowering the minimum fluidization velocity). The feed ash values obtained for the -2.8mm ranges were lower than for the +2.8mm ranges, leading to the conclusion that most of the valuable macerals have been liberated and hence carrying less mineral matter. Nevertheless, the -2.8mm beneficiated product still adhered to the local thermal coal market demands.

KEYWORDS: Small coal, dry coal beneficiation, magnetite, air dense medium fluidization

Introduction

Despite the fact that several countries are progressively investigating and developing clean energy alternatives, these are unlikely to replace coal as an energy resource in the near future because of its abundance and affordability (WEC, [20182022](#)). South Africa's national energy resource base is currently dominated by coal, with coal-fired power stations meeting more than 75% of the country's electricity demands (DOE, 2017). The South African Chamber of Mines (2018) recently noted that roughly 30% of the coal produced domestically is exported leaving coal as the largest contributor to the country's GDP. This highlights the importance of coal export in terms of foreign revenues. South Africa has recently been labelled a water-stressed country with a recorded average annual rainfall of 492mm, which is half of the global average (Rand Water, 2017), putting strain on washing coal, decreasing in

quality. This calls for efficient beneficiation techniques that do not require copious amounts of water or high operating and plant costs (De Korte, 2017). As such, coal processing industries worldwide are investigating dry processing instead of conventional wet processing as a possible solution. FGX separators, X-Ray sorters, air jigging, pneumatic oscillating tables and air dense medium fluidized beds (ADMFB) are some of the emerging dry beneficiation technologies that have been investigated in recent years and the performance was acceptable (Chen & Wei, 2003; Zhao *et al.*, 2015). ADMFB holds the advantages of reducing the risks associated with water pollution and costly dewatering processes by eliminating the use of water (Chen & Yang, 2003). The principle of air dense medium fluidization is that coal particles are stratified according to differences in density, with less dense coal particles moving to the top, and more dense coal particles (gangue) sinking (He *et al.*, 2016). For the past two decades, numerous small-scale successes for beneficiating coal, in a number of countries, using ADMFB's technology have been recorded. Table 5.1-1 shows that the separation efficiency (E_p) recorded for numerous countries. From the information in Table 5.1-1, it is apparent that the separation efficiency becomes worse as the particle size decreases, which indicates an increased degree in difficulty when beneficiating smaller coal sizes.

Table 5.1-1: Separation results of various coal size fractions

Country	PSD	E_p (g/cm³)	Medium	Reference
China	-50+6mm	0.05-0.07	Magnetite	Zhao <i>et al.</i> (2011)
India	-25+13mm	0.06	Magnetite	Mohanta <i>et al.</i> (2011)
India	-13+4.75mm	0.19	Magnetite	Mohanta <i>et al.</i> (2011)
Canada	-1+0.42mm	0.10	Magnetite	Choung <i>et al.</i> (2006)

Air dense medium fluidization has been implemented successfully in China (Mohanta *et al.*, 2011); however, South Africa's coal seams differ from other coal-producing countries and are notorious for being difficult to beneficiate. Therefore, thorough evaluation of the effectiveness of this coal processing technology on South African coal seams is necessary (De Korte, 2017).

Experimental

Materials used

The coal used for this study was sourced from the Witbank area in South Africa. The feed to the ADMFB was prepared by drying, crushing and screening the coal into various particle size fractions. Research revealed that adequate fluidization of coal particles is attained at

surface moisture levels below 5%_{wt} (Sahu *et al.*, 2009; Terblanche, 2013). The coal (as received) was air-dried by laying it out on a flat surface, at ambient air conditions, for 24 hours. Hereafter it was crushed and screened into the various particle size ranges including **PSD-11** (+11.2-13.2mm), **PSD-10** (+9.5-11.2mm), **PSD-09** (+8.0-9.5mm), **PSD-08** (+6.7-8.0mm), **PSD-07** (+5.6-6.7mm), **PSD-06** (+4.75-5.6mm), **PSD-05** (+4.0-4.75mm), **PSD-04** (+2.8-4.0mm), **PSD-03** (+2.0-2.8mm), **PSD-02** (+1.0-2.0mm) and **PSD-01** (+0.5-1.0mm). A representative sample of the material in each PSD was taken by the cone-and-quarter method. The proximate analysis and calorific values for each PSD fraction were determined and the results fell within the ranges tabulated below.

Table 5.1-2: Range of average initial proximate and calorific values for feed coal

Description	Range	Standards
Moisture content (% _{wt}) Air dried	0.6-3.4	ACT-TPM-010 based on SANS 5925: 2007
Volatiles (% _{wt}) Air dried	21.4-25.4	ACT-TPM-012 based on ISO 562: 2010
Ash yield (% _{wt}) Air dried	27.0-33.2	ACT-TPM-011 based on ISO 1171: 2010
Fixed carbon (% _{wt}) By difference	41.3-49.5	By difference
Gross calorific value (MJ/kg)	21.7-25.5	ACT-TPM-014 based on ISO 1928: 2009

The magnetite dense medium used for this study was prepared by air-drying it to <5%_{wt} and screening to +300µm (d₅₀ of 490µm). The true density of the magnetite was 4.8 g/cm³ and was determined by helium pycnometry (SANS 1014:1985).

Equipment used

An air dense medium fluidized bed (ADMFB) unit, shown in *Figure 5.1-1*, was used to conduct batchwise experimental runs. The unit consisted of three conjoined sections that are identified as (1) an *airflow system*, (2) *fluidized bed* and (3) *dust control* section. The (1) *airflow system* consists of an air blower, airflow sensor as well as an air-distribution compartment. Air from the blower is distributed by a distribution device at the bottom of the bed, which consists of a packed bed of ceramic balls sandwiched between two distribution plates and a 100µm wire mesh. Eight identical rectangular layers (0.3m x 0.3m x 0.05m) were constructed from clear blue PVC and stacked on top of each other to form the (2) *fluidized bed section* of 0.4m high. Each layer could be detached and removed from the fluidized bed section, individually, to reveal a cut-point height of the material. The fluidized bed section was extended with a 0.3m layer, also constructed of PVC, to allow enough space for the particles to fluidize. The (3) *dust control* section of the ADMFB unit is composed of a tapered trapezoidal extension (20° hopper angle), covered by a 100µm stainless steel wire mesh. This effectively limits ultra-fine particles from escaping the unit as

the pressure drop is increased and the particles fall back into the bed. A 0.18kW oscillating vibration motor was mounted onto the outside structure of the unit, which could induce vibration to the bed. The frequency and amplitude range provided by the vibration motor is 15Hz and 0.7-1.0mm, respectively.

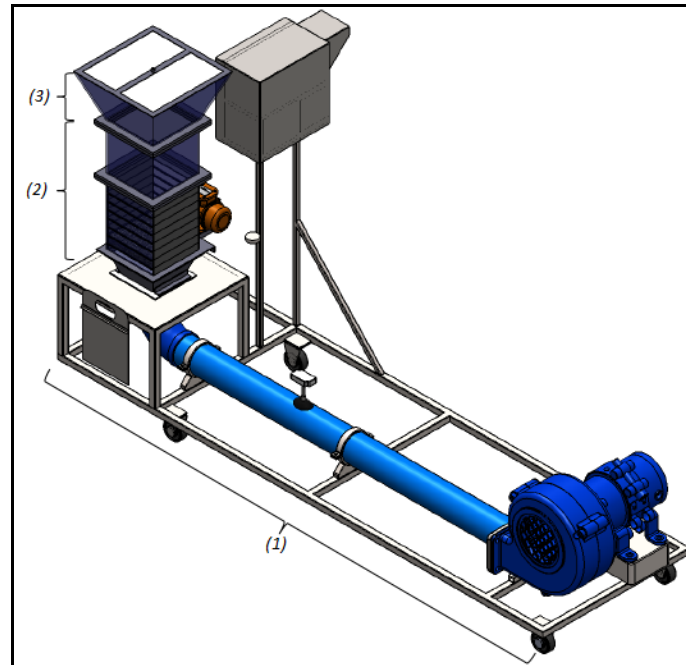


Figure 5.1-1: Three-dimensional illustration of the ADMFB with (1) airflow system, (2) fluidized bed system and (3) dust control section

The (2) *fluidized bed* section was loaded with a predetermined mass ratio of magnetite to coal. The load was fluidized at an operating velocity (U^*) of 1.05-1.30 times the minimum fluidization velocity (U_{mf}). After 10 minutes of fluidization, the ADMFB was switched off and the bed sections were disassembled from the top to the bottom. The material in each layer was retained in the individual frames by inserting a cutting plate (flat metal plate) between the layers and lifting the layer off the stack with the contents intact. The contents of each layer were coned and quartered until a representative sample for ash, calorific value and density analysis was obtained.

Results and discussion

To study the influence of PSD, dense medium to coal mass ratio (DM) as well as the effect of vibration (vibration activated, VA, or vibration deactivated, VD), the ash yield and calorific value (CV) for each layer along the height (bottom to top) of the bed was measured and recorded. PSD-06, without vibration (VD) and dense medium (DM 0:1) was selected for

demonstrative purposes. The ash yield (%_{wt}) of the four bed layers, along the bed height, together with the feed ash value is shown in the figure below.

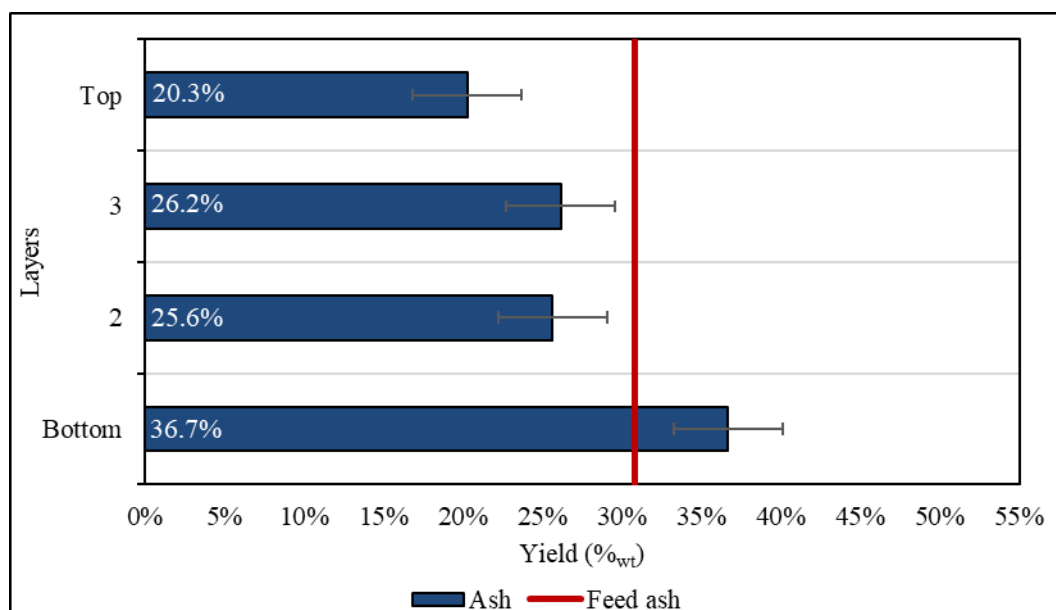


Figure 5.1-2: Ash values of control run according to bed layers (PSD-06; DM 0:1; VD)

The feed coal (the red line in *Figure 5.1-2*) had an ash yield and CV value of 30.8%_{wt} and 22.51MJ/kg, respectively, producing a top layer, which measured a 20.3%_{wt} ash value after fluidization and CV of 25.22MJ/kg. The ash yield recorded in the bottom layer, after fluidization, increased by about 16.5%_{wt}, which is indicative of de-stoning. The ash yields in the second and third layer of the bed (middlings portion) differs marginally, indicating that individual near-dense coal particles are constantly moving between the layers during operation and only settle once the airflow is switched off.

Effect of particle size range

The effect of particle size range on the minimum (U_{mf}) and operating (U) fluidization velocities of the bed is briefly discussed in this section. Several PSD's (DM 0:1), without vibration, along with the respective bed masses, U_{mf} and U values are summarized in *Table 5.1-3*.

Table 5.1-3: Minimum (U_{mf}) and operating (U) fluidizing conditions for various PSD's

PSD (mm)	Bed mass (kg)	U_{mf} (m/s)	U (m/s)
PSD-11 (+11.2-13.2)	15.06	15.1	16.1
PSD-09 (+8.0-9.5)	15.35	14.7	14.8
PSD-07 (+5.6-6.7)	15.20	9.2	10.4
PSD-05 (+4.0-4.75)	11.78	7.1	7.9
PSD-03 (+2.0-2.8)	12.67	5.7	6.3

PSD-01 (+0.5-1.0)	12.43	0.9	1.0
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It is apparent from the values in *Table 5.1-3* that an increase in U_{mf} is prompted by an increase in PSD. The reason for this is that a greater upward force is required due to the increased surface area of the of the larger particles, and in order to ensure the particles are fully suspended; a larger minimum fluidization velocity is therefore required. For pure coal beds, an operating velocity in close range with U_{mf} delivered a stable suspended bed with minimal turbulence and back mixing. The results obtained for PSD-11, PSD-09 and PSD-07 feed to the bed, operating it without dense medium (DM 0:1) or addition of vibration (VD), are summarized in *Table 5.1-4*.

Table 5.1-4: Ash, CV and density value comparison of PSD-11, PSD-09 and PSD-07

PSD (mm)	Layer	Ash (% _{wt})	CV (MJ/kg)	Density (g/cm ³)
PSD-11 (+11.2-13.2)	<i>Feed</i>	28.3	21.8	1.61
	Top	21.6	25.0	1.56
	Bottom	36.2	19.3	1.72
PSD-09 (+8.0-9.5)	<i>Feed</i>	30.6	21.3	1.66
	Top	24.0	24.0	1.54
	Bottom	41.6	17.1	1.84
PSD-07 (+5.6-6.7)	<i>Feed</i>	33.4	20.4	1.73
	Top	23.1	24.3	1.59
	Bottom	43.2	16.5	1.89

Comparable results were obtained for the bottom and top layers for PSD-11, PSD-09 and PSD-07 in *Table 5.1-4*. The degree of separation for PSD-11 to PSD-07 from top to bottom remained similar, however it seems the bottom layers were somewhat dependent on the feed coal characteristics. The ash values in the bottom layer increases in accordance with the feed ash values, which may be indicative of an increased fraction of dense particles present in the feed relocating to the bottom layer.

The ash yield obtained for the cumulative ash percentage in the product is represented in the form of performance curves like the one shown in *Figure 5.1-3*. The results of PSD-11, PSD-04 and PSD-01 are collectively compared to the float-sink test results (solid line in *Figure 3*) of PSD-11. The float-sink results for PSD-11 were selected for comparisons, since the float-sink curves of the remaining particle size ranges generated similar results (standard deviation of 0.14). Moreover, the yield versus the cumulative ash percentage is shown by this curve and it sets the basic standard for the best performance obtainable for the given coal samples.

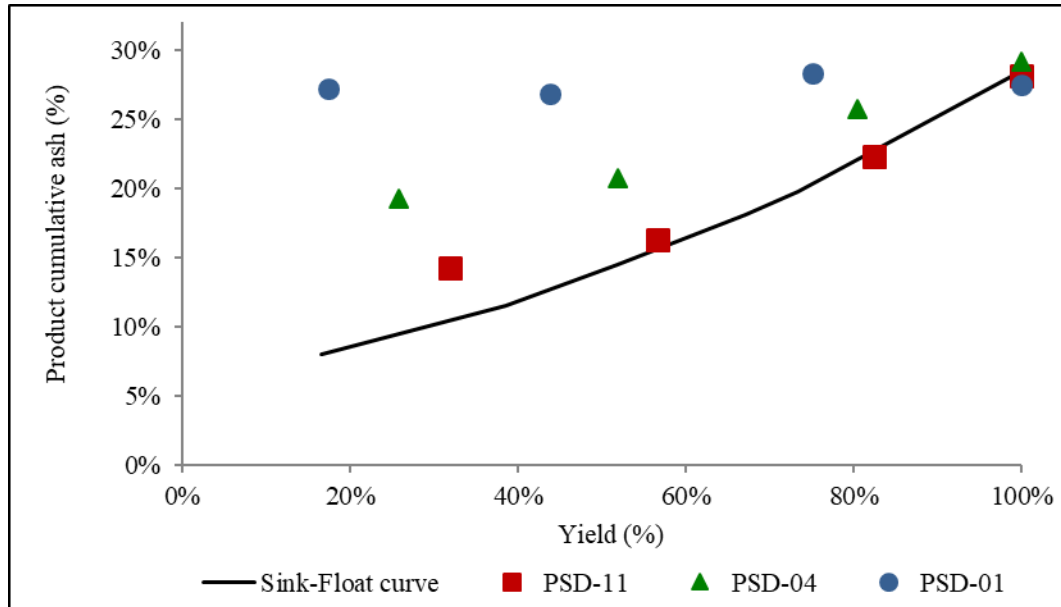


Figure 5.1-3: Comparison of bed performance curves for various PSD's (DM 0:1; VD)

The curves in Figure 5.1-3 highlight the ease of separation for the larger particle size ranges as compared to the smaller ones. Both the PSD-11 and PSD-04 produced acceptable performance to the float-sink curve at high yields. However, a limit in ash close to 15% and 20% was reached for PSD-11 and PSD-04, at coal yields of 30% and 25%, respectively. In contrast, at the same yield, the ash value of PSD-01 was found to be close to that of the feed coal, showing little to no beneficiation in the bed for these sized particles. This indicates that the stratification effect of the pseudo fluid bed is stronger for the larger particle sizes. Yang *et al.* (2015) found that a decrease in PSD yielded substandard results, which is confirmed by the performance of PSD-01 observed in Figure 5.1-3.

Effect of dense medium

The effect of dense medium on the separation performance of the ADMFB was investigated by varying the coal to dense medium ratio in the feed and comparing the results in the top and bottom layers of the bed after fluidization. For the experimental runs employing dense medium, the ratios were determined by weighing and mixing the dense medium and coal masses before placing it in the bed. About 7.5kg magnetite and 7.5kg coal was mixed for DM 1:1 experiment while 10kg magnetite and 5kg coal was mixed for DM 2:1 experiments. In order to ensure diverse, uniform distribution of coal and magnetite particles across the bed, it is essential to mix the magnetite and coal prior to loading. Table 5.1-5 compares the ash yields of the top and the bottom layers of the experimental runs for PSD-11 with DM 0:1, 1:1 and 2:1.

Table 5.1-5: Comparison of dense medium ratio of PSD-11 coal (VD)

Ash Yield (% _{wrt})			
Layer	DM 0:1	DM 1:1	DM 2:1
Feed	28.3	22.9	23.6
Top	21.2	20.3	21.2
Bottom	53.8	33.7	34.8

Minute differences are visible in the ash values of the top layers of the different beds, shown in *Table 5.1-5*, which indicates that the results obtained are close to best performance for each of the three beds. On the other hand, a difference in the ash values of the bottom layers of the dense medium beds in comparison to the pure coal bed was noticed. Both the dense medium beds behaved similarly to yield ash values in the proximity of 34%, while the ash values of the pure coal bed were 20 percentage points higher, close to 54%. A larger feed ash for the pure coal bed did contribute to this discrepancy, however it was evident that better destoning occurred within the pure coal bed. This was observed for all the dense medium beds at each PSD interval, with and without the addition of vibration to the bed.

In addition, it was also established from this study that the minimum fluidization velocity (U_{mf}) of a bed is influenced by the dense medium to coal feed ratios. In principle, a fluidized bed is fed with dense medium to create a fluid-like suspension of magnetite particles in which coal particles sink or float according to density. An experimental run using a pure magnetite bed (DM 1:0) of 15kg was tested and the minimum fluidization velocity (U_{mf}) and bed density thereof was found to be 8.0 m/s and 1.31 g/m³, respectively. This was done to compare the minimum fluidization of the DM1:1 and DM 2:1 bed to the pure magnetite (DM 1:0) bed. The minimum fluidization velocity (U_{mf}) values of PSD-11, PSD-07 and PSD-03 coal (total material mass per experimental run was recorded as ± 15 kg) as well as the operating velocities (U), are shown in *Table 5.1-6*, both without the addition of vibration.

Table 5.1-6: U_{mf} (m/s) and U (m/s) of PSD-11 and PSD-07 coal

DM ratio	Bed height (cm)	U_{mf} (m/s)	U (m/s)
PSD-11			
0:1	20	15.1	16.1
1:1	15	8.6	9.8
2:1	15	9.2	10.5
PSD-07			
0:1	20	9.2	10.4
1:1	15	7.2	8.3
2:1	15	5.2	6.9
PSD-03			

0:1	20	5.7	6.3
1:1	15	4.2	5.0
2:1	15	2.6	2.9

When dense medium is introduced to the system, the bed height as well as the minimum fluidization velocity is effectively reduced for all the PSD's. Once dense medium is added to the bed, a very fine particle with high density is added to the system. These high-density, fine particles collect in the pores between the large particles (blocking them), essentially exposing a solid surface to the air and thus is lifted with more ease from its resting position. It is apparent that the minimum fluidization velocity (U_{mf}) of the PSD's in *Table 5.1-6* falls within a narrow range from the 8.0m/s of the magnetite bed, with the exception of PSD-11 (DM 0:1) and PSD-03 (DM 1:1), (DM 2:1). PSD-03 (2:1) has a minimum fluidization velocity (U_{mf}) of 2.6m/s, which is considerably low compared to the minimum fluidization velocity of the pure magnetite bed (DM 1:0). It was found that when dense medium is added to the - 2.8mm coal particle size ranges, the pseudo fluid effect of the bed dwindles. This is because the coal and magnetite particles form a tightly packed bed in the ADMFB, making it difficult for the air to penetrate the particles, causing particle segregation and bed stratification to suffer.

Effect of vibration

Vibration as a parameter was investigated by adding a vibration motor to the bed that was able to induce a horizontal vibration motion. The addition of vibration to the bed lowered the minimum fluidization velocity (U_{mf}) while improving fluidization stability due to sufficient particle movement and bubble formation (He *et al.*, 2015). The fluidization velocities of PSD-11, PSD-09 and PSD-07 in *Table 5.1-7* show the influence of adding vibration to the system.

Table 5.1-7: Comparison of minimum and operating fluidizing conditions for varying vibration (DM 0:1)

Vibration	U_{mf} (m/s)	U^*	U (m/s)
<i>PSD-11</i>			
No	15.1	1.07	16.1
Yes	14.7	1.04	15.3
<i>PSD-09</i>			
No	14.7	1.01	14.8
Yes	11.1	1.15	12.8
<i>PSD-07</i>			
No	9.2	1.13	10.4
Yes	8.0	1.15	9.2

A decrease in minimum fluidization velocity (U_{mf}) is evident in *Table 5.1-7* for PSD-11, PSD-09 and PSD-07, when introducing vibration to the system. The same pattern was observed for all the other particle sizes tested during this study.

Conclusion

Upgrading of the coal quality was achieved to varying degrees according to the set variables. During batch operation, most of the high ash coal collected in the bottom part of the bed, yielding sequentially lower ash yield values towards the top layer where ash ranging between 13.0% and 22.4% were found for corresponding yields of 82.2% and 74.8%. Negligible influences on the separation efficiency for all the +2.8mm fractions were observed when applying vibration to the bed or feeding dense medium, +300 μ m magnetite (d_{50} = 490 μ m) with the coal to the batch setup. The addition of the vibration to the bed did however lower the minimum fluidization velocity and aided in particle striation and bed stability. Very little upgrading was observed for all particle size fraction less than 2.8mm. In general, the feed ash to the bed for -2.8mm coal fractions were lower than for the +2.8mm, leading to the conclusion that the mineral matter in the -2.8mm have already been liberated during the crushing and sizing process (sample preparation). In addition, the separation of coal and magnetite at these size ranges were extremely difficult due to adherence of magnetite to the additional exposed coal surfaces. But even with these less than desirable results, it was still possible to produce a product that adhered to the local thermal coal market demands. It can be concluded from this project that operating the ADMFB without the addition of magnetite or vibration will yield a coal product that comply to thermal standards, and in some cases to metallurgical standards. This operation philosophy aids in downstream processing by eliminating magnetic separation and recycling of dense media.

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CHAPTER 6

Chapter 6 contains the additional information related to Paper 1, Paper 2 and 3. Chapter 6 serves as explanatory section to the experimental work conducted for this thesis.

6.1 Additional Information (Paper 1 to Paper 3)

6.1.1 Overview of study

As previously mentioned, this study is focussed on evaluation of ADMFB technology as a suitable technique for beneficiation of small South African coal. During this study, coal particle size distribution (PSD), dense medium to coal mass ratio (DM: C) and vibration activated and deactivated (VA and VD) were tested. An investigation matrix of the study is shown in Figure 6.1-1.

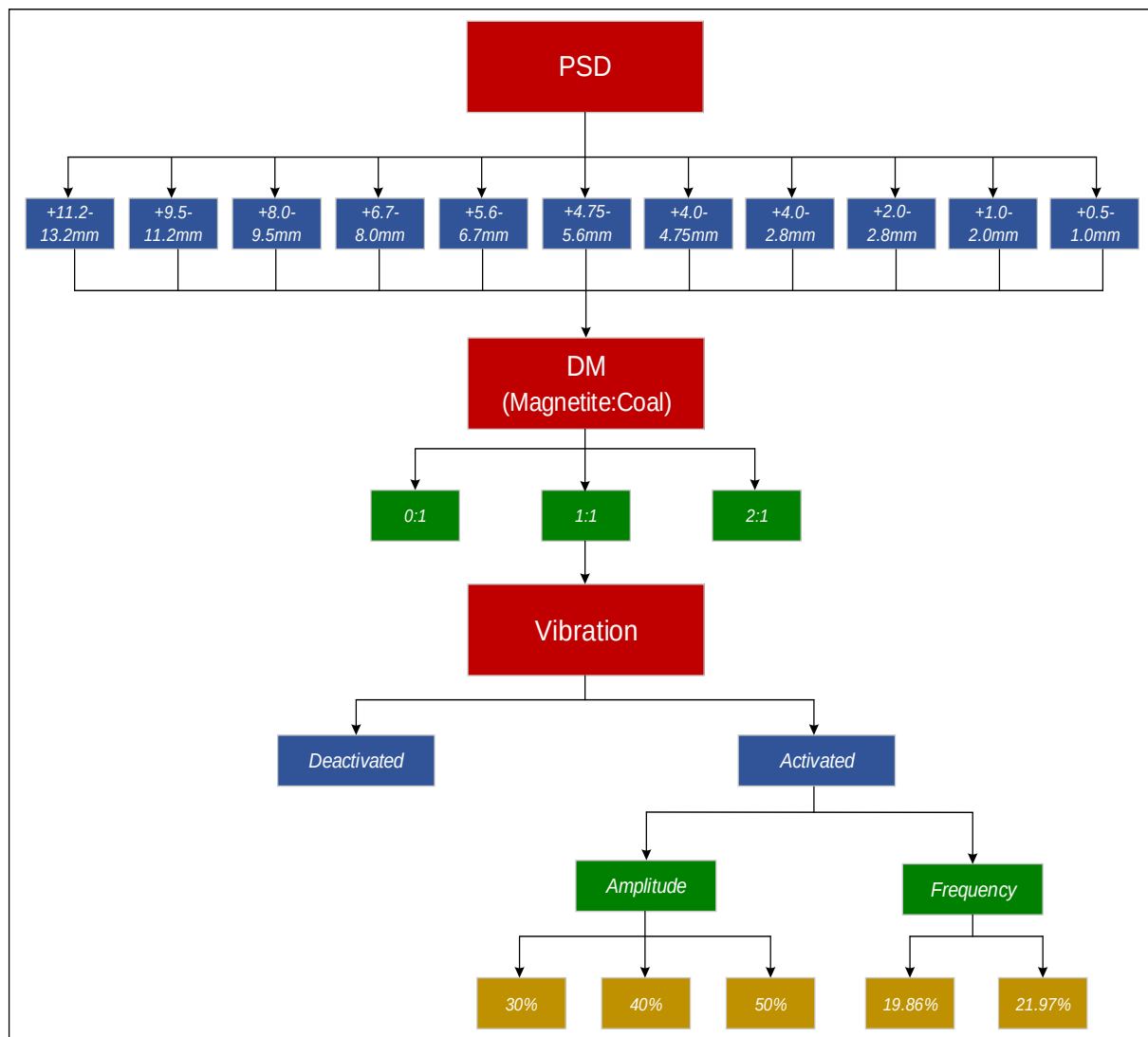


Figure 6.1-1: Investigation matrix

6.1.2 Experimental

6.1.2.1 Material properties

Coal from the number 4 seam in the Witbank area in South Africa was used for the experimental side study. The coal from this seam is bituminous, a thermal grade product, which is mined for the export market, with a small portion sold into the local market (Jeffrey, 2005).

Sample preparation consisted of drying, crushing and screening to provide the feed to the ADMFB. From research, it was established that coal fluidization is best achieved for particle surface moisture levels below 5%_{wt} (Sahu *et al.*, 2009; Terblanche, 2013). The coal (as received) was air-dried by laying it out on a flat surface, at ambient air conditions, for 24 hours. Hereafter it was crushed and screened into the particle size ranges shown in *Table 6.1-1*. The proximate analysis and calorific values for each PSD fraction were determined and it was found that the results were comparable across the different size ranges.

Table 6.1-1: Average initial proximate and calorific values for each coal particle size range

<i>Description</i>	<i>Moisture content (%wt)</i>	<i>Volatiles (%wt)</i>	<i>Ash yield (%wt)</i>	<i>Fixed carbon (%wt)</i>	<i>Gross calorific value (MJ/kg)</i>
+11.2mm-13.2mm	0.95	22.4	28.1	48.6	22.6
+9.5mm-11.2mm	1.04	22.1	28.4	48.5	22.5
+8.0mm-9.5mm	1.07	21.4	28	49.5	22.8
+6.7mm-8.0mm	0.61	22.9	30.8	45.7	21.7
+5.6mm-6.7mm	1.07	21.7	29.1	48.1	22.3
+4.75mm-5.6mm	1.67	21.6	32.6	44.1	24.3
+4.0mm-4.75mm	2.3	23.0	30.0	43.7	22.1
+2.8mm-4.0mm	2.46	23.9	30.8	42.9	25.5
+2.0mm-2.8mm	2.13	23.1	30.7	44.0	24.6
+1.0mm-2.0mm	1.93	23.6	29.2	45.3	25.2
+0.5mm-1.0mm	2.35	24.2	27.5	46.0	25.2
<i>Standards</i>	ACT-TPM-010 based on SANS 5925: 2007	ACT-TPM-012 based on ISO 562: 2010	ACT-TPM-011 based on ISO 1171: 2010	By difference	ACT-TPM-014 based on ISO 1928: 2009

A density profile of each particle size range was obtained by doing a float-sink analyses using zinc chloride solutions, for a range of 1.5 to 1.8g/cm³ (SANS 7936:2010). In addition

to the zinc chloride densimetric analyses, an organic solution was used to complete the densimetric analyses on the +0.5mm-1.0mm coal. White spirits and tetrachloroethylene were prepared at densities ranging from 1.30 to 1.60 g/cm³ and solutions of tetrachloroethylene and tetrabromoethane ranging from 1.70 and 1.80 g/cm³ were prepared. The coal samples were moved between these liquids to establish a density profile of that sample. Densimetric analysis were completed in two parts, +0.5mm-1.0mm coal particles were separated for densimetric analysis using organic liquids, while +1.0mm was analysed by using zinc chloride solutions. Organic liquids are substantially more expensive compared to zinc chloride and demineralized water. From an economic perspective, it was decided to use the expensive organic liquids for +0.5mm-1.0mm coal particles only, as this size range only requires a small amount. Additionally, the +0.5mm-1.0mm coal particles were small enough to wash through the mesh of the sieves used for the densimetric analysis with zinc chloride.

Magnetite powder was used as a dense medium in this study. It was prepared by drying it to <5%wt and screening to +300µm with a d₅₀ size of 490µm. The true density of the magnetite was found by helium pycnometry (SANS 1014:1985) to be 4.8g/cm³.

6.1.2.2 Equipment used

Experiments were conducted in a batch operated air dense medium fluidized bed (ADMFB) unit that consisted of three sections namely the airflow system, fluidized bed and dust control section, as shown in Figure 6.1-2. The airflow system consists of an air blower, airflow sensor (pitot tube) as well as an air-distribution compartment coupled to an air distribution plate. Eight rectangular frames (0.3m x 0.3m x 0.05m) manufactured from clear PVC were stacked on top of the distribution plate and clamped together to form a fluidizing unit having a height of 0.4m. Each layer could be detached and removed separately from the unit to aid in obtaining a sample cut of material at a certain height. The unit was extended at the top by a further 0.3m to allow enough space for the particles to fluidize. The dust control section of the ADMFB unit is composed of a tapered trapezoidal extension (20° hopper angle), covered by a 100µm stainless steel wired mesh. This limited the carry-over of ultra-fine particles by a reduction in air velocity at the top of the unit.

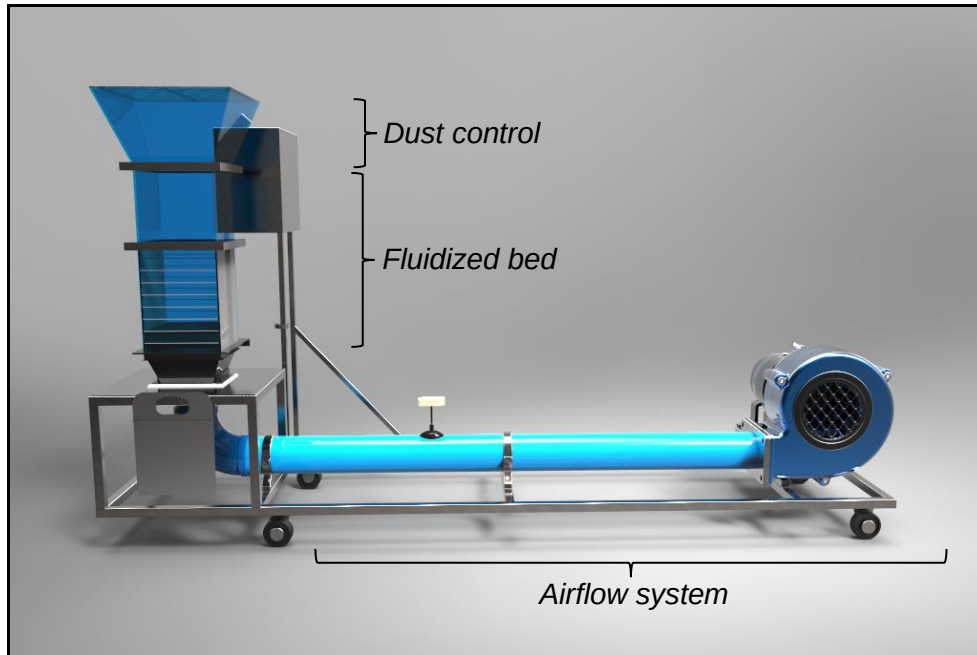


Figure 6.1-2: Three-dimensional illustration of the ADMFB

Predetermined mass ratios of coal and magnetite is loaded into the fluidized bed section. The mixture is fluidized with air at a predetermined minimum fluidization velocity (U_{mf}) for about 10 minutes. Afterwards, the layers were taken apart, the mixture harvested and analyzed for ash yield, calorific value, and density analysis.

Visual analysis was conducted on the feed coal batches to determine the shape and quality. The coal was found to disintegrate into a wide variety of shapes. Figure 6.1-3 is a visualization of these shapes.

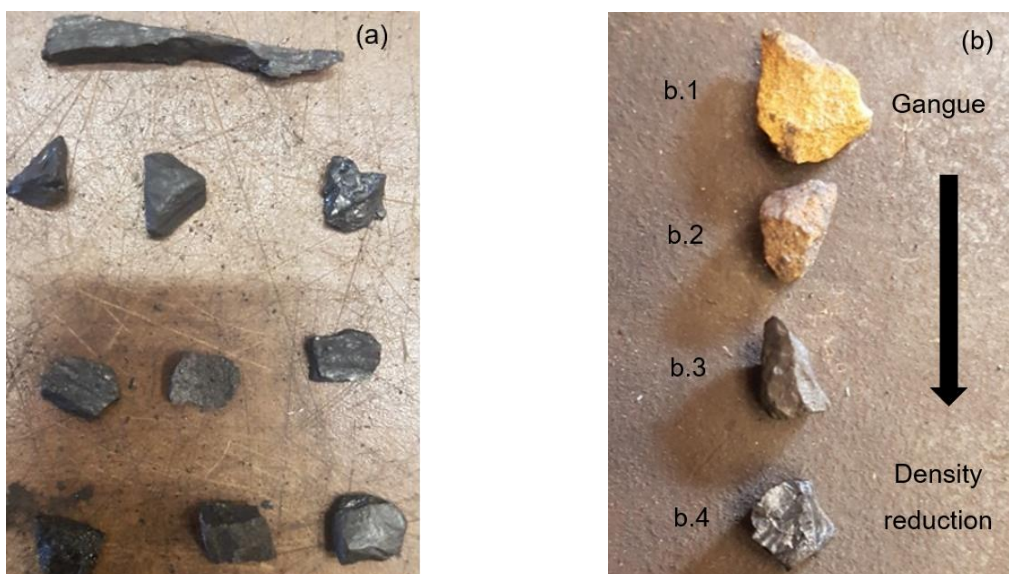


Figure 6.1-3: (a) Shapes of the crushed and sieved coal (b) Visual analysis of coalification of a sample

From Figure 6.1-3 (a) it is evident that various shapes of coal including cubical, longitudinal, rectangular and triangular shapes was present in the coal tested during this study. Figure 6.1-3 (b) illustrates liberation of coal and gangue from the same raw coal sample that was crushed during the ADMFB feed coal preparation process. The quality of the coal particles depicted in Figure 6.1-3 (b) contains less ash forming minerals moving from b.1 to b.4. Therefore, in bulk, beneficiation is required to separate these inconsistent particles.

Figure 6.1-4 shows the fluidized bed layers during the sampling process. A cutting plate, as shown in Figure 6.1-4 A, B and C, is used to dissect the bed layers for samples to be taken.



Figure 6.1-4: Sampling procedure

Before commencing with experimental runs, the minimum fluidization velocity (U_{mf}) has to be determined (repeated for all PSD's). Figure 6.1-5 depicts the pressure drop change when adjusting the airflow velocity. The minimum fluidization velocity (U_{mf}) is indicated by the green dotted line (denoted by MFV).

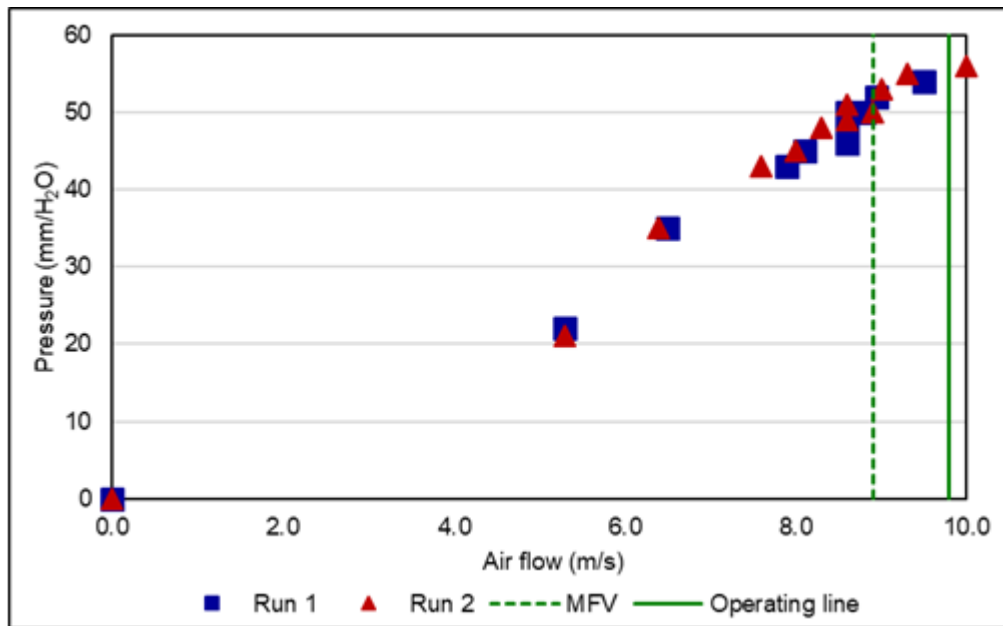


Figure 6.1-5: Pressure drop relative to airflow for -5.6+4.75mm particles (VD)

Hughes (2017) narrowed down the best operating fluidization velocity to between $1.1 \times U_{mf}$ - $1.3 \times U_{mf}$. Run 1 and Run 2 in Figure 6.1-4 shows U_{mf} to be around 8.9 m/s (as indicated by MFV in Figure 6.1-5.) and the operating velocity at about 9.8 m/s. Repeat experiments were conducted to establish the repeatability of the experimental procedure. The standard deviation of the U_{mf} was found to be 0.1m/s, which is a 1.14% error.

6.1.3 Results and discussion

Of the more than 70 experimental runs that constituted the main experimental matrix, shown in Figure 6.1-6, one experimental run, referred to as a control run, was identified to comprehensively discuss in this section.

Coal with a PSD of +11.2mm-13.2mm, without dense medium (DM 0:1) and vibration (VD) is chosen to present as a control run. Figure 6.1-6 illustrates the separation of coal particles according to the height of the ADMFB. The separation of the coal is depicted by the ash percentage that was analysed for each layer as well as for the feed. Figure 6.1-6 shows the ash content of +11.2mm-13.2mm coal particles with a feed ash content of 28.3%_{wt}, is reduced by 7.1% to layer 1 of the bed. Clear segregation is seen throughout the height of the bed, with a significant increase in ash content in the bottom bed layer compared to that in the feed. Moreover, marginal differences are reported for the ash values in the top bed layer and compared to that in the third layer of the bed.

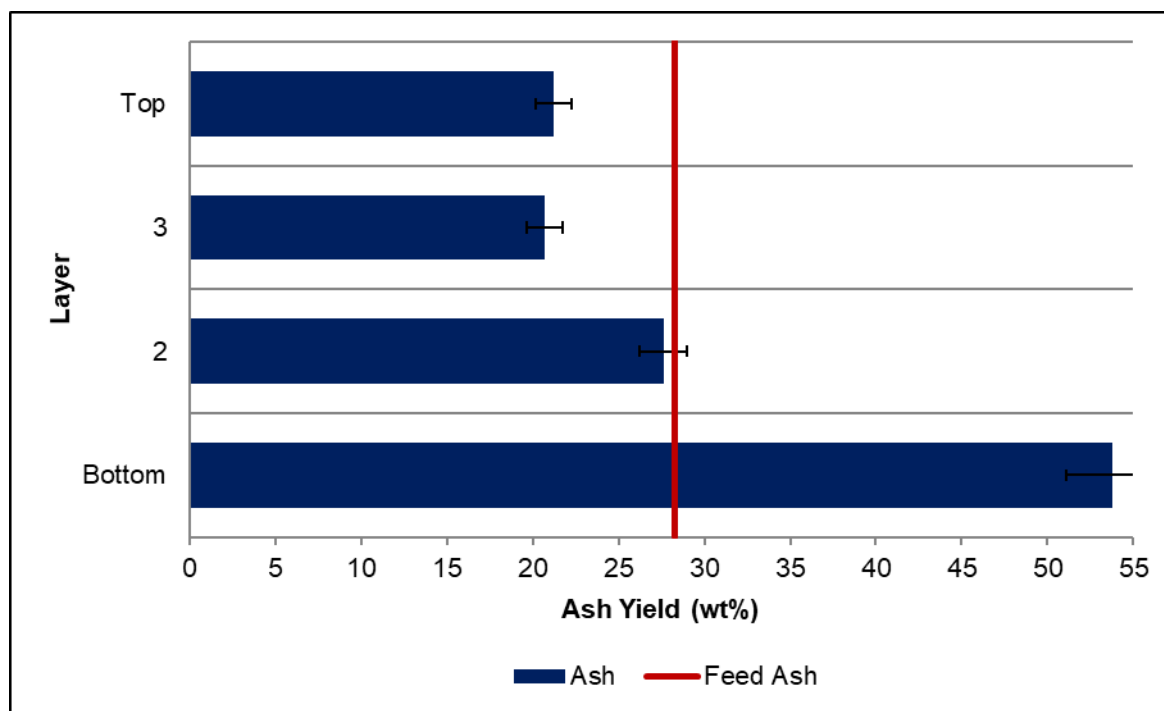


Figure 6.1-6: Particle segregation as per the ash yield for +11.2mm-13.2mm (DM 0:1 and VD)

A distinctive middling section with ash content of 27.6%_{wt} was produced. Exceptional destoning is captured in Figure 6.1-6 due to the exceptionally high percentage of ash forming minerals reporting to the bottom layer. This high destoning capability may be due to the particles size and the increased size of the inter-particulate voids. The increased void size provides additional area for the denser particles to move through and turn up in the bottom of the bed. This mechanism of particle transfer through the bed is similar to observations made by Yang *et al.* (2013), in that a decrease in particle size results in a decreased bed performance.

For each experimental run, total and ash mass balances were conducted on the system. The results of the total and ash mass balances are used to substantiate the ash results from the proximate analysis of each bed layer and to establish the amount of material lost during each fluidized bed cycle. Table 6.1-2 gives the total and ash mass balance for the feed as well as each respective bed layer of the control run (PSD +11.2-13.2mm; DM 0:1; VD).

Table 6.1-2: Total mass and ash balance for +11.2mm-13.2mm (DM :1 and VD)

Material	Feed (g)	Top layer (g)	Layer 3 (g)	Layer 2 (g)	Bottom Layer (g)	Bed total (g)
Coal	15060	4690	3630	3790	2950	14700
Ash	4262	994	751	1046	1393	4185
Coal Balance Error (%)			2.40	Ash Balance Error (%)		1.80

The ash balance error of 1.80% obtained for the bed is in agreement with the proximate results obtained for the feed and bed layers and falls within the recommended mass balance error of 5%. In addition, the mass distribution for the respective bed layers is provided. Although the ash yield produced for the bed layers are desired, the total mass of each distinct layer affects the overall yield of the bed. To be precise, the total mass of coal removed for every bed layer could deviate from the 25% of the total bed mass due to particle size and density.

Figure 6.1-7 illustrates the product and discard yield compared to bed height.

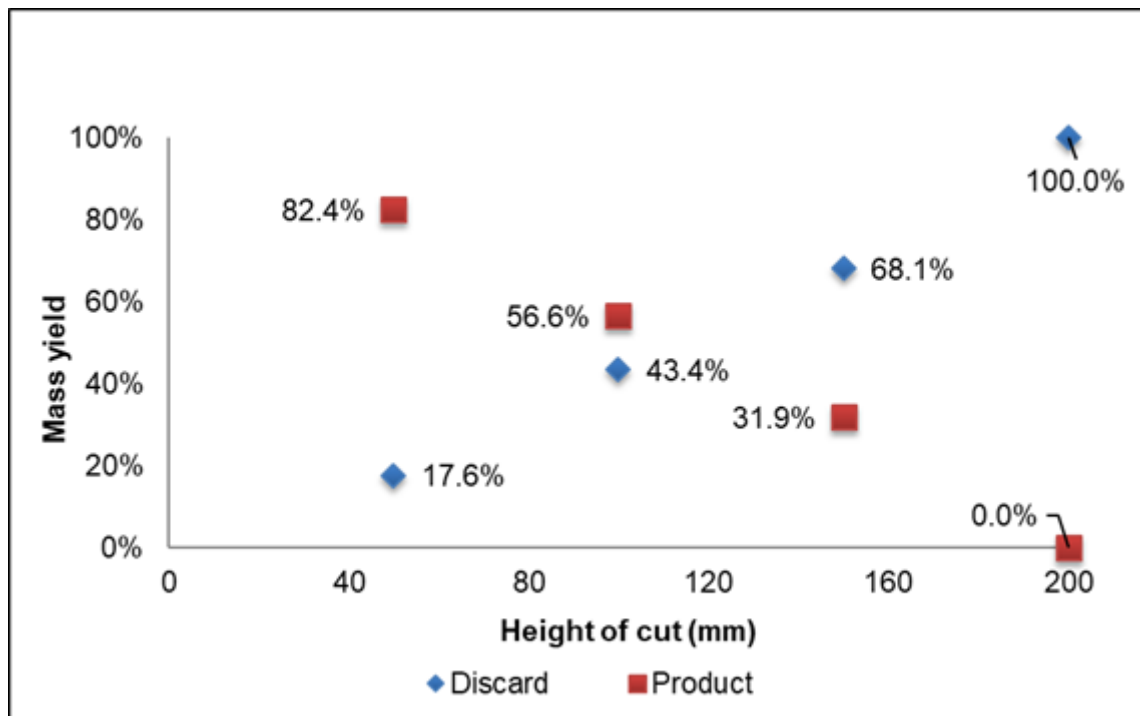


Figure 6.1-7: Mass yield compared to bed height for +11.2mm-13.2mm (DM 0:1 and VD)

From Figure 6.1-7, theoretical product and discard percentage (in terms of ash yield) can be seen on the vertical axis related to bed height on the horizontal axis (cut-point). If selecting a cut point of 50mm (collecting the top 3 layers as product and discarding the bottom layer), a mass yield of 82.4% with a corresponding ash yield of 22.3% is expected, which is not a substantial improvement from the 28.3%_{wt} feed ash content. If a product cut-point of 150mm (collecting the top layer as product and discarding the bottom 3 layers) is selected instead, a mass yield of 31.9% and ash yield of 14.2%_{wt} is obtainable. Therefore, the product quality can easily be altered by the cut-point selected for a specific bed. When the product requirement is driven by quality, one can obtain greatly improved coal ash yields by sacrificing mass yield.

When further evaluating the ash yield related to the coal discard, shown in Figure 6.1-8, it is evident that a reduction in discard ash is experienced as the theoretical cut point is increased in accordance with the bed height. This is an expected trend since the gangue migrate to the bottom layer of the bed based according to density. In the middling portion of the bed (layer 2 and 3), good quality coal and gangue is combined whereby a decrease in ash value directly results in an increase in cut point height.

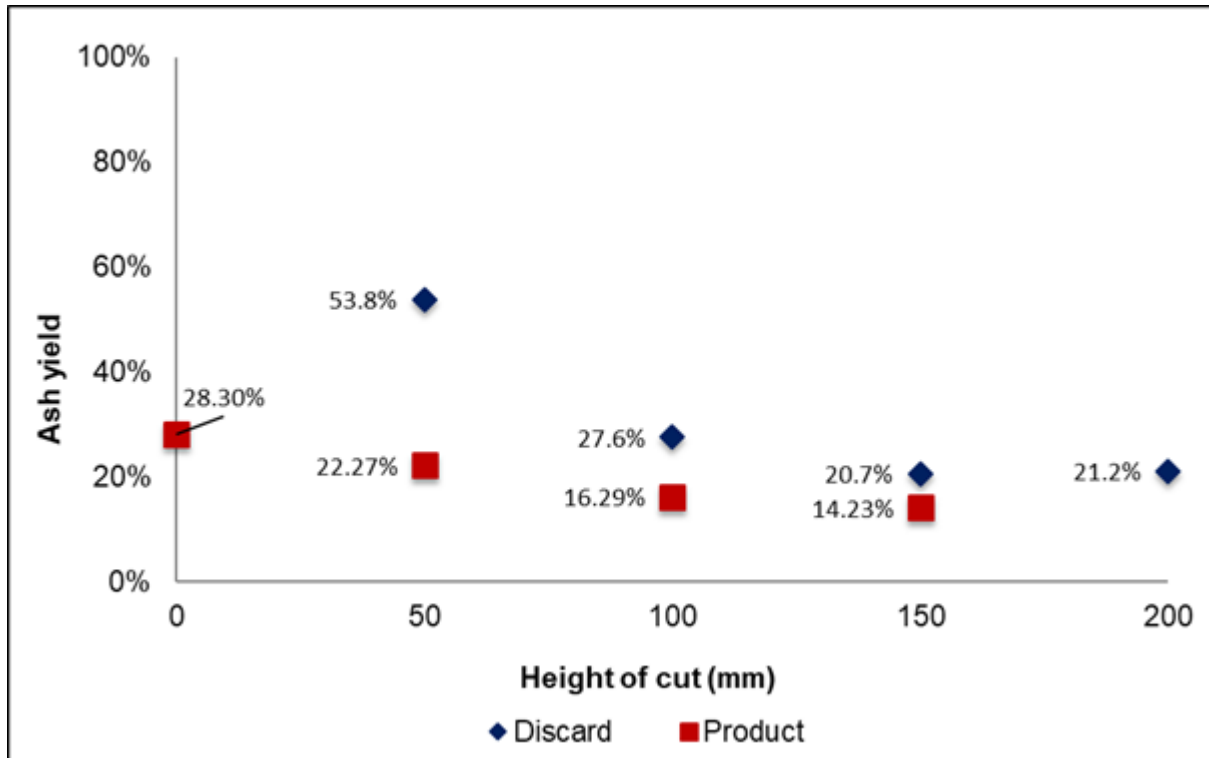


Figure 6.1-8: Ash yield compared to bed height for +11.2mm-13.2mm (DM 0:1 and VD)

Separation is at its sharpest at 50mm height as the biggest difference in product and discard ash values are seen in Figure 6.1-8. Therefore, in a case where both ash yield and mass yield are required to meet production specification, 50mm will be the best cut-point height.

In order to further illustrate the stratification achieved in the ADMFB, a performance curve was constructed. The performance curve, shown in Figure 6.1-9, denotes the cumulative ash yield percentage obtainable in the product.

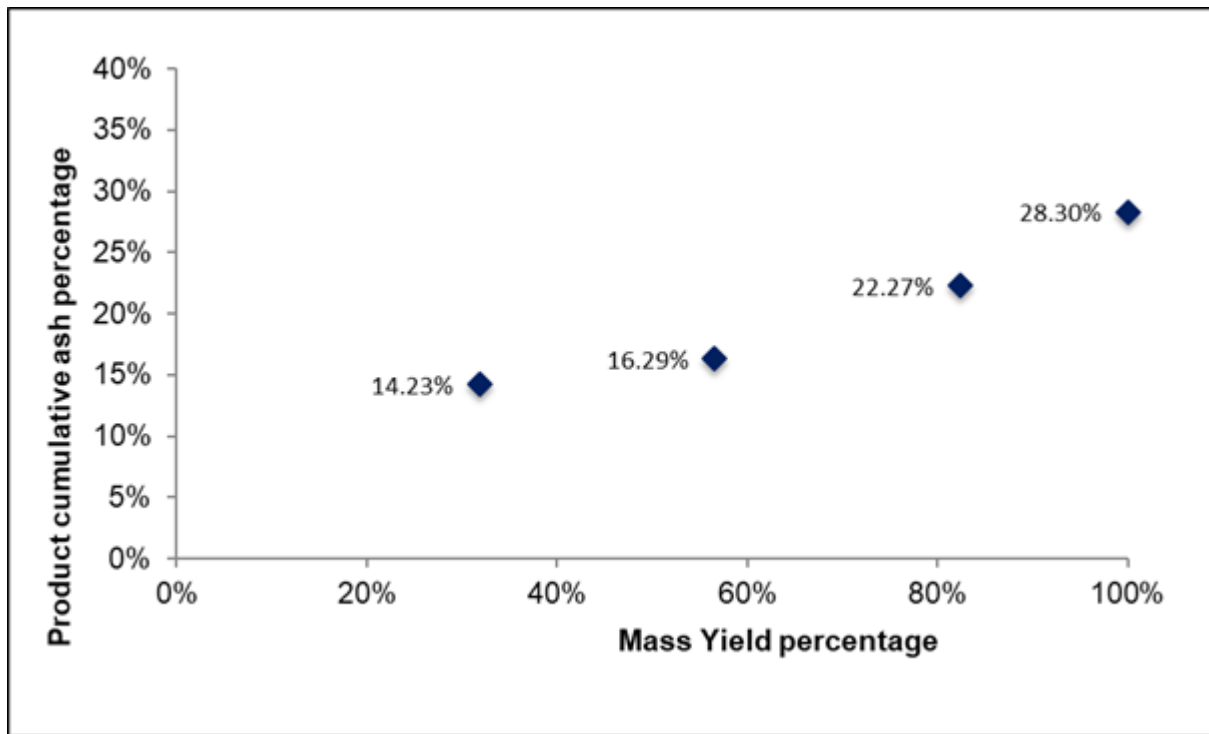


Figure 6.1-9: Cumulative ash yield compared to mass yield for +11.2mm-13.2mm (DM 0:1 and VD)

From Figure 6.1-9, recovering 100% of the mass (using the entire feed mass as theoretical product) will return unimproved coal with the same feed ash value (feed ash value for control run - 28.3%_{wt}) and as a result, an optimum cut point is required. The lowest and best cumulative ash product returned is produced at 31.9% yield, according to the control run presented in Figure 6.1-8 and Figure 6.1-9. By increasing the theoretical yield, an increase in cumulative ash yield is produced and as such, it is critical to identify and select a cut-point in the bed where coal of acceptable quality at adequate quantities is produced.

Theoretically, the cumulative floats curve, also referred to as the limiting performance curve or washability curve, provides the best performance that the coal can achieve with regards to limiting yield and ash value. Figure 6.1-10 provides the control run performance curve compared to the limiting performance curve of the coal. Evident in Figure 6.1-10, the ADMFB separation performance compared well to the washability curve considering the ash-yield curves follow the same trend.

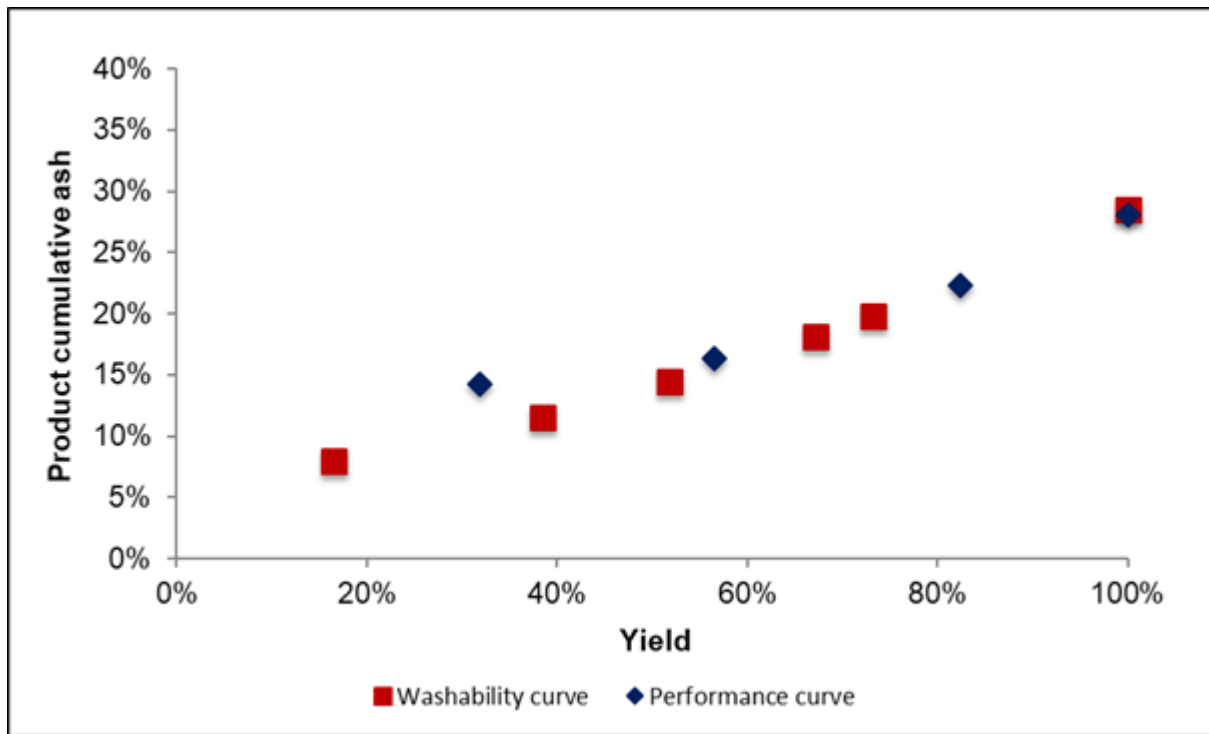


Figure 6.1-10: Comparison of bed performance and washability curve (limiting performance curve) for +11.2mm-13.2mm (DM 0:1 and VD)

6.1.3.1 Dense medium behaviour

One of the major challenges during operation of the ADMFB was the dense medium (magnetite) behaviour during fluidisation. Zhang *et al.* (2014) and He *et al.* (2016) established that the size of the dense medium is directly related to the fluidization stability that is achieved in the bed. It is well known from literature that -0.074mm magnetite causes severe instabilities in a fluidized bed (He *et al.*, 2016). Zhang *et al.* (2014) recommended using +0.074mm-0.30mm dense medium in the beneficiation process and as such the assumption was made that coarser magnetite would improve the stability of the fluidized bed. Therefore, the magnetite that was prepared to use as dense medium was screened to +300 μ m with a d_{50} of 490 μ m. Despite the employment of coarser magnetite, the coal to dense medium size ratio remains high (the magnetite particles are grouped according to a Gerhard type C). Some degree of difficulty in fluidization was experienced, as the magnetite adapted a solid-like behaviour instead of acting as a lubricant for particles to be displaced.

The weaker gas-solid forces compared to the inter-particle forces, may be the reason for the congealing behaviour of the dense medium. The resistance caused by the tightly packed dense medium particles hinders and effectively limits flow of air through the inter-particulate voids resulting in a slight increase in pressure beneath the magnetite bed. When the upward

air flow velocity reaches a point where the upward force overcomes the inter-particulate forces, the packed bed was disrupted, and the fluidization behaviour returned to normal. A similar response was experienced when coal was added to the fluidized bed system containing magnetite.

When considering the experimental runs for +11.2-13.2mm coal without vibration (VD), the dense medium proves to have no significant effect on the ash values in the top layer of the ADMFB. Figure 6.1-11 shows a comparison in ash values obtained by varying dense medium ratios (DM 0:1, DM 1:1 and 2:1).

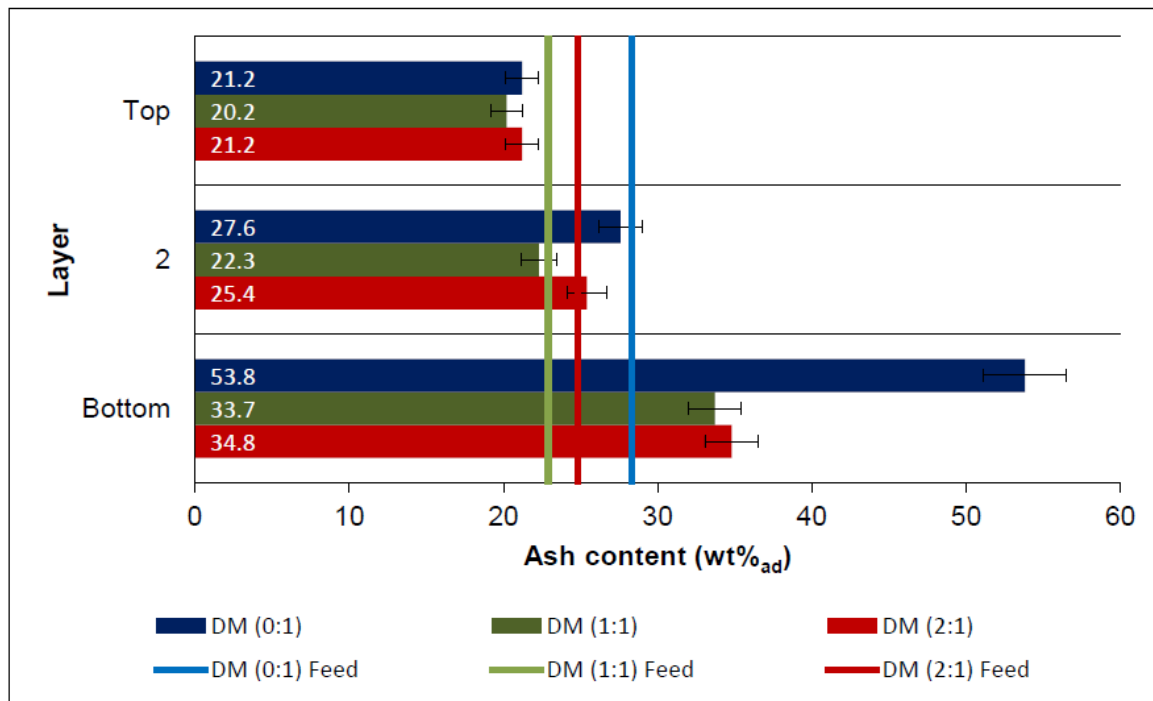


Figure 6.1-11: Comparison of bed performance for DM 0:1, DM 1:1 and DM 2:1 (+11.2mm-13.2mm and VD)

From Figure 6.1-11, the ash values in the top layer are constant for all DM to coal mass ratios tested. The same trend was found for all the coal PSDs tested and indicates that addition of dense medium will not yield better results compared to the ADMFB without addition of dense medium. However, a significant difference between the ash values is observed in the bottom layers of the bed in that it decreases with increased dense medium to coal ratio (biggest discrepancy seen between DM 0:1 and 1:1 and smallest between DM 1:1 and 2:1).

The performance curves for dense medium ratio comparison for +11.2mm-13.2mm coal without vibration are shown in Figure 6.1-12. It is evident from the performance curve that the run without dense medium yielded the best separation performance.

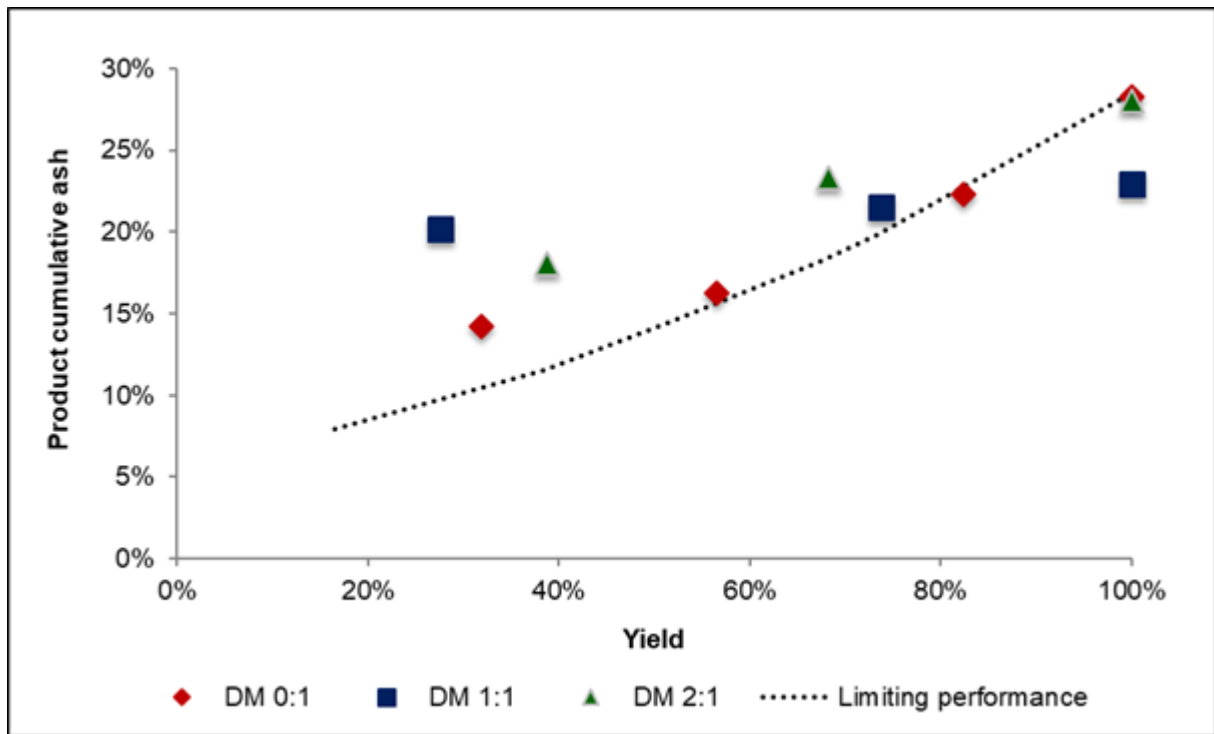


Figure 6.1-12: Comparison of bed performance for DM 0:1, DM 1:1 and DM 2:1 (+11.2mm-13.2mm and VD)

When considering all PSD ranges tested, variances from the trend in Figure 6.1-12 for the top bed layers were observed. When taking into account the overall segregation behaviour of all experimental runs with dense medium, it is evident that the separation achieved in the bed is inconsistent and poor. An ash comparison chart is provided in Figure 6.1-13 to exemplify this trend.

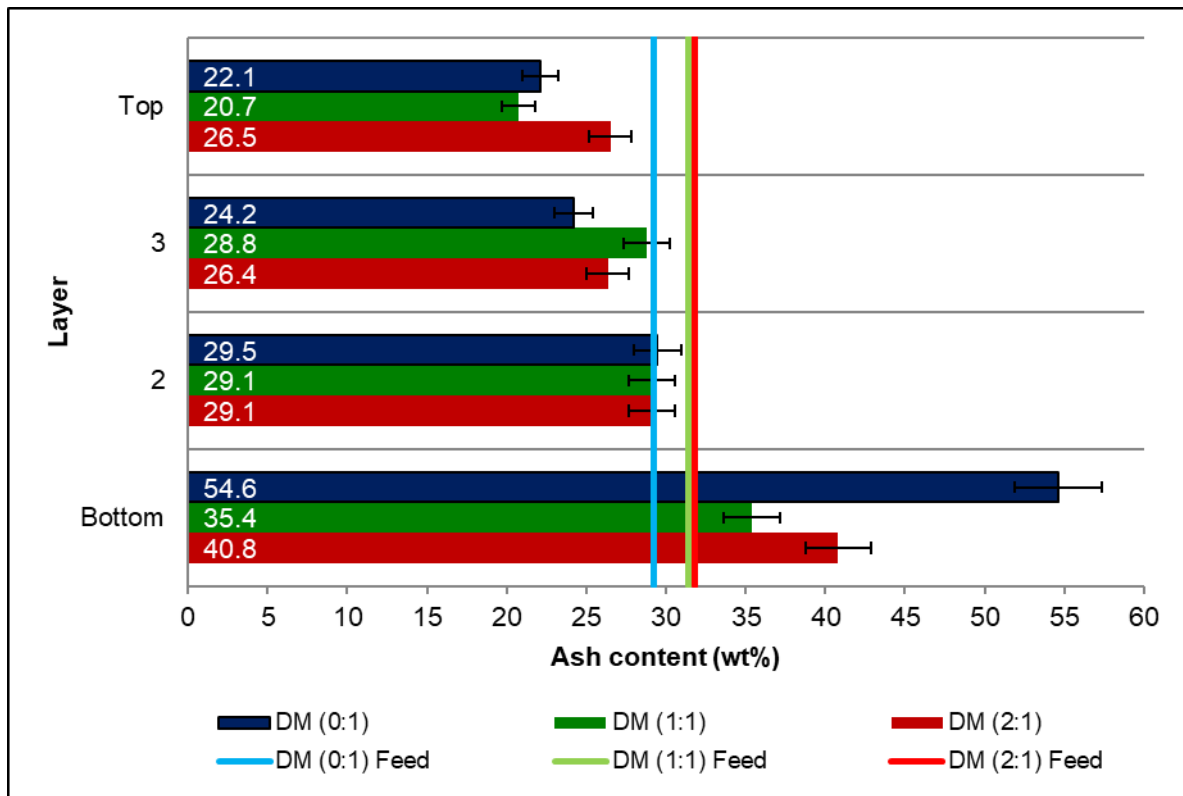


Figure 6.1-13: Comparison of bed performance for DM 0:1, DM 1:1 and DM 2:1 (+2.8mm-4.0mm and VD)

Figure 6.1-13 shows that the +2.8mm-4.0mm coal bed without dense medium generated more acceptable results when considering the top and bottom layer ash yields as well as bed stability, compared to beds with dense medium. An especially distinct difference is observed in between the bottom layers and may be due to the destoning capabilities of the. DM 2:1 show minimal upgrading from the feed.

When fluidizing a bed containing a mixture of dense medium and coal, the magnetite is usually suspended, and the coal simply displace according to the difference in relative density in the bed. The minimum fluidization velocity of the dense medium beds should not vary drastically provided the stationary-unfluidized bed height remains constant. Therefore, when the dense medium ratio is increased, a decrease in U_{mf} observed, not because of the mass or volume of magnetite added but rather due to the difference in the stationary-unfluidized bed height. Addition of dense medium proved to have no significant improvement on the performance or upgrade for all of the particle size ranges tested during this study. This effect seemed to even worsen when dense medium is added to the -2.8mm coal PSD ranges, the pseudo fluid effect of the fluidized bed weakens. To further explain, once dense medium is added to the bed, ultra-fine particles with a high density is added to the ADMFB system which complicates the operational aspects of the bed. These fine, highly

dense particles accumulate within the pores creates between the larger particles, essentially blocking the pores and exposing a congealed surface to the air. This tight packing of coal and magnetite particles in the ADMFB, makes it tough for the upward air flow to perforate the particles, eventually resulting in an impairment of particle segregation and stratification.

Magnetite particles are often used as DM for ADMFB beneficiation due to their good flowability and recoverability through magnetic means (Xu & Guan, 2003). The effective bed density of a magnetite fluidized bed under moderate superficial gas velocities is however not low enough to act as an effective DM for ROM coal particles (Fu, Zhu, Barghi, Zhao, *et al.*, 2019), (Fu, Zhu, Barghi, & Zhao, 2019). Therefore, binary mixtures of magnetite and a lighter material offers a method of dynamically adjusting the bed density so that coal particle segregation is effective under varying operating conditions (Fu, Zhu, Barghi, & Zhao, 2019), (Fu *et al.*, 2020a), (Huang *et al.*, 2017)

6.1.3.2 Vibration behaviour

Macpherson and Gavin (2010) conducted a study which validated that addition of vibration will effectively reduce the minimum fluidization velocity required by the ADMFB. This was further studied by He *et al.* (2015), who found that the addition of vibration allows for a better stabilized pressure performance in the bed, resulting in steady fluidization. The effect of bed vibration is induced, when required, by activating a variable frequency- amplitude 0.18kW oscillating vibratory motor. A 2-pole BM 200/3 model OMB motor was attached to the horizontal base of the bed frame. Variable speed drive enables the operator to adjust the input frequency of vibration. The motor is rated for a maximum of 3000 revolutions per minute (RPM) with amplitude and frequency specifications of 10mm and 50Hz, respectively. A motor efficiency of 69.4% along with a 1.83KN force is expected if weights are set to 100%. The frequency of the motor is the electrical frequency which is received by the motor. By adjusting the weights on the motor, the amplitude output of the motor can be manipulated.

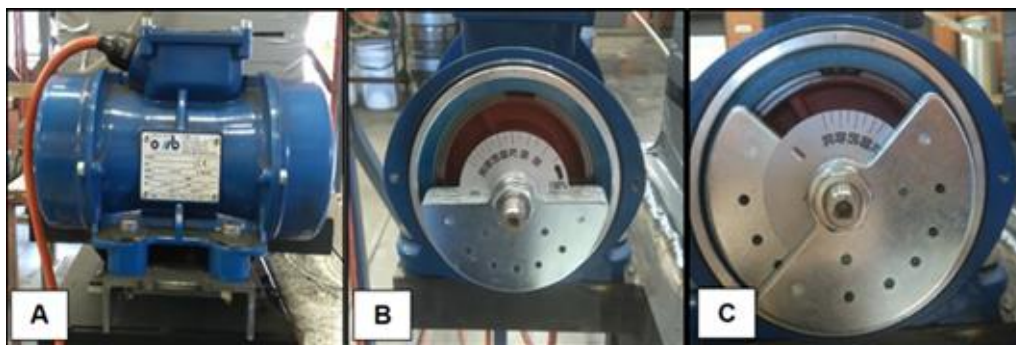


Figure 6.1-14: Oscillating vibration motor (A) adjustments of weights (B and C)

The position of the motor on the frame of the bed induces a horizontal agitating motion of the bed structure. To assist with reinforcing the bed structure during vibration, a steel brace is placed over the fluidized bed portion and bolted onto the base of the bed.

An initial baseline frequency of 20Hz was selected for the experiments that have vibration activated to ensure the vibration is not excessive. A clear decrease in minimum fluidization velocity (MFV) and more stable pressure drop was experienced within the fluidized bed. However, after completing the experimental runs it was found that simply activating and deactivating vibration will not guarantee optimal performance of the bed and so some additional tests were required. Upon activation of vibration, an increased backpressure was experienced, which was indicative of particle segregation resulting in a bed with more densely packed particles. If vibration is induced prematurely, a negative effect on fluidization may be experienced as the ability of particles to move throughout the bed may be suppressed. This behaviour was evident when the bed was dismantled for sampling after fluidization with vibration. The static bed height visible in the top lower of the bed was somewhat lower along the bed height than before the bed was fluidized.

For most of the coal particle sizes investigated, vibration was either activated (frequency of 20Hz applied) or deactivated. However, since addition of vibration at these settings did not significantly contribute to the performance of the ADMFB, the amplitude and frequency effect warranted further investigation. Therefore, the influence that vibration has on the separation performance of two selected PSD's, +4.75mm-5.6mm and 2.8mm-4.0mm, in the ADMFB, was studied. Vibration was tested at amplitudes of 30%, 50% and 70% of the vibratory motor's output and frequencies of 19.86 Hz and 21.97Hz addition of dense medium. Figure 6.1-14 shows an ash value comparison for +4.75mm-5.6mm coal with vibration activated and vibration deactivated.

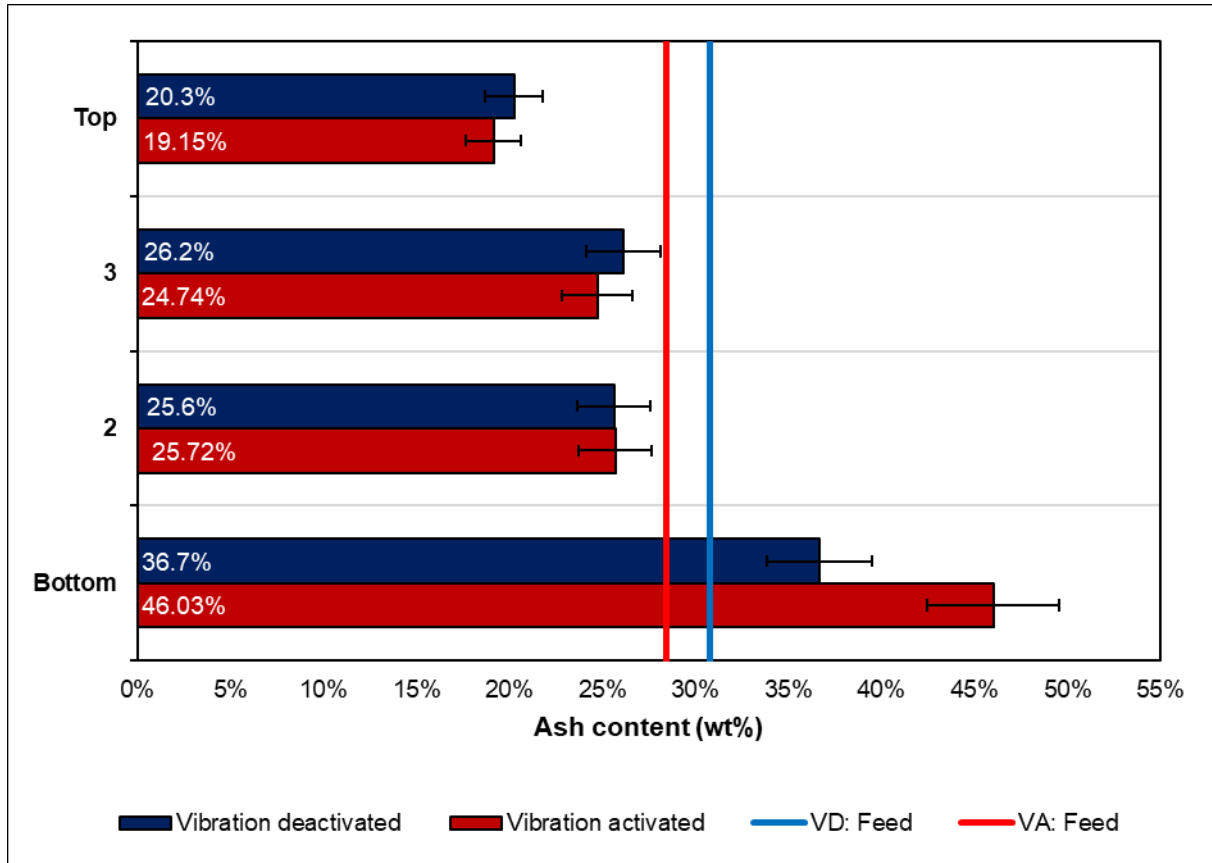


Figure 6.1-15: Ash values for +4.75mm-5.6mm coal with vibration deactivated and activated (DM 0:1)

The vibration activated run in Figure 6.1-14 shows well improved ash differential in the bottom of the bed when vibration is activated. When considering the top three layers of the bed, it was noted that addition of vibration resulted in a more stable bed, with increased ash values from layer 1 to layer 4.

6.1.3.2.1 Influence of amplitude on vibration activated runs

To further investigate the influence of amplitude during vibration activated cases, +4.75mm-5.6mm coal was selected and while maintaining the PSD and frequency constant, the amplitude was varied between 30% and 70%. In Figure 6.1-15, a frequency of 21.97Hz was maintained while the bed was operated.

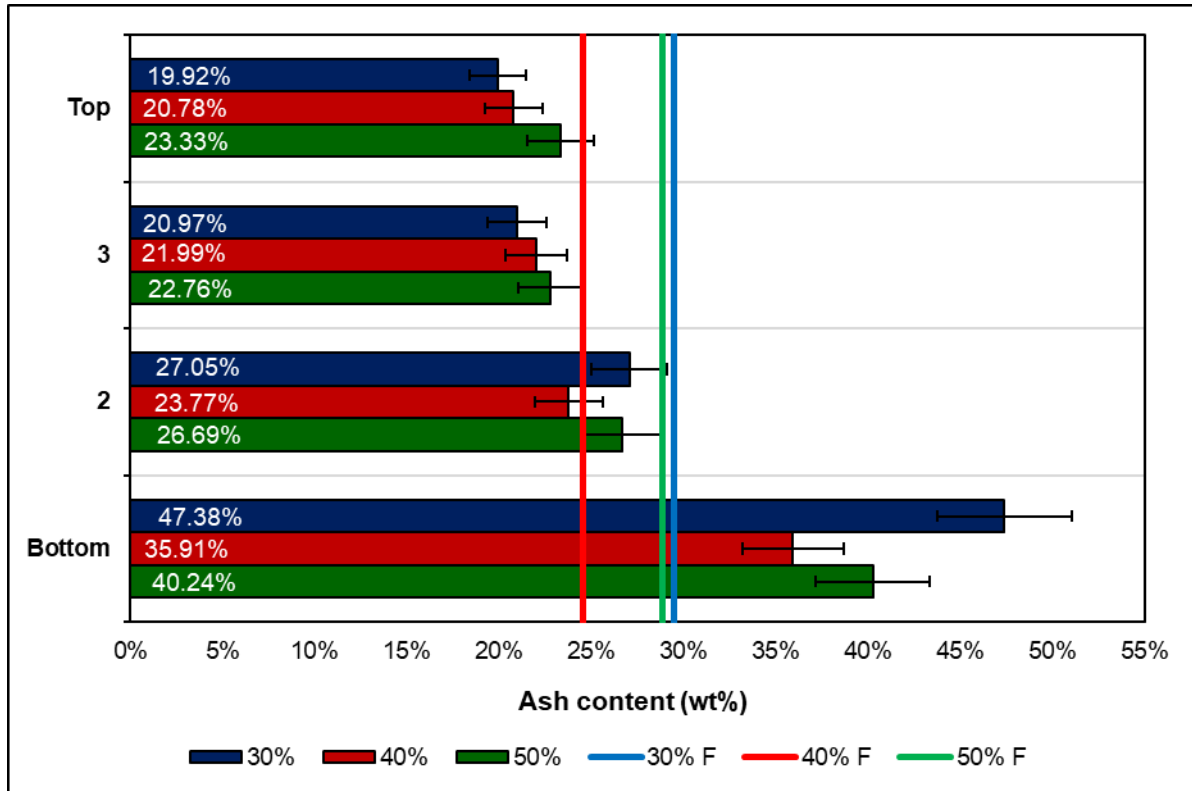


Figure 6.1-16: Ash values for +4.75mm-5.6mm coal at 21.97 Hz

The influence of the amplitude is quite evident especially when considering runs with amplitudes of 30% and 50%, seeing as the feed ash values for these cases differ only marginally. The bottom layer of the bed in Figure 6.1-15, shows that the ash yield at an amplitude of 30% is about 7% more than that obtained for an amplitude at 50%. This implies that the separation performance of the fluidized bed dwindles when the amplitude of vibration is increased. He *et al.* (2015) found that excessive vibrational energy inhibits uniform micro-bubble formation throughout the bed, resulting in reduced separation performance.

6.1.3.2.2 Frequency

In order to evaluate vibration frequency on the separation performance of coal particles in the ADMFB, the amplitude and PSD is kept, while the frequency of the vibration motor were varied between 19.86 Hz and 21.97 Hz. Figure 6.1-16 shows the ash value comparison of +4.75mm-5.6mm coal particles at 30% amplitude subjected to frequencies of 19.86Hz and 21.97Hz. This experimental run was conducted without dense medium.

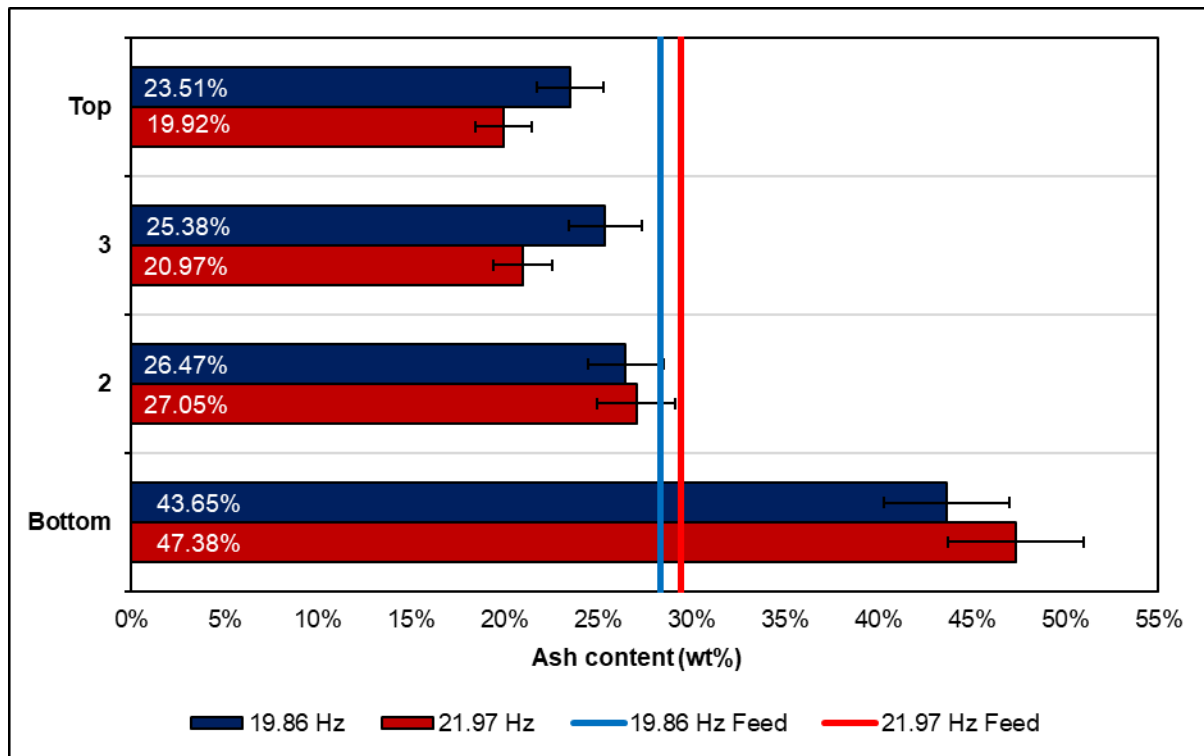


Figure 6.1-17: Ash values for +4.75mm-5.6mm particles at 30% amplitude

The bed layers portrayed in Figure 6.1-16; the ash yield was improved by a frequency increase. By increasing the vibration frequency, a stable bed is created by increased uniform micro-bubble formation that leads to improved bed separation performance (He *et al.*, 2015). However, even though frequency has an influence on the bed fluidization behaviour and separation performance, it was quite marginal in comparison to vibration amplitude.

6.1.3.2.3 Combined effect of amplitude and frequency

Figure 6.1-17 illustrates the impact that various amplitudes at a constant frequency of 19.86Hz has on the separation performance of +4.75mm-5.6mm coal particles.

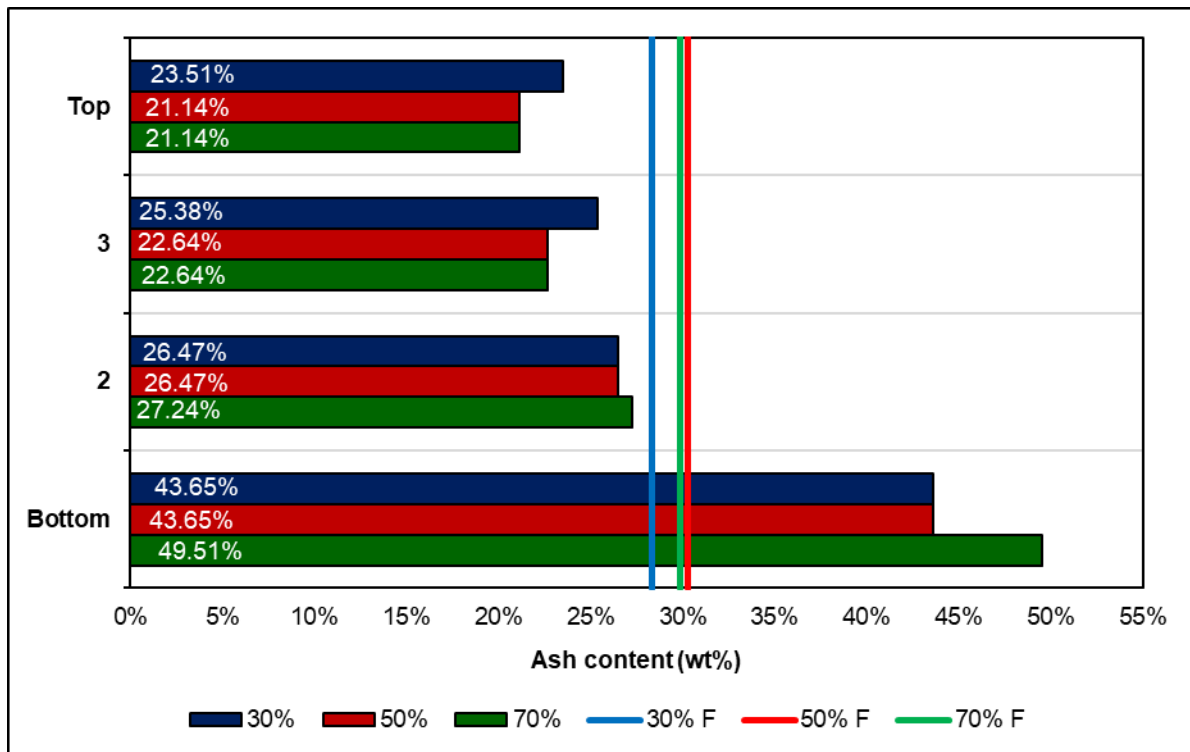


Figure 6.1-18: Ash values for +4.75mm-5.6mm coal particles at 19.86 Hz

It is evident in Figure 6.1-17 that increase ash values are experienced from the top to the bottom for all the amplitudes. When the amplitude of vibration is increased, the variance between the ash values of the feed and bed layers increase. This is particularly noticeable in the bottom bed layer where the feed ash value differs from the bed layer ash values for 30% and 70% amplitudes by 16.24% and 19.67%, respectively. This because the vibration induced at 19.86Hz and 30% amplitude for the +4.75mm-5.6mm coal particles is not intense enough and as there is an increase in intensity of vibration an increase in separation performance is also experienced (He *et al.*, 2015). Based on observation, instabilities are experienced in the fluidization of the bed when the amplitude is increased to 21.97Hz. This may be a result of the vibration intensity being excessively high causing over compaction and expansion in the bed which in turn leads to back mixing (He *et al.*, 2015).

A performance curve, shown in Figure 6.1-18, was constructed to identify the best suited vibration conditions to obtain the superlative bed performance for +4.75mm-5.6mm coal particles.

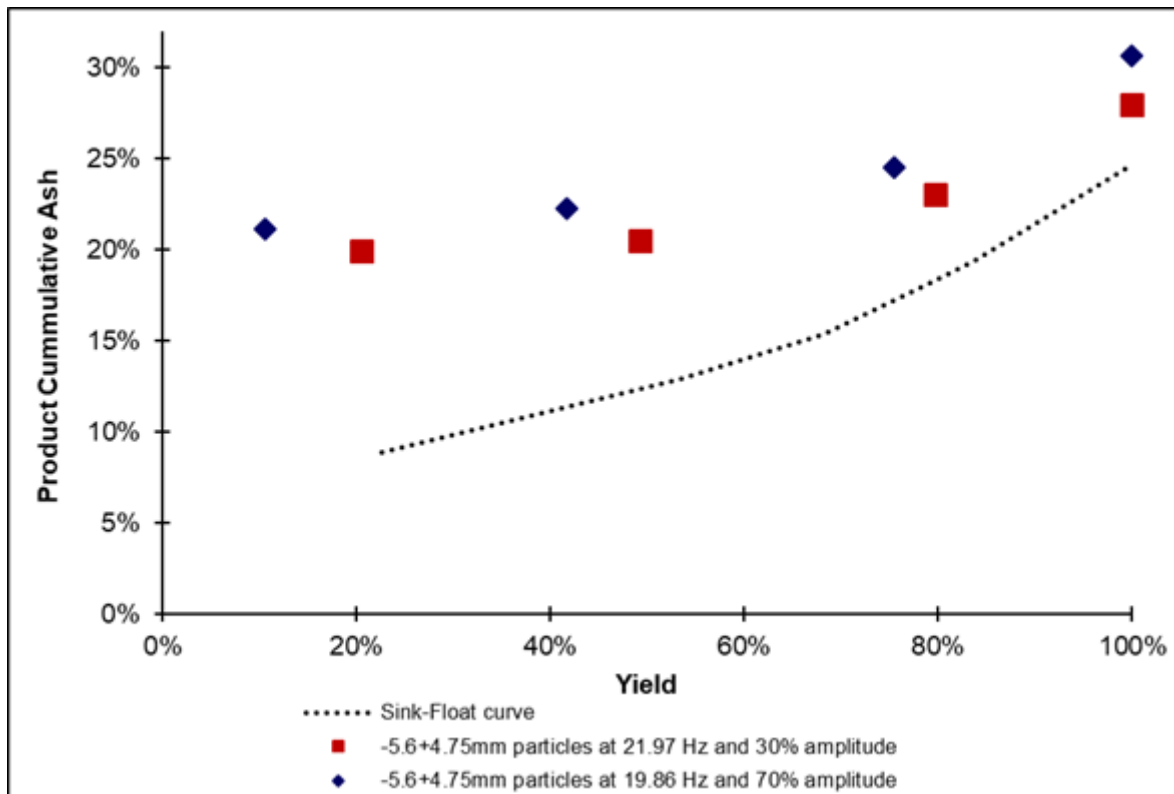


Figure 6.1-19: Performance curves for +4.75mm-5.6mm coal particles

The performance curves in Figure 6.1-18 follows a similar trend, however the datapoints are not within close proximity to the sink-float curve (washability or limiting performance curve). This implies that the ADMFB is not operating within its best performance range yet, nevertheless, results obtained for the 21.97Hz and 30% amplitude run datapoints falls marginally closer to the limiting performance curve. Therefore, for +4.75mm-5.6mm coal particles with added vibration at 21.97Hz and 30% amplitude is most suitable to obtain achieve desired separation performance.

For +2.8mm-4.0mm coal particles at 21.97Hz, none of the vibration combinations were as effective as the combinations at 19.86Hz. As a result, the best combination for upgrading +2.8mm-4.0mm coal particles was identified by comparing the performance curves of the two best results obtained from the best ash values achieved at 19.86Hz. Figure 6.1-19 contains the performance curves for the +2.8mm-4.0mm coal particles selected based on the condition described.

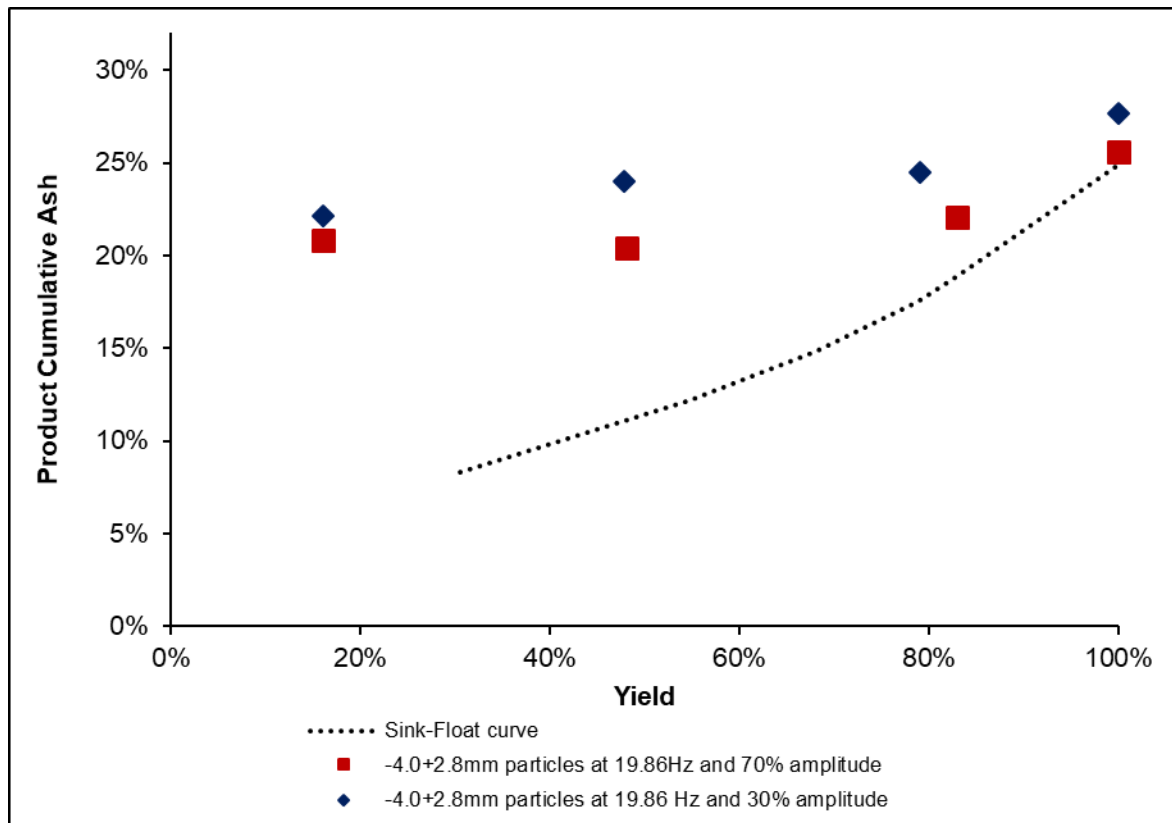


Figure 6.1-20: Performance curves for +2.8mm-4.0mm coal particles

The combination of 19.86Hz and 30% amplitude shown in Figure 6.1-1 lies slightly further from the limiting performance curve. Like the +4.75mm-5.6mm coal particles, the two performance curves in Figure 6.1-19 are not within proximity of the limiting performance curve, indicating the bed has yet to reach full optimization conditions. Nevertheless, for the +2.8mm-4.0mm coal particles, the best frequency and amplitude combination is concluded to be 19.86Hz and 70% amplitude.

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CHAPTER 7

Chapter 7 contains the paper titled “Use of CFD modelling to assist with air dense medium fluidization of South African coal”. In this paper, CFD software, NEPTUNE_CFD, is used to study the dynamic mechanisms of the ADMFB utilized for the experimental portion of this study. This paper aids to better understand experimental findings by investigating the dynamic mechanism of operation of ADMFB. The paper is divided into three portions:

- The effects of material properties and operational parameters on the segregation performance of particles in an ADMFB is investigated
- The fluidization properties of a DM made up from a binary mixture of magnetite and fine coal is investigated
- The ability of a DM binary mixture to effectively separate coal is numerically evaluated

Use of CFD modelling to assist with air dense medium fluidization of South African coal (Paper 4)

This section provides information regarding contributions made by the various authors towards completion of the paper.

Contribution by authors

Misses E.S. Diedericks conducted a coal beneficiation simulation study with the guidance and supervision of Mr. L. Le Grange, Prof. M. le Roux and Prof. Q.P. Campbell. The concept and scope of work for this portion of the study was formulated by Mr. L. Le Grange, Misses E.S. Diedericks, Prof. M. le Roux and Prof. Q.P. Campbell. Misses E.S. Diedericks was responsible for conducting the experimental work, collecting data and processing of the collected data. The results from experimental runs were discussed and assessed during multiple consultations between Misses E.S. Diedericks, Mr. L. Le Grange, Prof. M. le Roux and Prof. Q.P. Campbell. Mr. L. Le Grange assisted with the modelling portion of the paper. The CFD software used for the numerical simulations is called NEPTUNE_CFD and follows a multifluid Eulerian approach with a polydispersed fluid-particle capability implemented by The Institute of Fluid Mechanics of Toulouse (IMFT). A 3-D rectangular mesh was set up by Mr. L Grange representing the experimental fluidized bed. Numerical simulations were performed using a parallel computer at the Center for High Performance Computing (CHPC) in Cape Town. After completion of the numerical simulations, the results were compared to the experimental results. After conclusions were drawn, the paper was discussed and reviewed by Misses E.S. Diedericks, Mr. L. Le Grange, Prof. M. Le Roux and Prof. QP. Campbell.

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7.1 Use of CFD modelling to assist with air dense medium fluidization of South African coal

ES Diedericks, L Le Grange, M le Roux and QP Campbell*

School of Chemical and Minerals Engineering, North-West University, Potchefstroom, South Africa.

**marco.leroux@nwu.ac.za*

Abstract

This paper is focussed on evaluating the effects of material properties and operational parameters on the segregation of particles in an air dense medium fluidized bed (ADMFB) with the help of Computational Fluid Dynamics (CFD). In this study, CFD will be employed to improve the understanding of the dense medium (DM) selection process for best particle segregation. The bed consisted of monodispersed ash, coal and dense medium (DM) particles. The fluidized DM particles provide a bed density that acts as a cut density in separating coal from ash on the principle of buoyancy.

During this study, segregation of particles in a ADMFB were varied through a parametric study involving DM particles size, DM material density, superficial gas velocity and the ratio of DM to coal and ash. The CFD results showed that segregation was successfully obtained from different combinations of fluidization velocity and DM properties. Moreover, the properties of a DM bed were investigated where the DM is represented by a binary mixture of heavy and light particles, which in this case was a combination of magnetite and fine coal. By varying fluidization velocity as well as the coal to magnetite ratio, it was found that a homogenous bed could be obtain at low fluidization velocity for a low DM composition ratio.

Lastly, an ADMFB was simulated using a binary mixture of fine coal and magnetite as the DM to separate coal from gangue particles. The simulations showed the range in fluidization velocities and DM properties that could be effective to separate high- and low-density coal particles.

Keywords: Coal, ash, dry beneficiation, dense medium (DM), magnetite, air dense medium fluidization, minimum fluidization velocity, Computational Fluid Dynamics (CFD)

Introduction

South Africa's national energy-resource base is dominated by coal; with coal meeting about 77% of the countrywide energy demand (DOE, 2022).

SA has vast amounts of low-quality coal implicating a high ash content. Harmful elements in ash include hard minerals such as quarts that causes wear of mills and boiler internals. Certain coal also contains pyritic sulphur that is released after combustion as Sulfur Dioxide (SO_2) polluting the air. Feeding ROM coal directly to coal-fired power stations along with the continuous increase in global energy demand will inevitably increase air pollution. Coal with high moisture and high ash forming mineral content, referred to as low quality coal, is also considered the main reason for low efficiencies on coal-fired power stations (Chen & Yang, 2003).

SO_2 is commonly removed using Flue Gas Desulphurisation (FGD) in a post-combustion process. Another method is to remove pyritic sulphur from raw coal during coal beneficiation in a pre-combustion process. Coal beneficiation also remove hard minerals such as quarts that causes wear on mills and boiler internals. Coal beneficiation is therefore important from an economic, health and environmental perspective. Coal beneficiation techniques may include wet and dry beneficiation techniques. Currently industrialized, wet beneficiation processes generate high separation efficiencies however, it requires large volumes of water. Various challenges are associated with wet beneficiation, including handling ability, storage and transport issues. In terms of economics, capital requirement increases, and excessive operating and transportation cost arise from excess water retained in coal (Choung *et al.*, 2006). Moreover, water is a scarce commodity in coal producing and processing countries, such as South Africa, Australia and Mongolia, which makes wet coal beneficiation techniques not ideal in such countries (Chen & Yang, 2003; Zhao *et al.*, 2015). The industry is researching dry coal beneficiation methods in the hopes of finding a suitable alternative that is capable of producing good separation efficiencies. Dry beneficiation techniques, such as Air Dense Medium Fluidized Bed (ADMFB) system, is a solution to remove mineral matter content from ROM coal to enhance coal quality and alleviate application issues as well as reduce carbon footprint.

In a coal-magnetite ADMFB, the magnetite is fluidized (in suspension) and the coal particles float or sink in accordance with density. The selection of a good dense medium (DM) (density, particle size and volume) is complex because its properties together with the coal properties and bed operational parameters are inter-dependant in optimizing an ADMFB.

Coal can theoretically be separated in an ADMFB where a fluidized DM provides a bed density cut point separating coal from gangue (high ash forming mineral matter content) on the basis of buoyancy (Singh *et al.*, 2019), (Huang *et al.*, 2017). This desired bed density can be manipulated by varying several operational and material properties of the fluidized bed and the DM respectively. These properties included fluidization velocity, density of the DM and DM particle size. Fluidization velocity will directly affect bed density through bed porosity. Too high fluidization velocities may result in over-mixing and will inhibit segregation. The particle size of the DM can also be varied however, it should not be too small due to increased intergranular viscous forces and not be too big due to increased minimum fluidization velocity and accompanying bubbling velocity characteristics (Zhao *et al.*, 2015).

A study conducted at the NWU investigated Air Dense Medium Fluidised Bed (ADMFB) as a suitable dry beneficiation technique to upgrade small South African coal (Diedericks *et al.*, 2020). The separation performance of the ADMFB was evaluated by monitoring the effect of feed particle size, DM (100% magnetite) to coal feed ratios and the addition of vibration. In the majority of the cases, superior cumulative products were obtained for ADMFB experimental runs that employed no DM and no addition of vibration (Diedericks *et al.*, 2020). Furthermore, high ash values were obtained in the bottom layers of the coal beds without DM, which was indicative of better de-stoning capabilities in these beds (Diedericks *et al.*, 2020). This phenomenon was unexpected as it was not in line with that reported in literature. The function of DM is to stabilize and enhance separation capabilities of the ADMFB system and since the experimental results was not portraying that, the decision was taken to further explore the DM selection process.

However, selecting a dense medium based on a specific density is not practical due to the limited availability of suitable DM materials. Magnetite particles are often used as DM for ADMFB beneficiation due to their good flowability and recoverability through magnetic means (Xu & Guan, 2003). The effective bed density of a magnetite fluidized bed under moderate superficial gas velocities is however not low enough to act as an effective DM for ROM coal particles (Fu, Zhu, Barghi, Zhao, *et al.*, 2019), (Fu, Zhu, Barghi, & Zhao, 2019). Therefore, binary mixtures of magnetite and a lighter material offers a method of dynamically adjusting the bed density so that coal particle segregation is effective under varying operating conditions (Fu, Zhu, Barghi, & Zhao, 2019), (Fu *et al.*, 2020a), (Huang *et al.*, 2017).

In order to comprehend the role of DM in the segregation process, computational fluid dynamics (CFD) was employed to investigate the particle dynamics of the ADMFB. The

study consists of three steps; Firstly, the effects of material properties and operational parameters on the segregation of particles in an ADMFB is considered. Secondly, the fluidization properties of a DM made up from a binary mixture of magnetite and fine coal is investigated and finally a DM binary mixture's ability to effectively separate coal is numerically evaluated.

Methodology

Material properties:

Coal from the number 4 seam in the Witbank area in South Africa was used for the experimental study. The coal from this seam is bituminous, a thermal grade product (Jeffrey, 2005).

Table 7.1-1: Average initial proximate and calorific values for each coal particle size range

<i>Description</i>	<i>Moisture content (%wt)</i>	<i>Volatiles (%wt)</i>	<i>Ash yield (%wt)</i>	<i>Fixed carbon (%wt)</i>	<i>Gross calorific value (MJ/kg)</i>
+0.5mm-13.2mm	0.61-2.46	21.4-24.2	27.5-32.6	42.9-49.5	21.7-25.5
<i>Standards</i>	ACT-TPM-010 based on SANS 5925: 2007	ACT-TPM-012 based on ISO 562: 2010	ACT-TPM-011 based on ISO 1171: 2010	By difference	ACT-TPM-014 based on ISO 1928: 2009

Magnetite powder, with a d_{50} size of 490 μ m, was used as a dense medium. The true density of the magnetite was found by helium pycnometry (SANS 1014:1985) to be 4800 kg/m³.

Experimental setup

Experiments were conducted in a batch operated air dense medium fluidized bed (ADMFB) unit that consisted of three sections namely the *airflow system*, *fluidized bed* and *dust control* section, as shown in Figure 6.1-1. The *airflow system* consists of an air blower, airflow sensor (pitot tube) and air-distribution compartment coupled to an air distribution plate. Eight rectangular frames (0.3m x 0.3m x 0.05m) were stacked on top of the distribution plate to form a fluidizing unit having a height of 0.4m. Each layer could be detached and removed separately from the unit to aid in obtaining a sample cut of material at a certain height. The unit was extended at the top by a further 0.3m to allow enough space for the particles to fluidize. The *dust control* section of the ADMFB unit is composed of a tapered trapezoidal

extension (20° hopper angle), covered by a 100µm stainless steel wired mesh. This limited the carry-over of ultra-fine particles by a reduction in air velocity at the top of the unit.

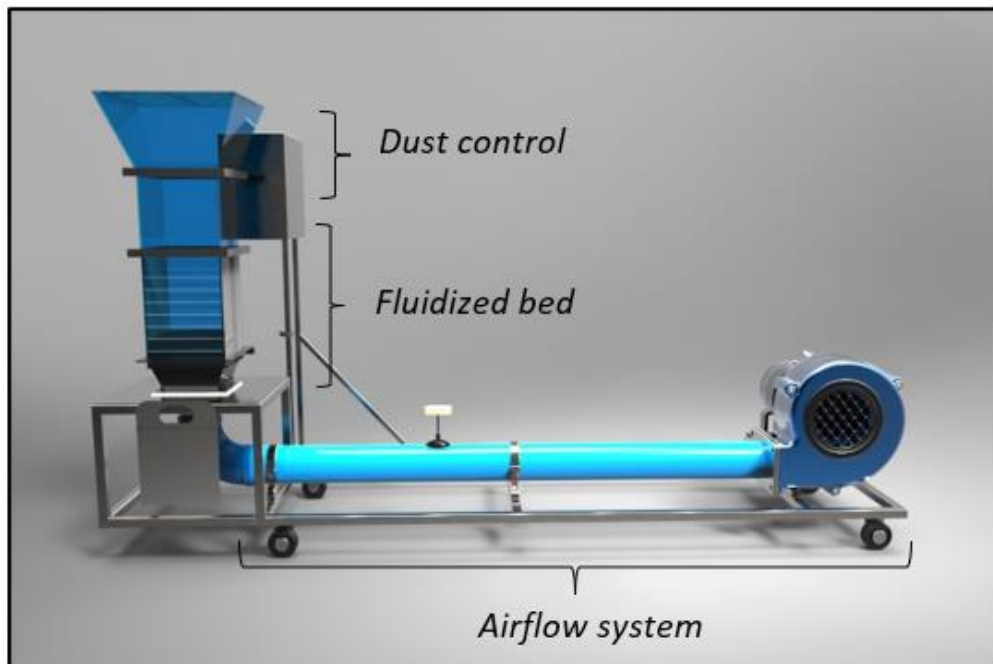


Figure 7.1-1: Three-dimensional illustration of the ADMFB

Simulation model

Theory

The CFD software used for the numerical simulations is NEPTUNE_CFD (<https://www.code-saturne.org/cms/web/NEPTUNECFD>). The code is developed in the framework of the NEPTUNE project financially supported by the Commissariat for Atomic Energy and Alternative Energies (CEA), Electricity of France (EDF), The Institute for Radiation Protection and Nuclear Safety (IRSN) and AREVA-NP. The code is based on a multifluid Eulerian approach with a poly-dispersed fluid-particle capability implemented by The Institute of Fluid Mechanics of Toulouse (IMFT) (Neau et al., 2013).

The theoretical approach outlined below describes the continuum-based methodology followed in NEPTUNE_CFD as described in (Fotovvat et al., 2015).

The Eulerian multi-fluid approach used is a hybrid approach (Morioka and Nakajima, 1987) and with the transport equations derived by phase ensemble averaging for the continuous phase and by using kinetic theory of granular flows supplemented by fluid and turbulent effects for the dispersed phase through a joint fluid–particle probability density function (PDF) approach. In the proposed modeling approach, transport equations for mass,

momentum and fluctuating kinetic energy are solved for each phase and coupled through inter-phase transfer terms. The momentum transfer between gas and particle phases is modeled using the drag law of Wen and Yu limited by Ergun equation for the dense flows (Gobin *et al.*, 2003; Peirano *et al.*, 2002). The collisional particle stress tensor is derived in the frame of the kinetic theory of granular media (Boelle *et al.*, 1995). The fluid turbulence modeling is achieved by the two equations of k - ϵ model extended to particle-laden flows accounting for additional source terms due to the inter-phase interactions (Vermorel *et al.*, 2003). For the dispersed phase, a coupled transport equation system is solved on particle fluctuating kinetic energy and fluid-particle fluctuating covariance. The effects of the particle-particle contact force in the very dense zone of the flow are taken into account in the particle stress tensor by the additional frictional stress tensor (Srivastava and Sundaresan, 2003).

Mesh and boundary conditions

A 3-D rectangular mesh with dimensions 300mm (width) x 300mm (depth) x 500mm (height) is used representing the experimental fluidized bed. The cell sizes are uniform with 5mm edges. Numerical simulations are performed with parallel processing using a supercomputer at the Center for High Performance Computing (CHPC) in Cape Town utilizing 48 cores and with each core handling 7500 computational cells.

At the bottom of the bed a uniform superficial inlet velocity is imposed corresponding to the fluidization velocity and with a zero flux of particles. At the top of the fluidized bed, a free outlet for both gas and particles is defined. The wall-type boundary condition is friction for the gas and a no-slip one for the particles (Fede *et al.*, 2009).

Results and discussion

The numerical investigation comprises three aspects; Firstly, a parametric study to better understand the fundamental principles of an ADMFB and in particular the role of the DM particle size, DM particle density, DM to coal ratio and fluidization velocity. Secondly a parametric study is performed to investigate DM suitability when made up of a binary mixture of magnetite and fine coal and finally the evaluation of a DM binary mixture of magnetite and fine coal to effectively separate a coal particle mixture.

ADMFB parameter study

The purpose of this parameter study is to investigate the fundamental principles of density-based particle segregation using a fluidized bed with a dense medium. The combined effects of DM particle size, DM material density, fluidization velocity and the ratio of DM to coal are simulated to predict whether the bed is in a complete mixed state, whether segregation takes place and also if the DM settles. The phase properties and parameter envelope are given in Table 7.1-2 and Table 7.1-3 respectively. The individual case studies are shown in Table 7.1-4.

Table 7.1-2: Properties of gas and particulate phases

Property	Units	Value
Gas density	kg/m ³	1.18
Gas viscosity	Pa.s	1.85--5
Ash (C1) density	kg/m ³	1900
Ash (C1) size	mm	5
Clean coal (C2) density	kg/m ³	1400
Clean coal (C2) size	mm	5
Magnetite (M) density	kg/m ³	4600
Magnetite (M) size	mm	0.232
Fine coal density	kg/m ³	1400
Fine coal size	mm	0.396

Table 7.1-3: Parametric study variables

Variable	Units	C1	C2	DM
Particle size	mm	5	5	0.1, 0.3
Particle density	kg/m ³	1900	1400	2800, 3200, 3600, 4000
Gas velocity	cm/s	30, 24, 22, 20, 18, 16, 14, 8, 6, 4, 3, 2, 1		
Volumetric solids ratio (C1:C2:DM)	%	12.5 : 12.5 : 75, 25 : 25 : 50, 37.5 : 37.5 : 25		

Table 7.1-4: ADMFB case definitions

Case	DM size [mm]	DM density [kg/m ³]	Velocity [cm/s]	Volumetric ratio [1] (C1 : C2 : DM)
1	1	4000	8, 6, 4, 3, 2, 1	(37.5 : 37.5 : 25)
2	3	4000	30, 24, 22, 20, 18, 16, 14	(37.5 : 37.5 : 25)
3	1	3600	8, 6, 4, 3, 2, 1	(37.5 : 37.5 : 25)
4	3	3600	30, 24, 22, 20, 18, 16, 14	(37.5 : 37.5 : 25)

5	1	3400	8, 6, 4, 3, 2, 1	(37.5 : 37.5 : 25)
6	3	3400	30, 24, 22, 20, 18, 16, 14	(37.5 : 37.5 : 25)
7	1	2800	8, 6, 4, 3, 2, 1	(37.5 : 37.5 : 25)
8	3	2800	30, 24, 22, 20, 18, 16, 14	(37.5 : 37.5 : 25)
9	1	4000	8, 6, 4, 3, 2, 1	(25.0 : 25.0 : 50)
10	3	4000	30, 24, 22, 20, 18, 16, 14	(25.0 : 25.0 : 50)
11	1	3600	8, 6, 4, 3, 2, 1	(25.0 : 25.0 : 50)
12	3	3600	30, 24, 22, 20, 18, 16, 14	(25.0 : 25.0 : 50)
13	1	3400	8, 6, 4, 3, 2, 1	(25.0 : 25.0 : 50)
14	3	3400	30, 24, 22, 20, 18, 16, 14	(25.0 : 25.0 : 50)
15	1	2800	8, 6, 4, 3, 2, 1	(25.0 : 25.0 : 50)
16	3	2800	30, 24, 22, 20, 18, 16, 14	(12.5 : 12.5 : 75)
17	1	4000	8, 6, 4, 3, 2, 1	(12.5 : 12.5 : 75)
18	3	4000	30, 24, 22, 20, 18, 16, 14	(12.5 : 12.5 : 75)
19	1	3600	8, 6, 4, 3, 2, 1	(12.5 : 12.5 : 75)
20	3	3600	30, 24, 22, 20, 18, 16, 14	(12.5 : 12.5 : 75)
21	1	3400	8, 6, 4, 3, 2, 1	(12.5 : 12.5 : 75)
22	3	3400	30, 24, 22, 20, 18, 16, 14	(12.5 : 12.5 : 75)
23	1	2800	8, 6, 4, 3, 2, 1	(12.5 : 12.5 : 75)
24	3	2800	30, 24, 22, 20, 18, 16, 14	(12.5 : 12.5 : 75)

Solution initialization

The bed is initialized with solid fractions for C1 (Ash), C2 (Coal) and DM (dense medium). Initially conditions are set so that air makes up 60% of the bed volume up to a height of 0.2 m from the base plate and the remaining 40% filled with solids according to the volumetric ratios (C1 : C2 : DM) given in Table 7.1-4..

Simulations are performed for each of the 24 cases defined in Table 7.1-4 starting with a large fluidization velocity to induce complete mixing and then slowly decreasing until the DM settles at the bottom of the bed. Between each change in velocity, 5 seconds of physical simulation time is allowed for flow adjustment after which results are recorded while performing time-averaging for a further 10 seconds. Transient simulations are performed with an averaged time step of 0.0002 seconds.

Bed component solid fractions

With four phases in the bed namely air, C1, C2 and DM, a typical ADMFB (Case 11) is simulated under conditions of complete mixing, effective segregation and DM settling. Color contour plots of DM solid fraction for the 3 modes of mixing in the bed are shown in Figure 7.1-2, Figure 7.1-4 and Figure 7.1-6 respectively. Mixture component solid fractions along the height for the 3 modes of mixing, averaged over line A, line B and line C (see contour plots) are shown in Figure 7.1-5, Figure 7.1-5 and Figure 7.1-7 respectively.

- Complete mixing

DM solid fractions in the bed for Case 11 with complete mixing are shown in Figure 7.1-2 and Figure 7.1-3. Large bubbles are clearly visible on the color contour plot. The line graph shows mostly horizontal lines, which indicate that the solid fraction of the components remain almost constant along the bed height.

- Effective segregation

As the fluidization velocity for Case 11 is reduced, segregation starts. Effective segregation conditions are shown in Figure 7.1-4 and Figure 7.1-5. On the contour plot one can see that the DM solid fraction is the highest in the middle of the bed indicating that C1 probably settled on the floor while C2 is probably floating at the top of the bed. The line graphs of solid fractions of C1, C2 and DM confirm this.

- DM settling

Solid fractions of DM in the bed for Case 11 at lower fluidization velocities are shown in Figure 7.1-6 and Figure 7.1-7. The DM starts to settle at the bottom as can be seen in the color contour plot. The solid fraction line graph also shows the highest concentration of DM at the bottom of the bed where it starts to displace C1.

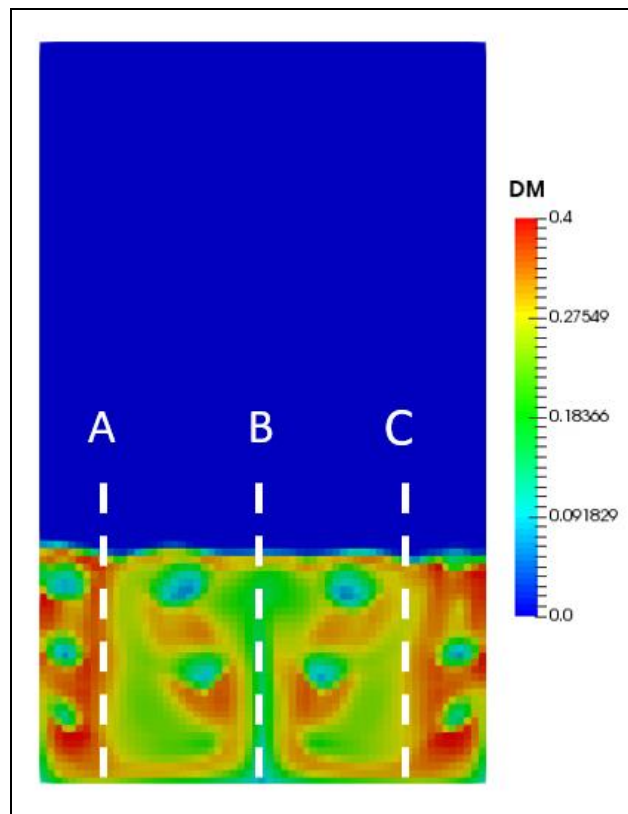


Figure 7.1-2: Case 11 – complete mixing – fluidization velocity too high

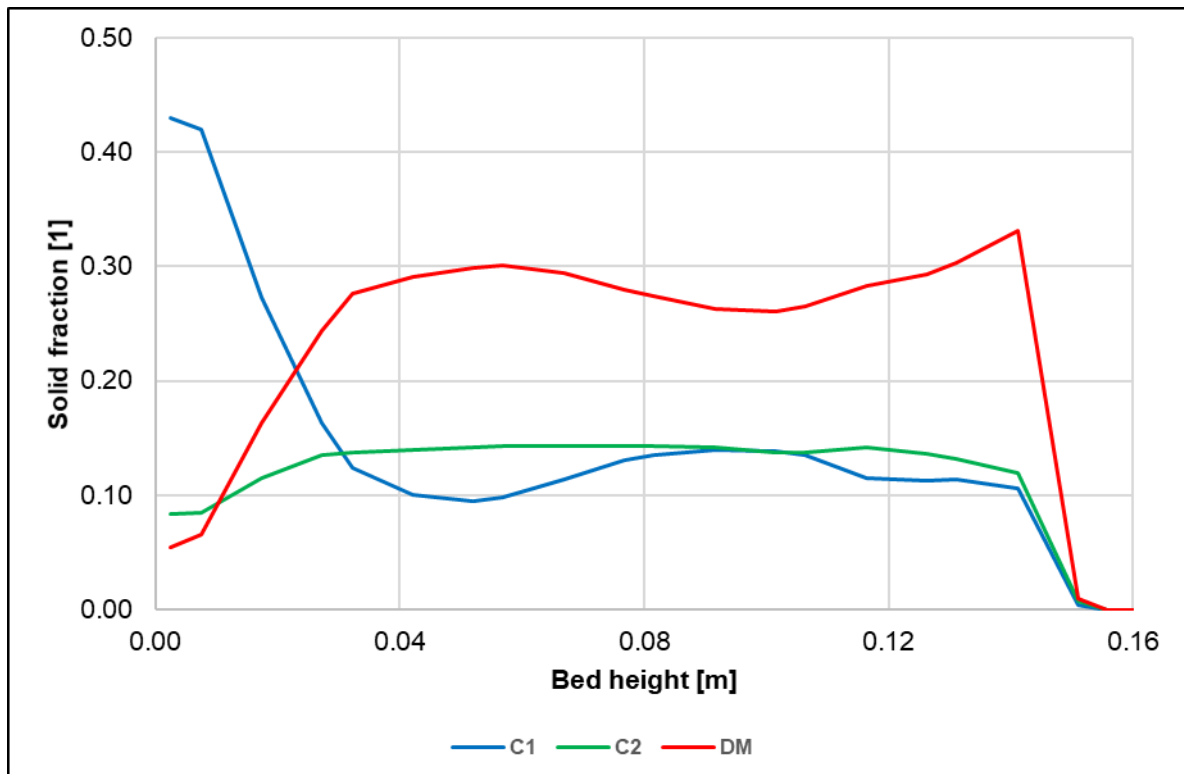


Figure 7.1-3: Case 11 – complete mixing - vertical solid fractions of components

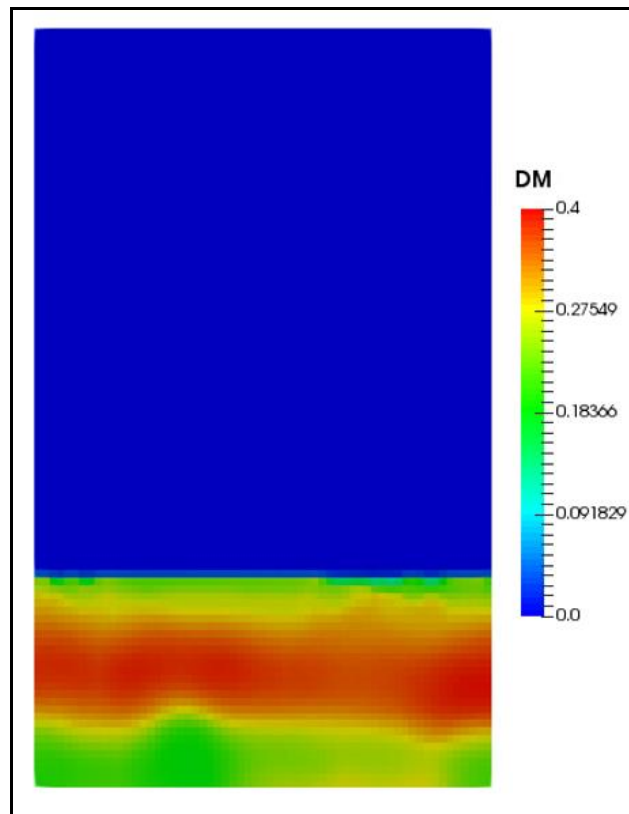


Figure 7.1-4: Case 11 – successful segregation

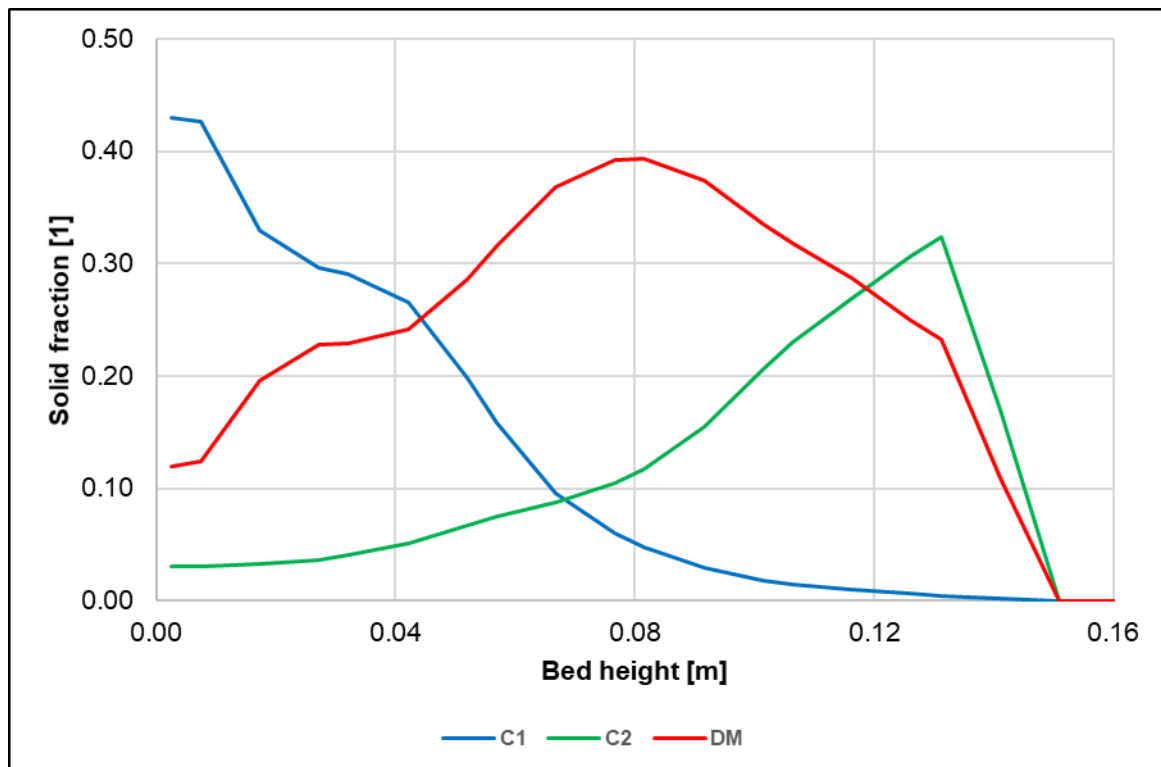


Figure 7.1-5: Case 11 – successful segregation - vertical solid fractions of components

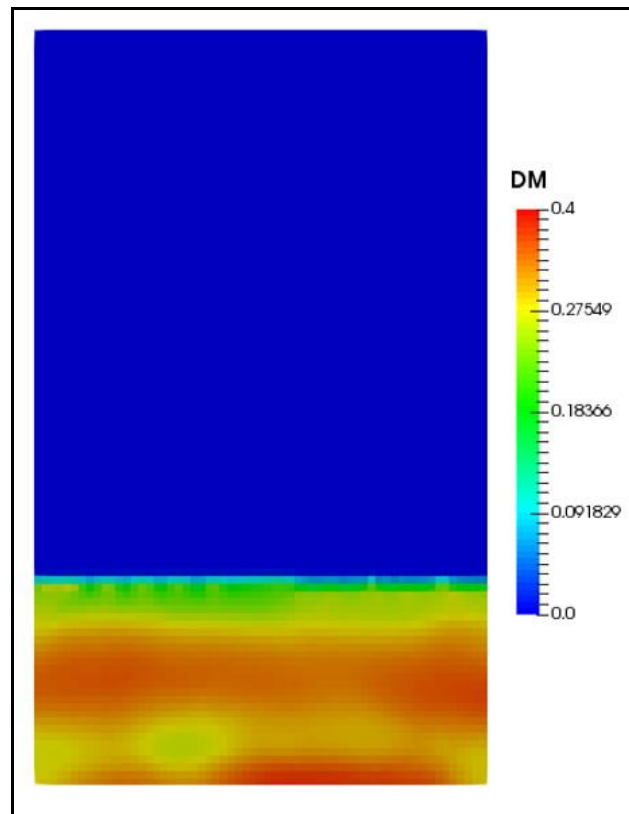


Figure 7.1-6: Case 11 – DM settling – fluidization velocity too low.

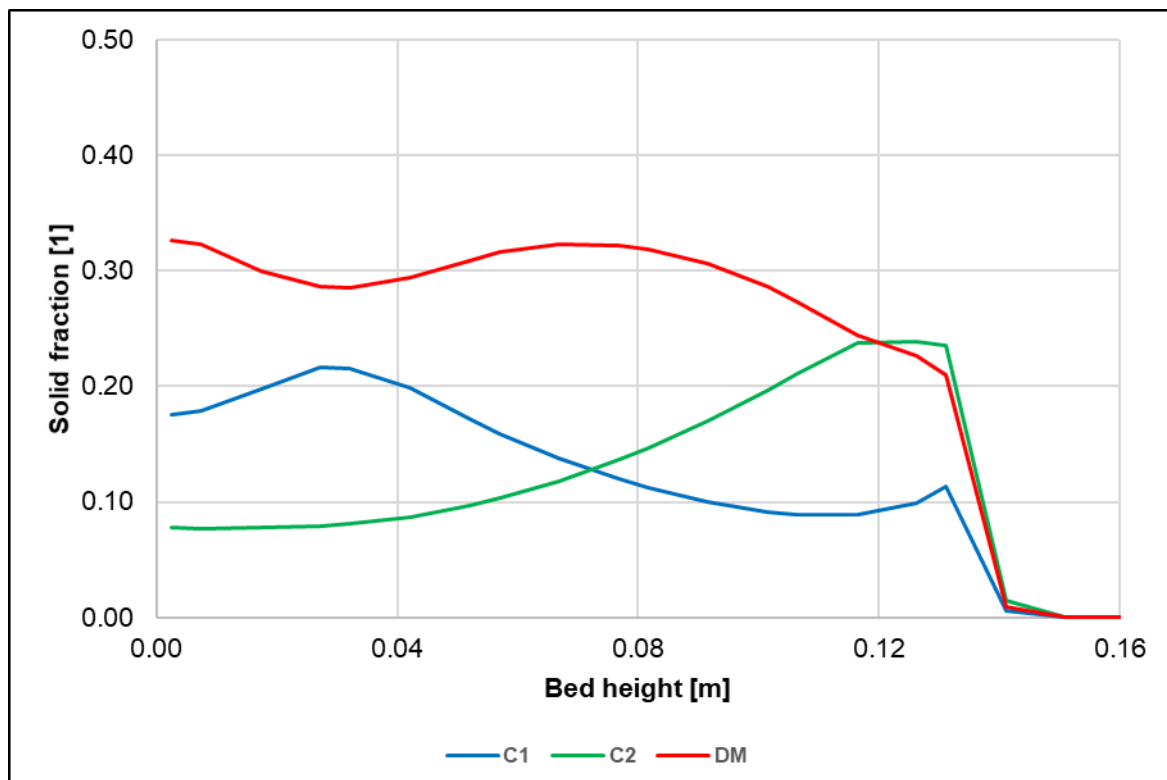


Figure 7.1-7: Case 11 – DM settling - vertical solid fractions of components

Extend of segregation

In an ADMFB where density-based segregation is accomplished through buoyancy, less dense particles will rise to the top (flotsam), more dense particles will sink to the bottom (jetsam) while most DM particles will find themselves wedged in the middle. A vertical Center-Of-Mass (COM) calculation will show three distinct heights – jetsam having the lowest COM, flotsam the highest COM and with the DM in between. With complete mixing this vertical order in COM will clearly be disrupted. The extend of segregation of components in various mixtures and under various fluidization conditions are therefore presented in terms of the COM of each component along the height of the bed using the following calculation for each component:

$$i = \frac{\sum h_j \phi_{ij}}{\sum \phi_{ij}} \quad \text{Equation 6.1}$$

with h_j the height at vertical increment j in the bed and ϕ_i the solid volume fraction of component i .

Identifying effective particle segregation

The extend of segregation of solid components (represented as the vertical COM) for Case 11 is displayed in Figure 7.1-8. The COM data is calculated at each of the line probes A, B and C (as shown in Figure 7.1-2), averaged and then displayed for different fluidization velocities. The blue, red and green bars in Figure 7.1-8 represent the COM of C1 (1900kg/m³), C2 (1400kg/m³) and DM (3600kg/m³), respectively.

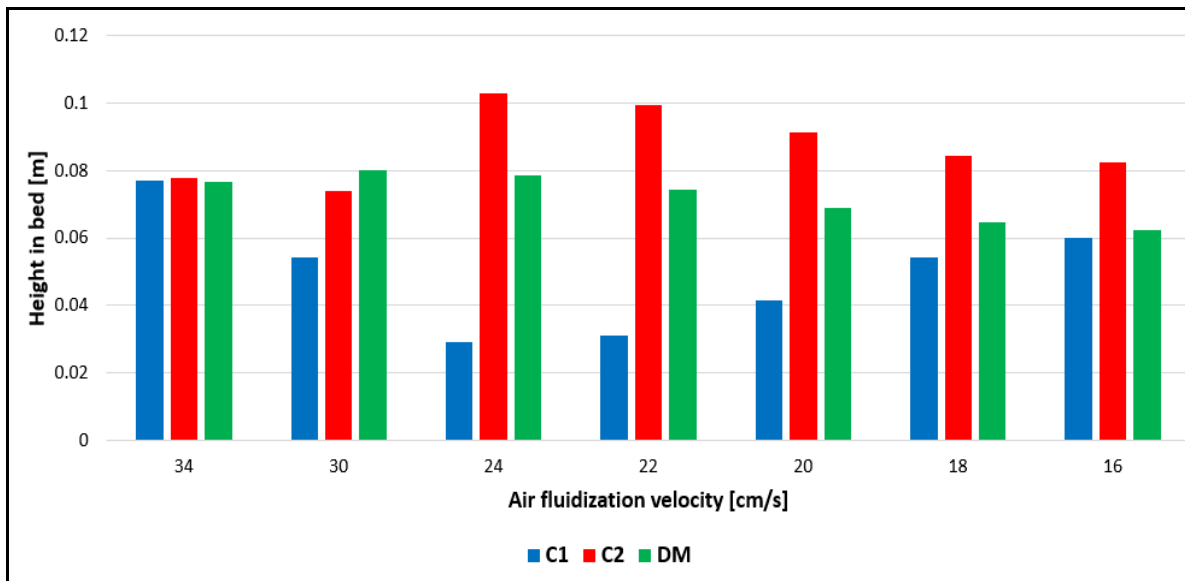


Figure 7.1-8: Case 11 – Segregation of 5mm coal fluidized with 0.3mm dense medium at various fluidization velocities

In this case the combination of heavy material (C1) and the light material (C2) represents a coal sample with more dense particles (coal with large fraction of ash forming minerals) and less dense particles (coal with small fraction of ash forming minerals). The volume fraction in a collapsed bed for C1:C2:DM was 0.25:0.25:0.5, respectively. From Figure 7.1-8, effective segregation occurs at fluidization velocities from 24 cm/s down to 20 cm/s where the top of the green bar (COM of DM) is situated vertically between that of the blue bar (COM of C1) and the red bar (COM of C2). Figure 7.1-8 can therefore be used to select the best fluidization velocity for effective segregation. This order in height reflects the effects of buoyancy where the heavier material (C1) sinks to the bottom while the lighter material (C2) floats at the top.

Effect of fluidization velocity

The fluidization in each of Cases 1 to 24 follows a similar trend that it starts with complete mixing at high superficial air velocities and then particle segregation improves as the fluidization velocity is lowered. At too low fluidization velocities the DM particles starts to settle with the effect that the bed loses its capacity to segregate the C1 and C2 particles at low velocities on the basis of buoyancy.

Effect of DM density

The combination of DM density together with fluidization velocity (for a given DM particle size) is most important for it determines the effective bed density intended to separate heavy and light particle on the basis of buoyancy. The bed density can be altered either by increasing or decreasing the fluidization velocity or by increasing or decreasing the DM material density. In general, lower fluidization velocities are preferable from a bed stability point of view. In such a case a lower DM material density may be required to produce an effective cut-density.

The material density of the DM was varied between 2800 and 4000 kg/m³ in the CFD simulations. It turned out that the selection of DM material density of 2800kg/m³ did not produce successful segregation (Cases 7, 8, 15, 16, 23 and 24) while higher density values resulted in segregation in the fluidization velocity ranges considered. In Case 23 (see Figure 7.1-9), complete mixing prevails over the full fluidization velocity range due to the bed density being too low resulting in the DM floating on top of C1 and C2.

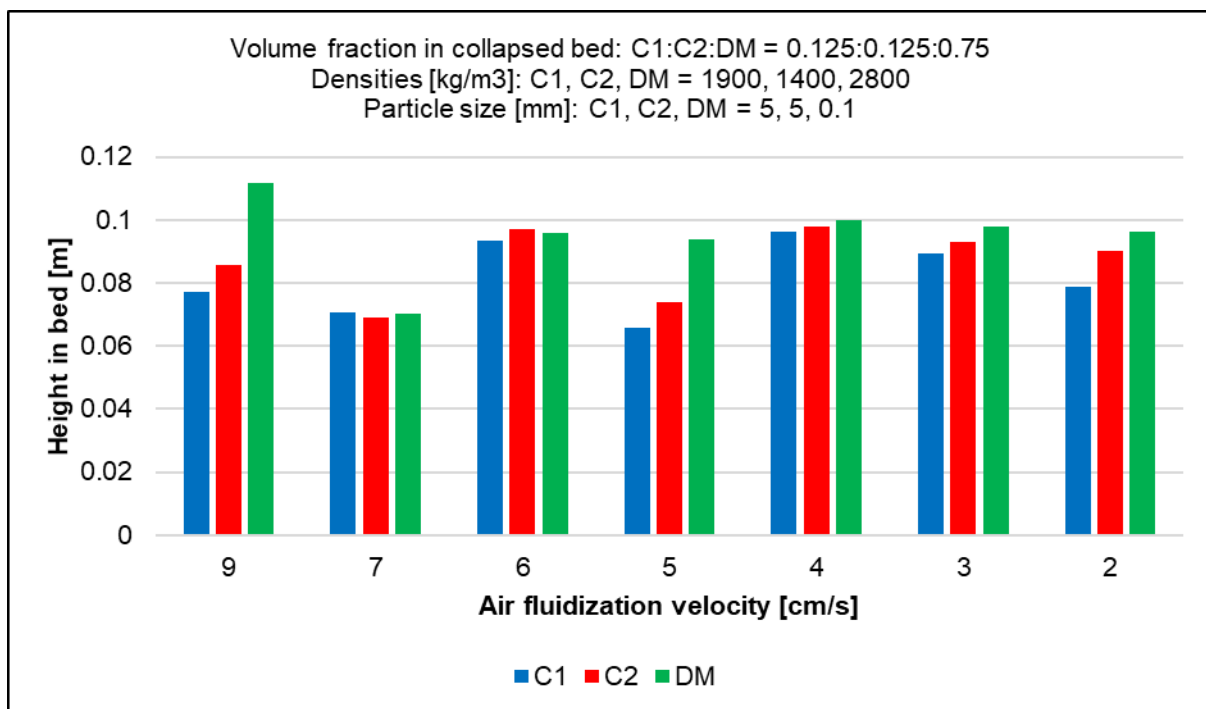


Figure 7.1-9: Case 23 – Complete mixing occurs over the entire range of fluidization velocities.

Also, from the results of these successful densities (3200, 3600 and 4000kg/m³) one could see that using a less dense DM resulted in effective segregation at a lower fluidization velocity, which is desirable from a bed stability point of view.

Case 18 (Figure 7.1-10), demonstrate the case where the dense medium settles almost over the full fluidization velocity range due to the bed density being too high. C1 and C2 floats on top of the DM.

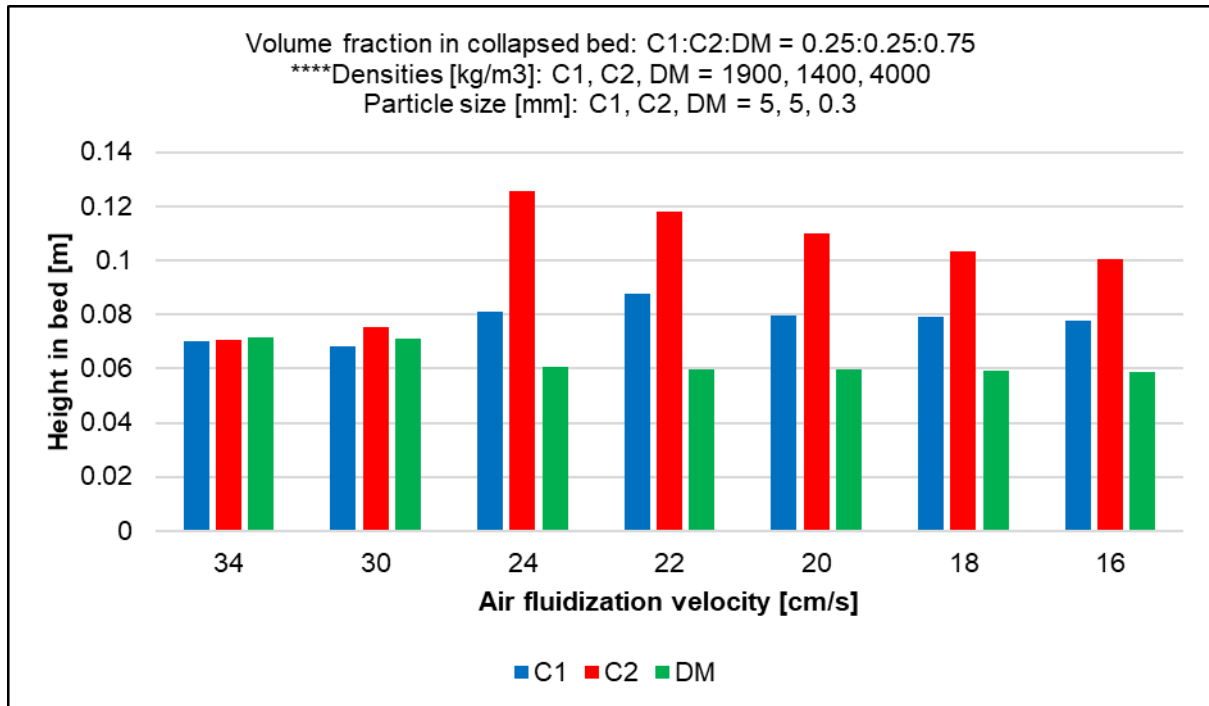


Figure 7.1-10: Case 18 – DM settling in fluidized bed.

Effect of DM particle size

The particle size of the DM plays a crucial role in the ability to fluidize and in the economics of segregation. Segregation of particles C1 and C2 are shown for DM particle sizes of 0.1mm and 0.3mm in Figure 7.1-11 and Figure 7.1-12 respectively. Segregation using 0.1mm DM particles occurs at velocities an order of magnitude lower than for 0.3mm DM particles. Lower superficial gas velocities require less power to fluidize. Small DM particles are however more difficult to fluidize due to increased surface forces including electrostatics (Rokkam et al., 2013; Jalalinejad et al., 2015).

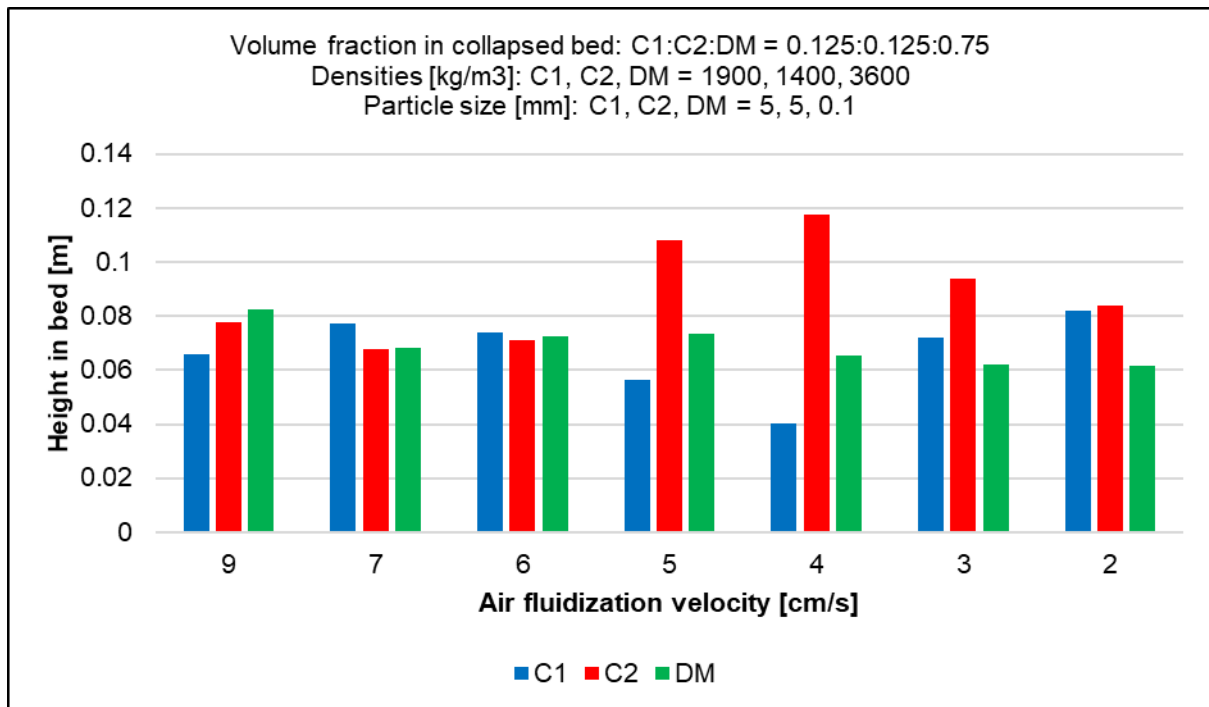


Figure 7.1-11: Case 19 – Particle segregation with 0.1mm DM particle size.

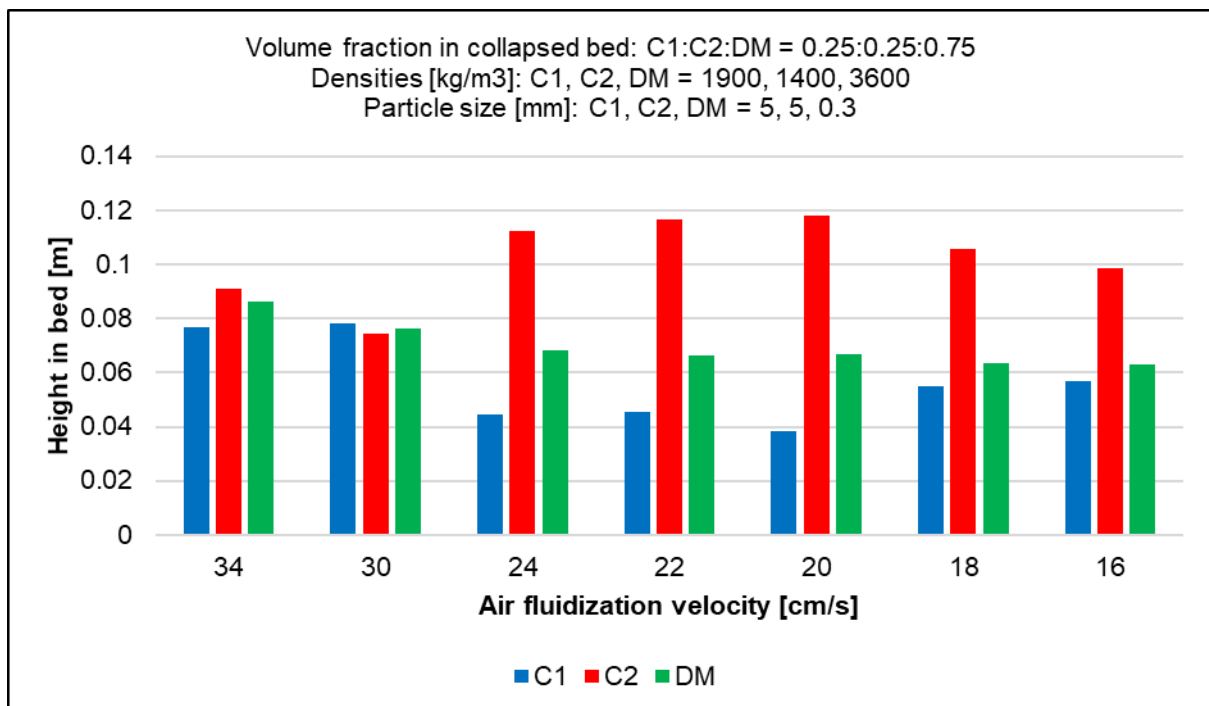


Figure 7.1-12: Case 20 – Particle segregation with 0.3mm DM particle size.

Effect of ratio between DM and coal

The volumetric ratio of DM to coal (C1, C2) is varied between 25%, 50% and 75% for the DM while the remaining part is split in half between C1 and C2. Comparison of the simulation results using these fractions are shown in Figure 7.1-13, Figure 7.1-14 and Figure 7.1-15. The most effective segregation occurs when the DM makes up most of the bed solid volume

At low DM solid fractions, the C1-C2 particles are spaced much closer implicating increased surface friction, which inhibits free movement in the bed and subsequently limits segregation.

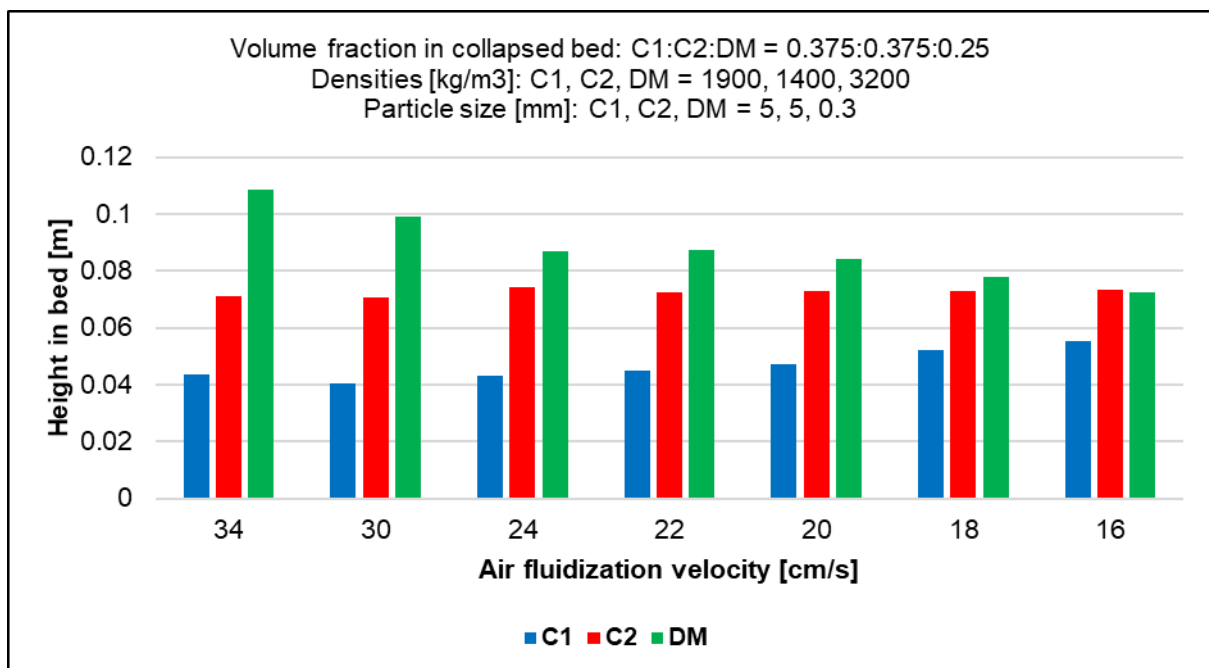


Figure 7.1-13: Case 6 – DM volumetric ratio to coal (25 : 75)

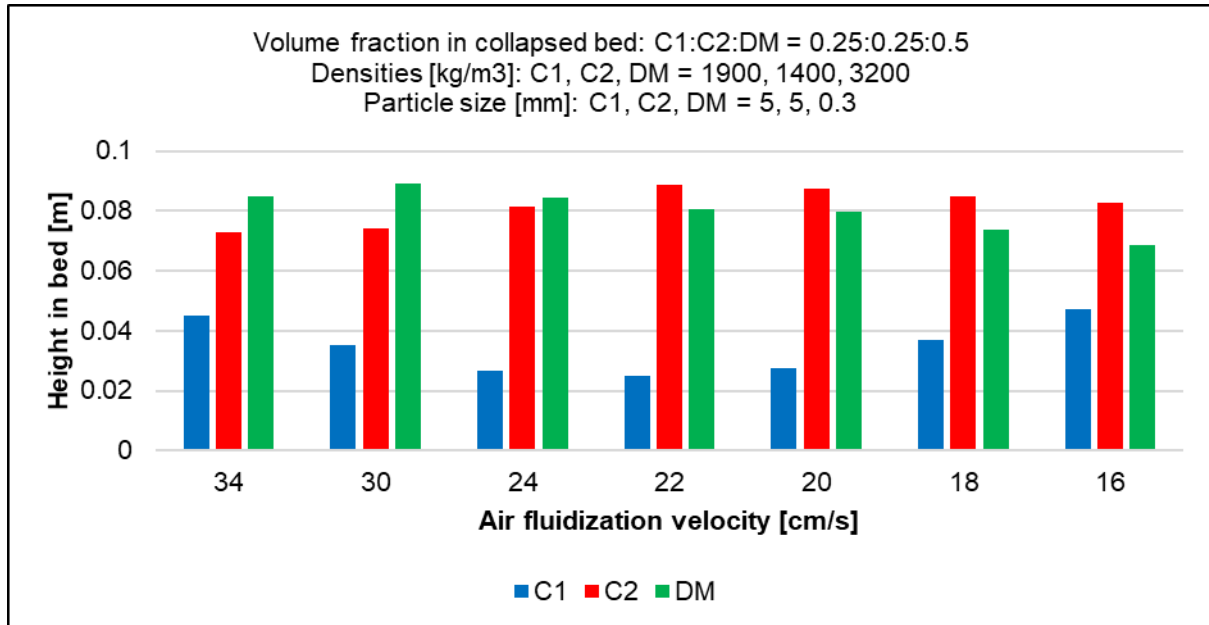


Figure 7.1-14: Case 14 – DM volumetric ratio to coal (50 : 50)

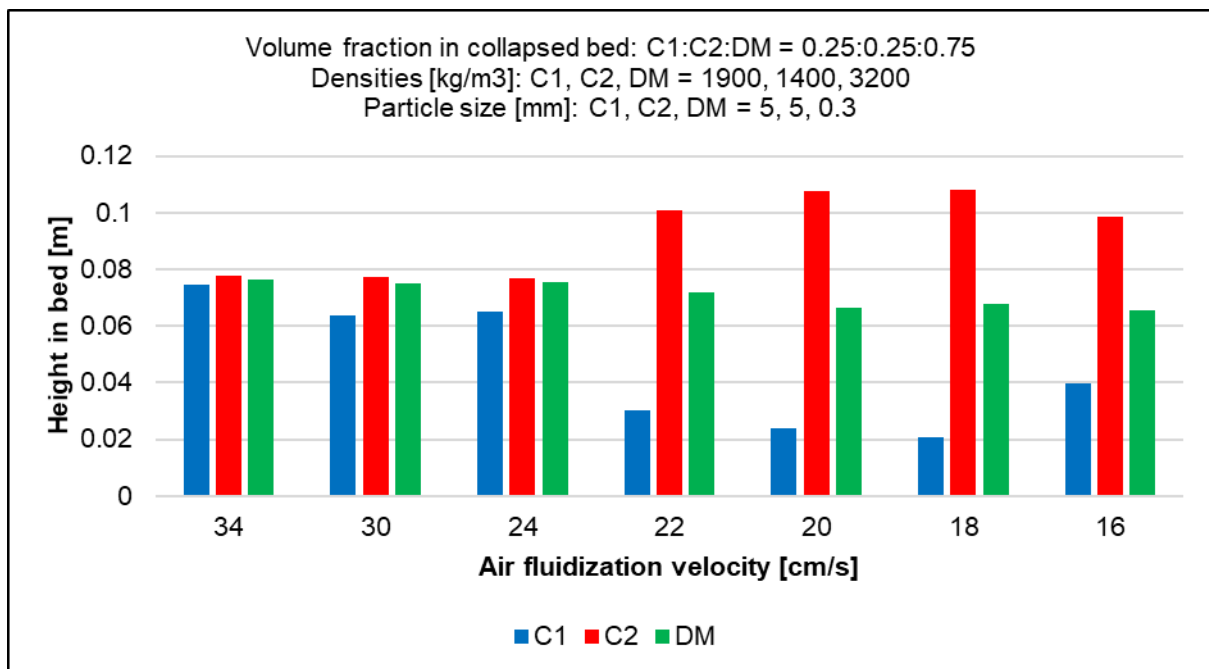


Figure 7.1-15: Case 20 – DM volumetric ratio to coal (75 : 25)

3.2 DM binary mixture parameter study

In practice, the selection of an appropriate DM is challenging since there are a limited number of suitable materials available. Examples are magnetite, ilmenite, ferrosilicon and sand (Fu *et al.*, 2020a). These materials have very specific material densities and most often not in a desirable range. However, the bed density obtained by fluidizing magnetite is generally higher than that required for effective dry coal beneficiation (Wei *et al.*, 2003; Yoshida *et al.*, 2008), since the ADMFB is a gravity-based separation method.

From the results presented in the previous sections, different bed conditions require specific DM densities in order to maximize segregation. In order to overcome this challenge the use of binary mixtures of DM materials have been used – typically a heavy and a much lighter component (Singh *et al.*, 2019), (Turrado *et al.*, 2018), (Fu, Zhu, Barghi, Zhao, *et al.*, 2019). A requirement is however that the binary mixture be well mixed to act as an effective DM. When the components of the DM have different properties such as size or density the mixture itself may segregate and that will limit the DM's effectiveness for particle segregation on the basis of density.

Research has been done on various binary mixtures to establish the suitability as an effective DM material (Fu *et al.*, 2020a). The study included the effects of particle density ratio, particle size ratio, effect of binary mixture composition, effect of superficial velocity and the effect of bed height. Three binary mixtures were considered namely magnetite mixed with sand, gangue and fine coal. In this study we will consider only one DM mixture of Magnetite and fine coal to establish its mixing characteristics under various superficial gas velocities and mixture composition and if the results compares with the findings of (Fu *et al.*, 2020a).

3.2.1 Effect of mixture composition and superficial velocity

A parametric study is performed to investigate the behavior of the binary mixture under different component compositions as well as superficial gas velocity. Four different volumetric ratios of magnetite to fine coal are considered and in each case the gas superficial velocity is varied from a high value producing complete mixing down to a low value where settling starts to take place. The material properties of the magnetite and fine coal mixture is taken the same as for a case considered by Fu *et al.*, 2020b and these binary mixture volumetric ratios (fine coal and magnetite) is displayed in Table 7.1-5 (Case 25-28).

Table 7.1-5: Binary mixture cases

Case	Velocity [cm/s]	Volumetric ratio [1]
25	30, 22, 18	92 : 8
26	30, 22, 18	88 : 12
27	30, 22, 18	75 : 25
28	30, 22, 18	63 : 37

An effective binary mixture that can act as a suitable DM needs to be well mixed with minimal segregation and be fluidized to such a degree that the bed density can act as a cut-density to separate the clean coal from the gangue. The solids distribution of a fluidized mixture of magnetite and fine coal is shown for four cases in Figure 7.1-16, Figure 7.1-17, Figure 7.1-18 and Figure 7.1-19. The figures show the solid volume fraction as a function of superficial gas velocity vertical in the bed. A well-mixed DM will display as vertical lines which indicates a uniform distribution of solid material throughout the entirety of the bed height. A lateral distribution is indicative of segregation of the components of the binary mixture, which is undesirable.

From the CFD results in Figure 7.1-16 (Case 25) to Figure 7.1-18 (Case 28), it is evident that mixing occurs at velocities 30cm/s (red line) and 22cm/s (blue line) but with settling of magnetite occurring at 18cm/s. The lower fluidization velocity of 22cm/s (blue line) would be preferred from a segregation point of view as this would minimize possible back mixing. When comparing composition of mixture fraction it shows that effective mixing at 22m/s only occurs for low volume fractions of fine coal, which is consistent with the findings of Fu *et al.*, 2020b.

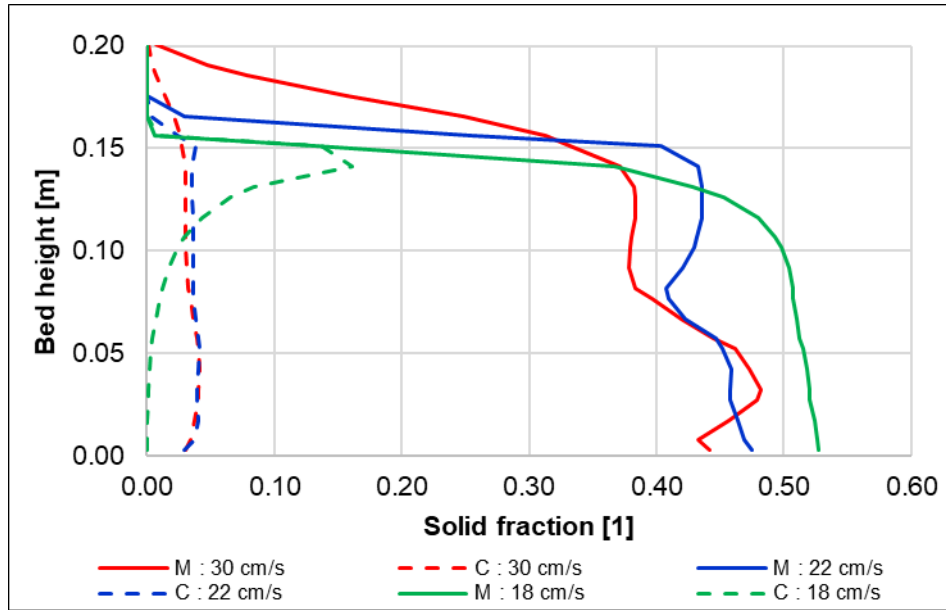


Figure 7.1-16: Binary mixture (DM ratio magnetite to coal 92:8)

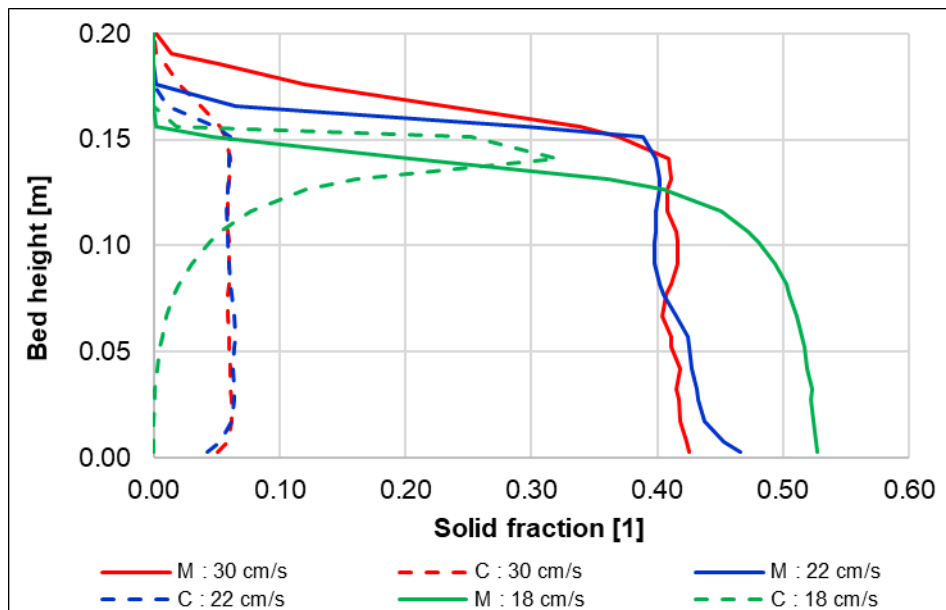


Figure 7.1-17: Binary mixture (DM ratio of magnetite to coal 88:12)

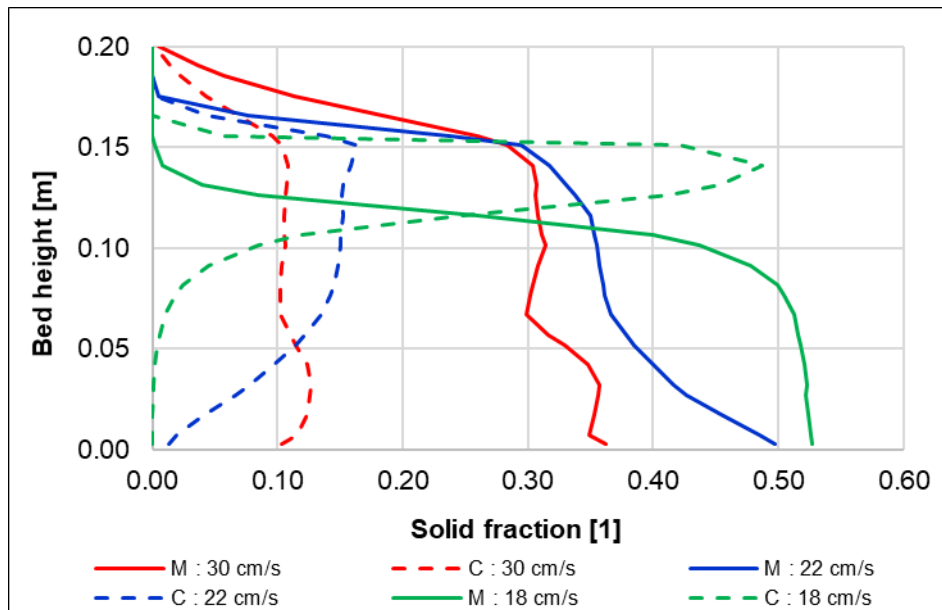


Figure 7.1-18: Binary mixture (DM ratio of magnetite to coal 75:25)

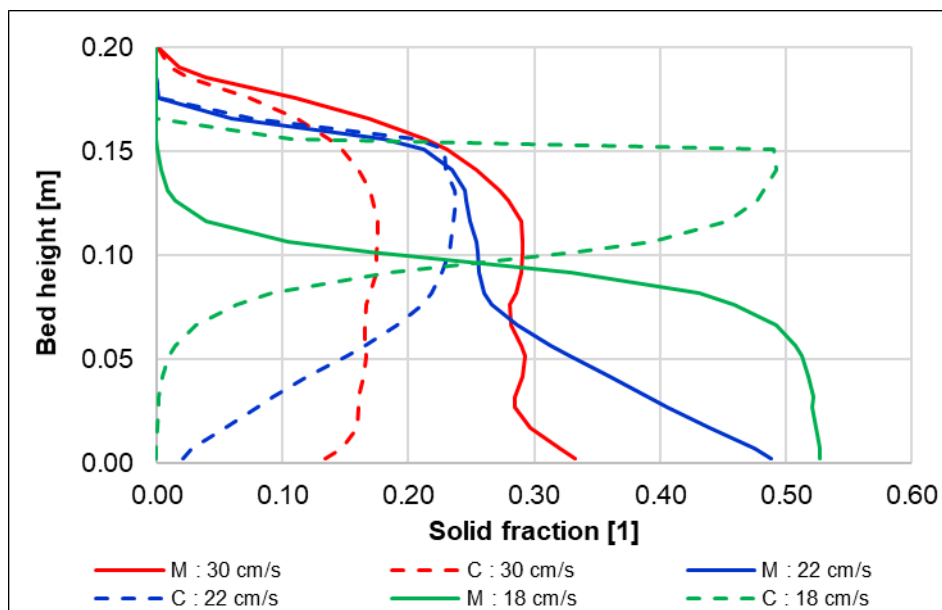


Figure 7.1-19: Binary mixture (DM ratio of magnetite to coal 63:37)

3.3 DM binary mixture application

An application of particle segregation using a binary mixture of magnetite and fine coal acting as a DM is simulated. The properties of the materials are shown in Table 7.1-2. The mixture is fluidized while the superficial gas velocity is slowly lowered from a high (complete mixing) to a low value (settling). The solid fractions of the solid particle components are shown in Figure 7.1-20 and Figure 7.1-21.

From Figure 7.1-20 one can see that all material components are fully mixed with a gas velocity of 29cm/s. This is evident from the almost vertical solid fractions of all four components along the height of the bed. When the gas velocity is lowered to 23cm/s one can see clear segregation of particles C1 and C2 while the binary mixture of magnetite and fine coal still shows fairly vertical solid fraction distributions, i.e., no segregation of the binary mixture components.

The comparison of solid fraction distribution for lower velocities is shown in Figure 7.1-21. It is clear that magnetite and fine coal completely segregated with the result of poor segregation between C1 and C2.

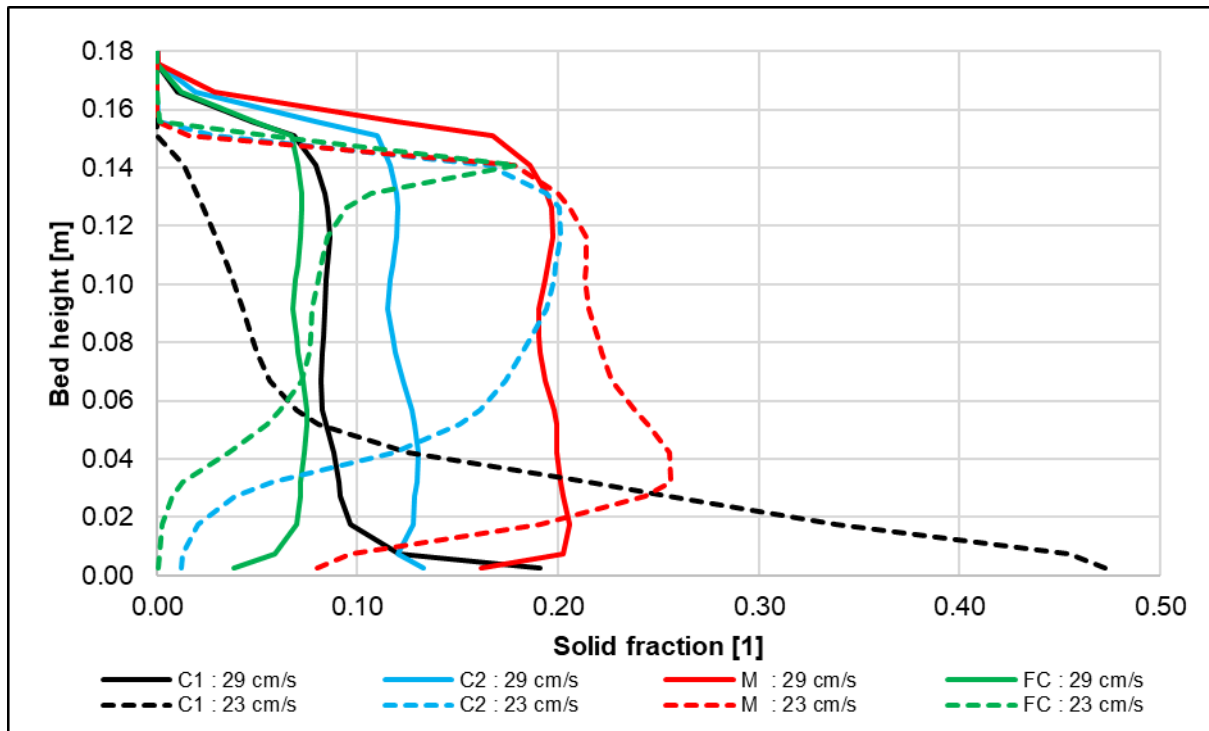


Figure 7.1-20: Particle Solid fraction versus bed-height

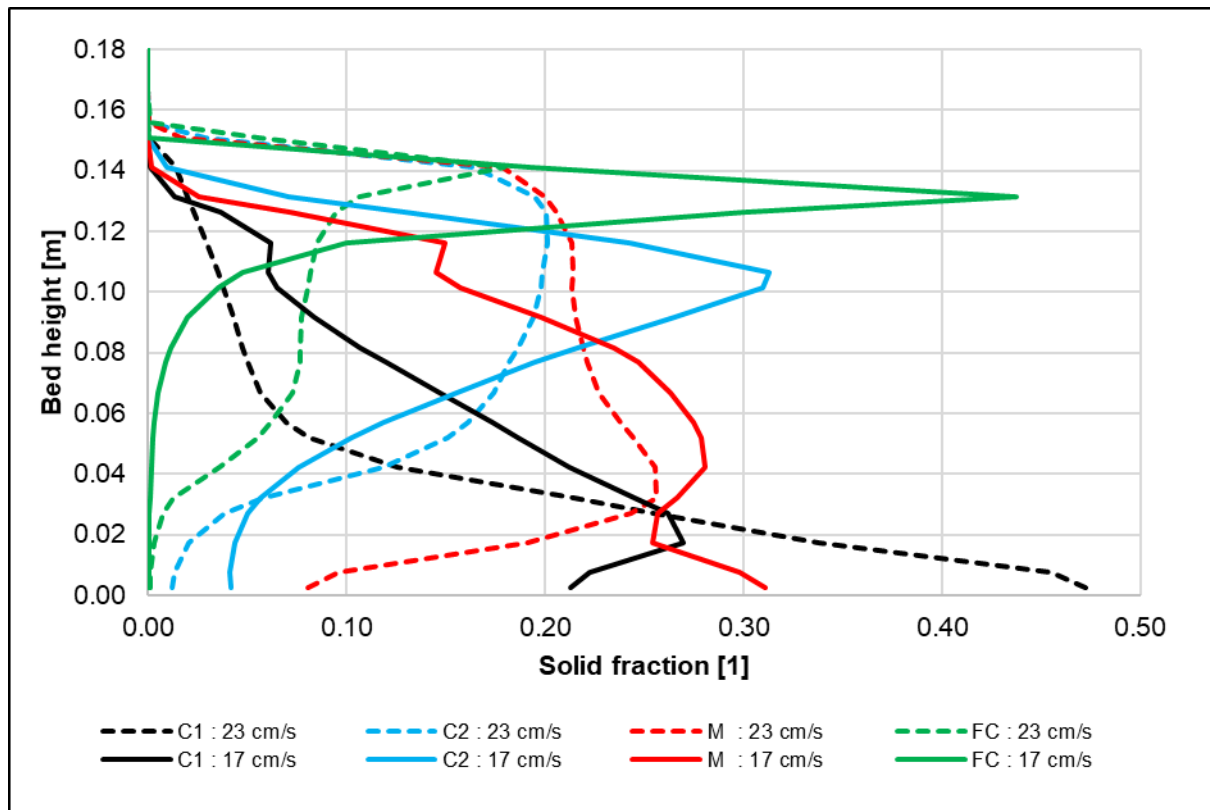


Figure 7.1-21: Particle Solid fraction versus bed-height

Summary

In this study CFD was used to gain insight into the behaviour of an ADMFB for the beneficiation of coal particles. The bed consisted of monodispersed ash, coal and DM particles. The fluidized DM particles provide a bed density that acts as a cut density in separating coal from ash on the principle of buoyancy.

In the first past part of the study the bed height and bed homogeneity were varied through a parametric study involving DM particles size, DM material density, superficial gas velocity and the ratio of DM to coal and ash. The CFD results showed how optimal segregation can be obtained from different combinations of fluidization velocity and DM properties. In a second part of the study CFD was used to investigate the properties of a DM bed when the DM is represented by a binary mixture of heavy and light particles, in this case magnetite and fine coal. By varying fluidization velocity as well as the ratio of coal to Magnetite, it was shown that a homogenous bed could be obtain at a relative low fluidization velocity but only for a low composition ratio. The experimental study conducted by Diedericks *et al.* (2020) showed addition of dense medium did not significantly improve the performance of the ADMFB. This paper addresses the occurrence by further investigating the separation performance of coal in the ADMFB in the presence of a binary mixtures of DM (fine coal and magnetite).

A final application of an ADMFB was simulated using a binary mixture of fine coal and Magnetite as the DM to separate coal and ash particles. Throughout the study the CFD model was able to demonstrate key characteristics of an ADMFB as found by various researchers in the field. An improved understanding of the role of the DM during density-based segregation of particles was obtained. The requirements of a suitable DM to enable particles segregation could also be demonstrated.

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7.2 Additional Information (Paper 4)

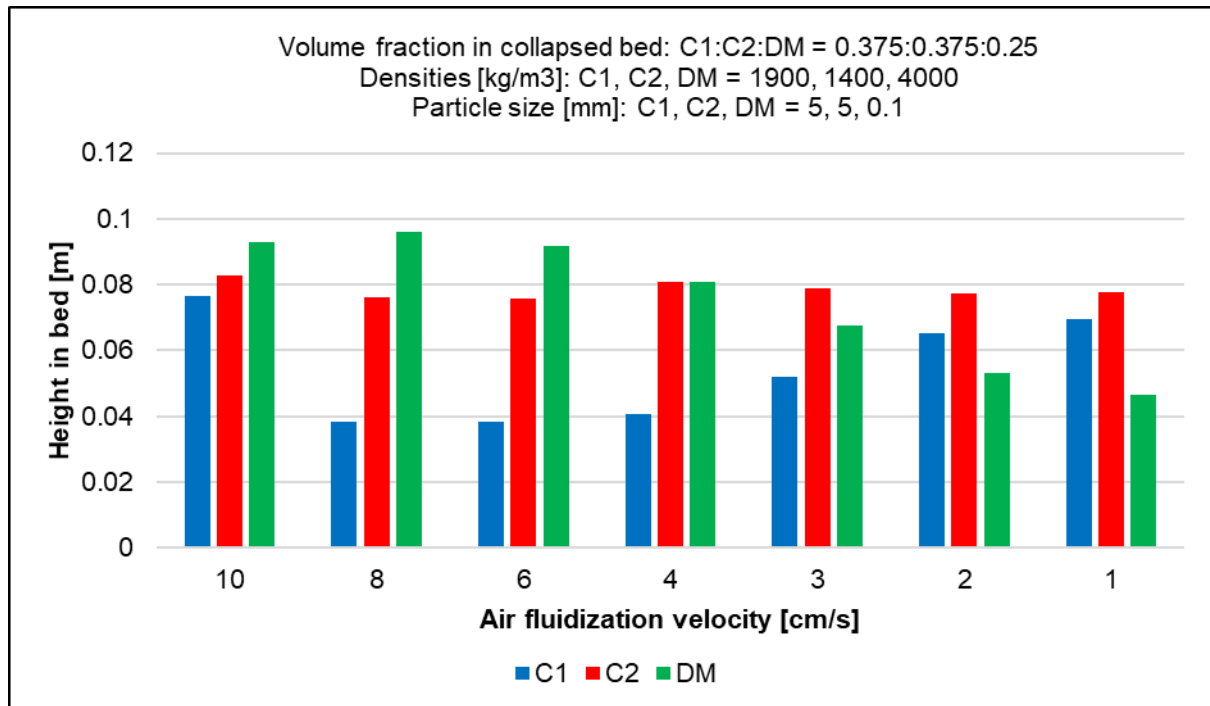


Figure 7.2-1: ADMFB CFD Case 1

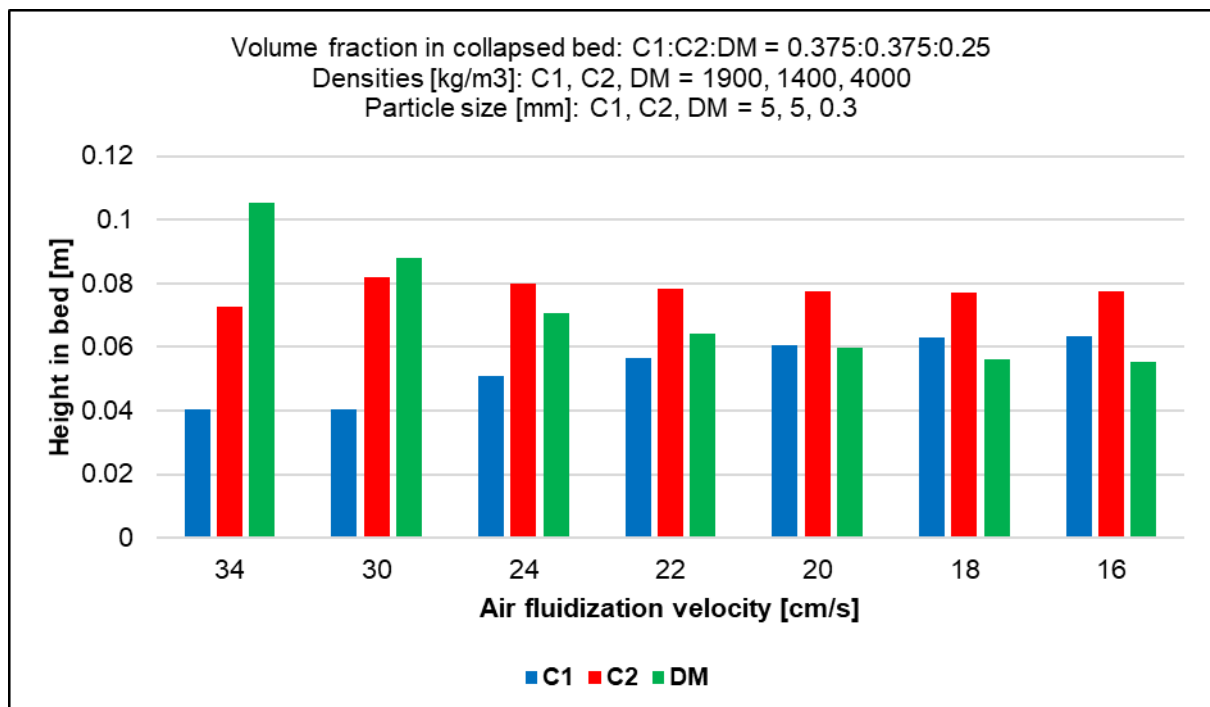


Figure 7.2-2: ADMFB CFD Case 2

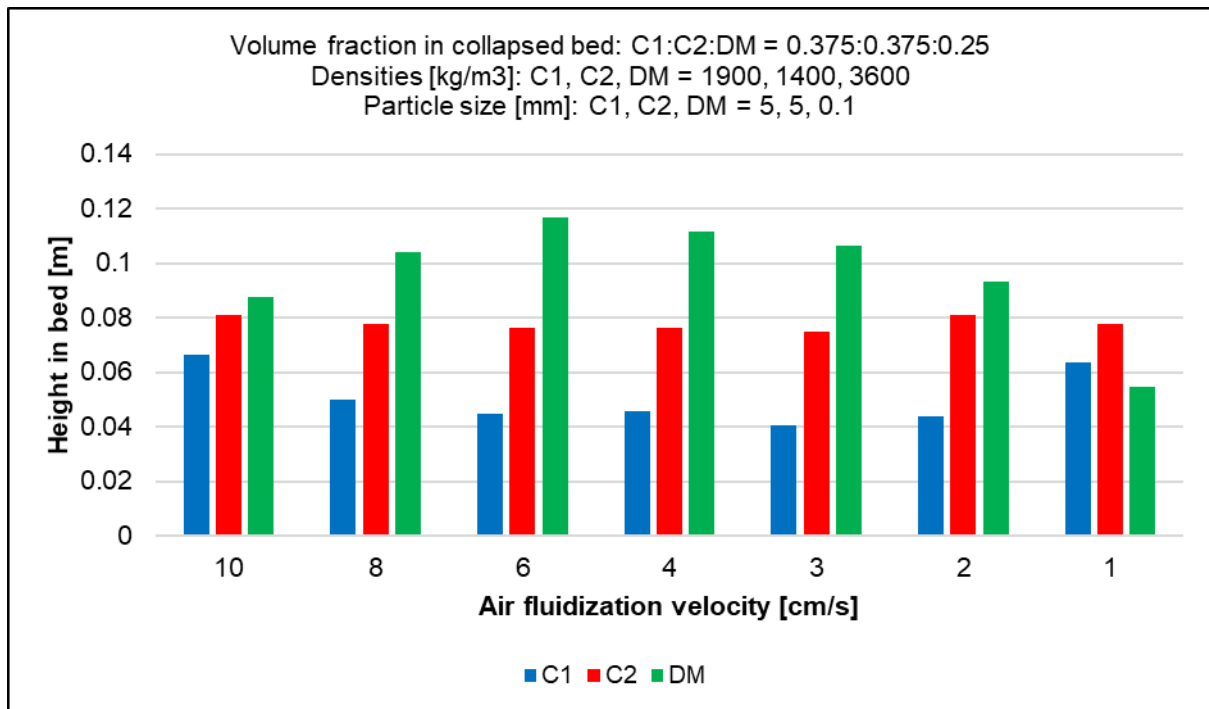


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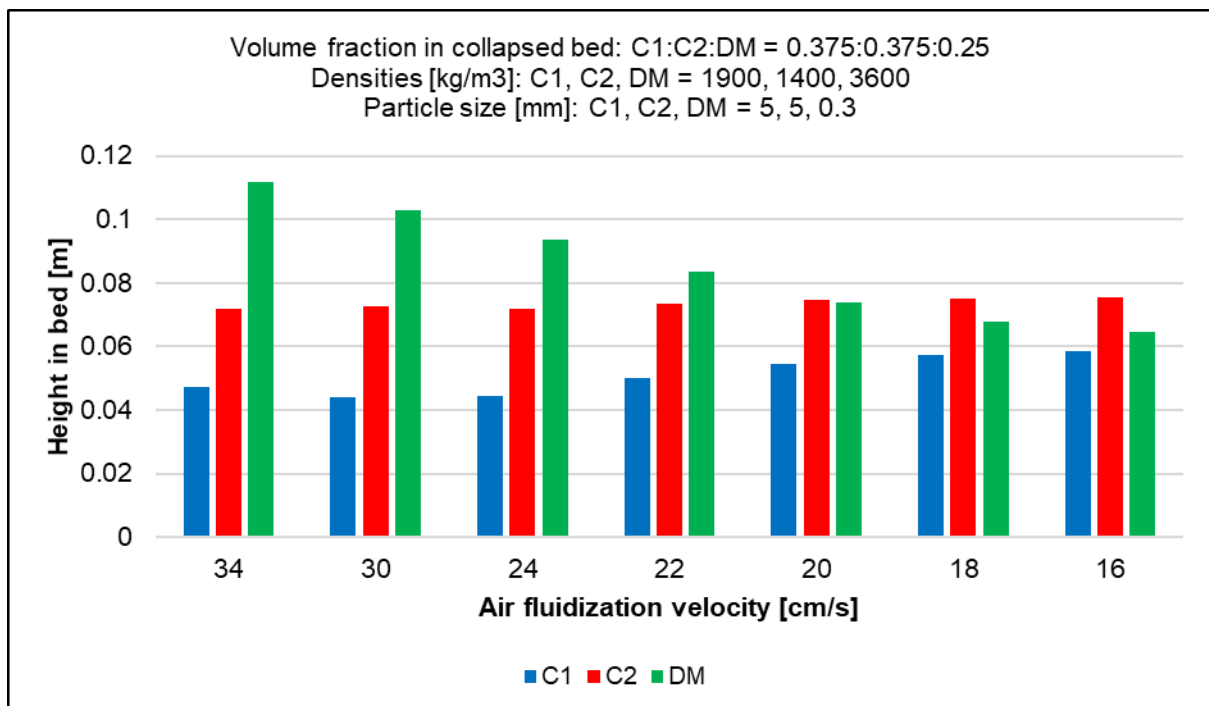


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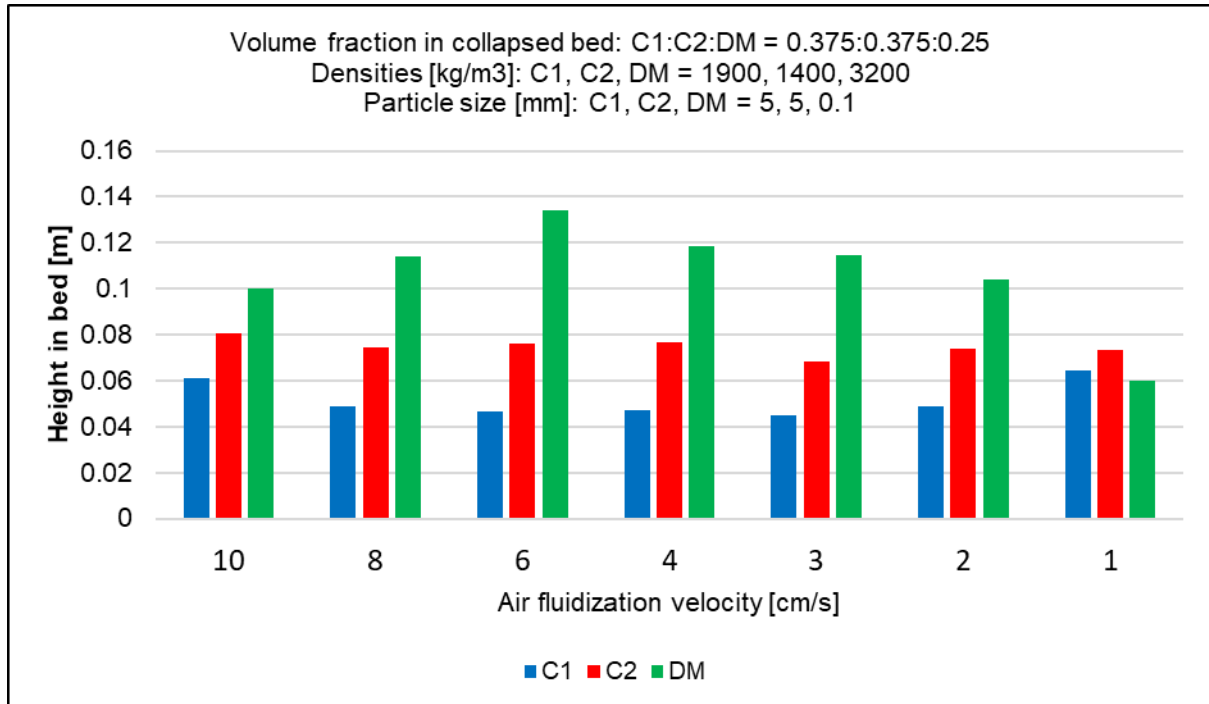


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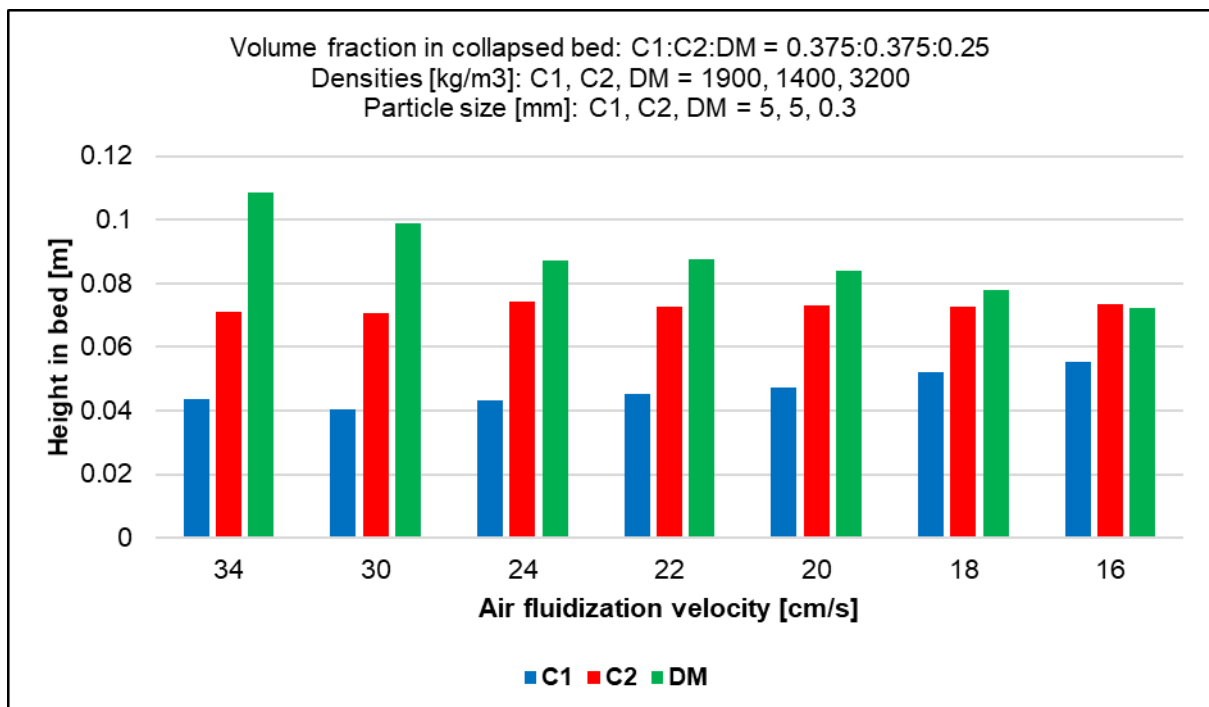


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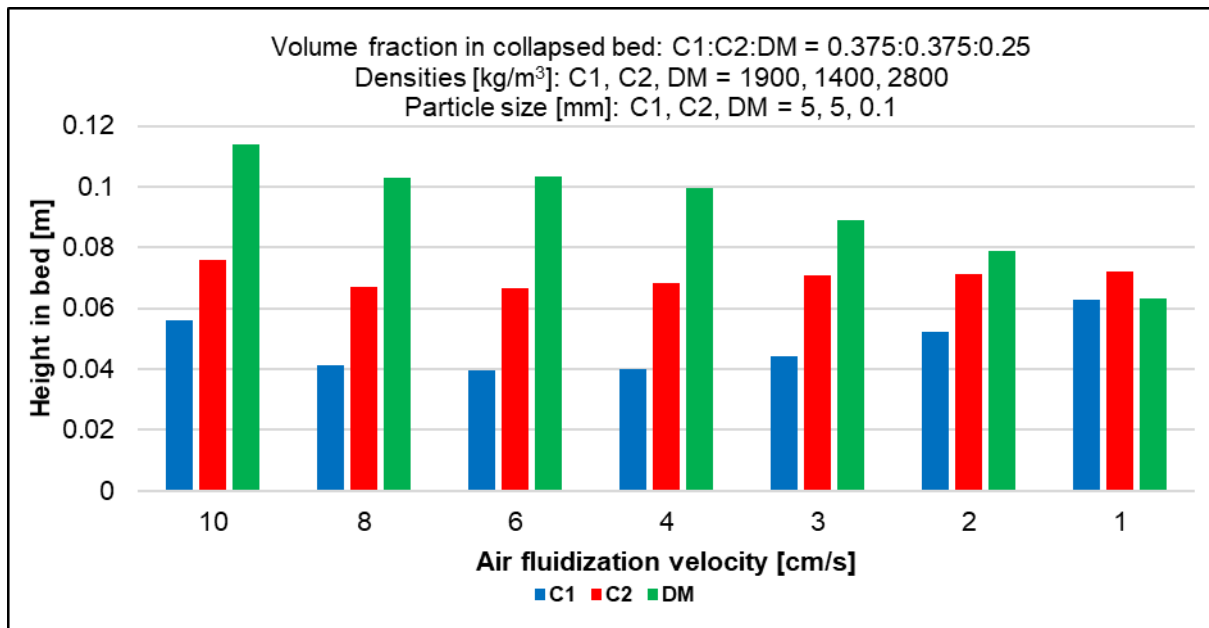


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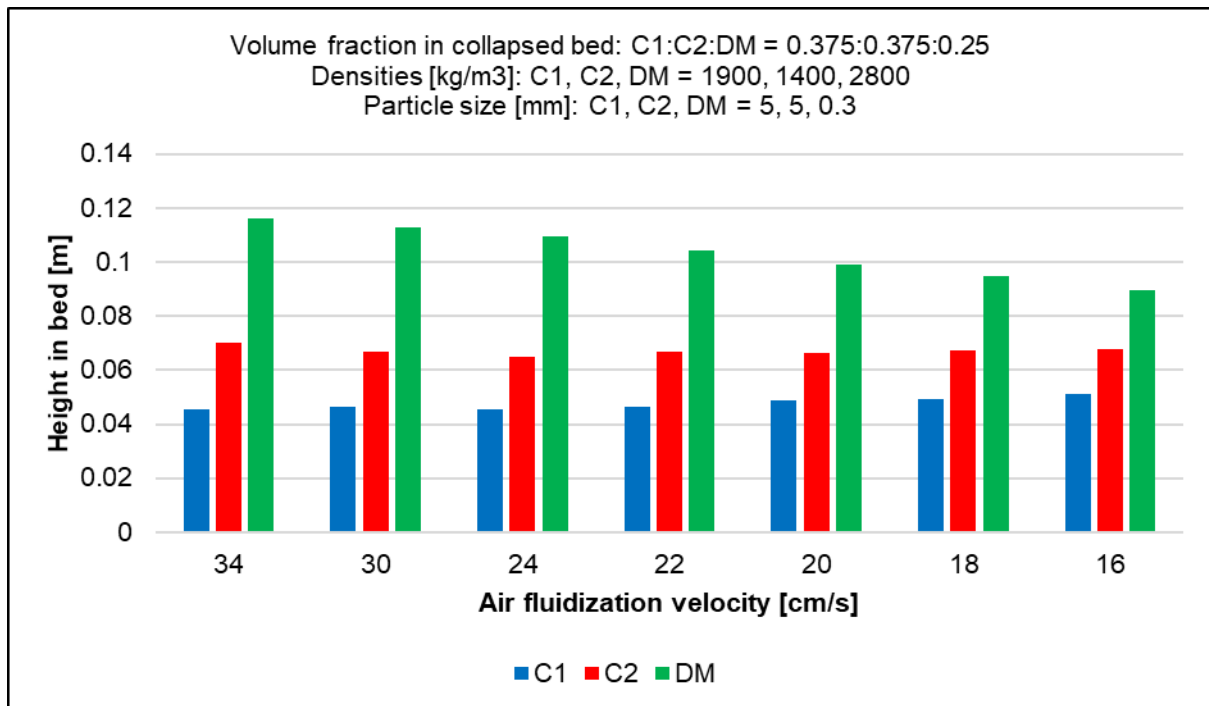


Figure 7.2-8: ADMFB CFD Case 8

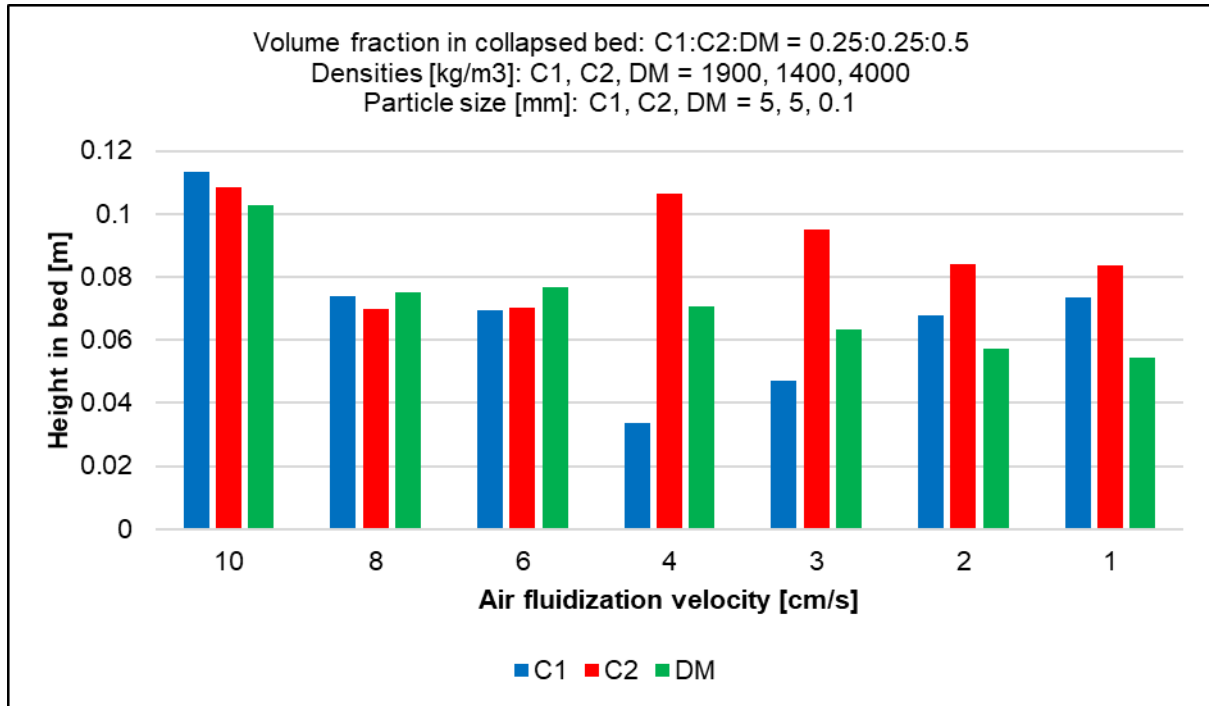


Figure 7.2-9: ADMFB CFD Case 9

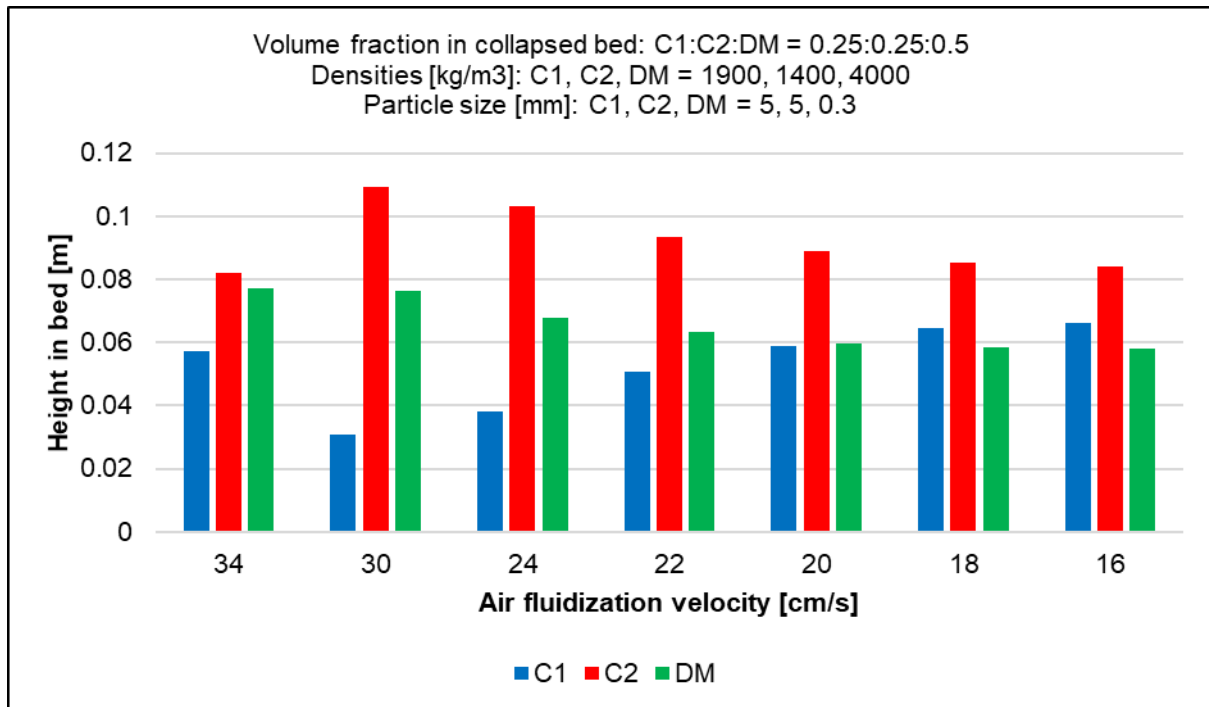


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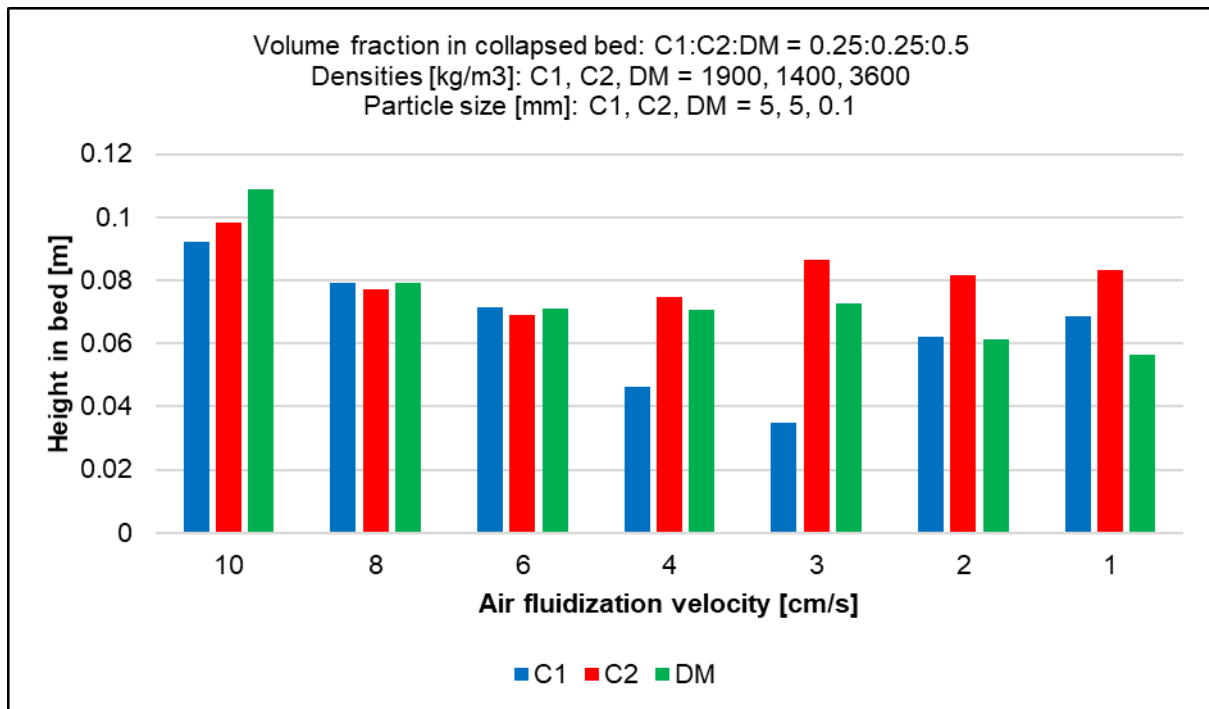


Figure 7.2-11: ADMFB CFD Case 11

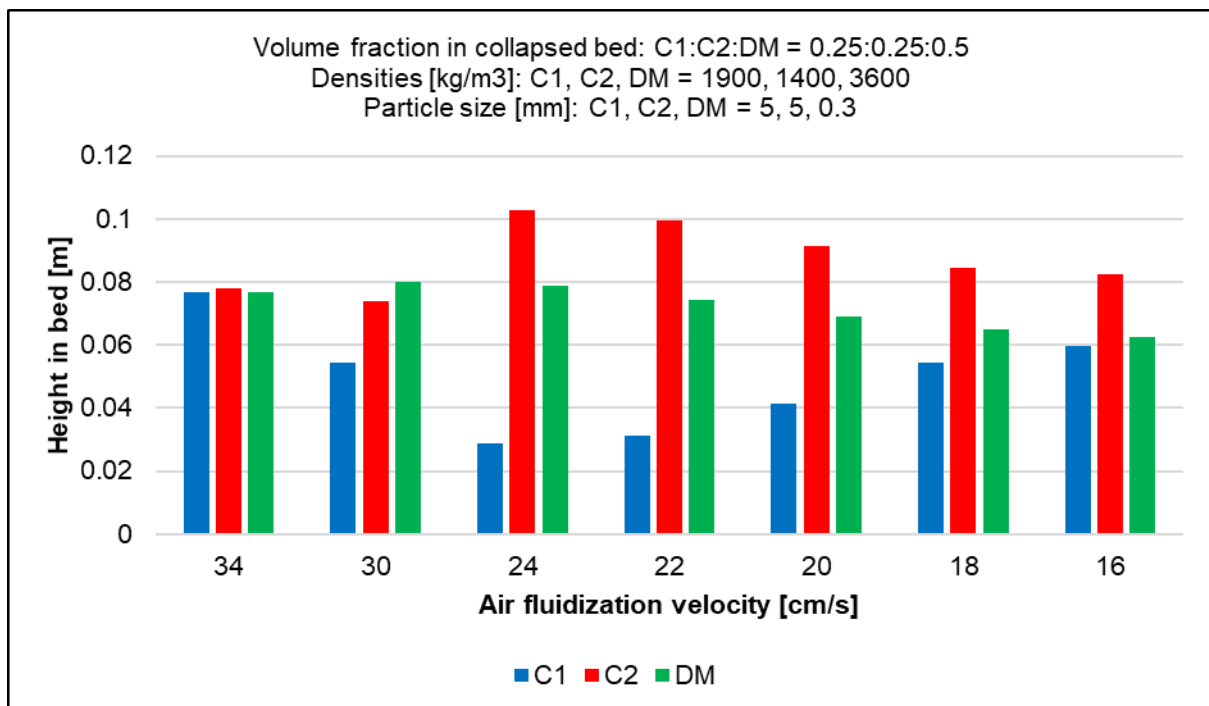


Figure 7.2-12: ADMFB CFD Case 12

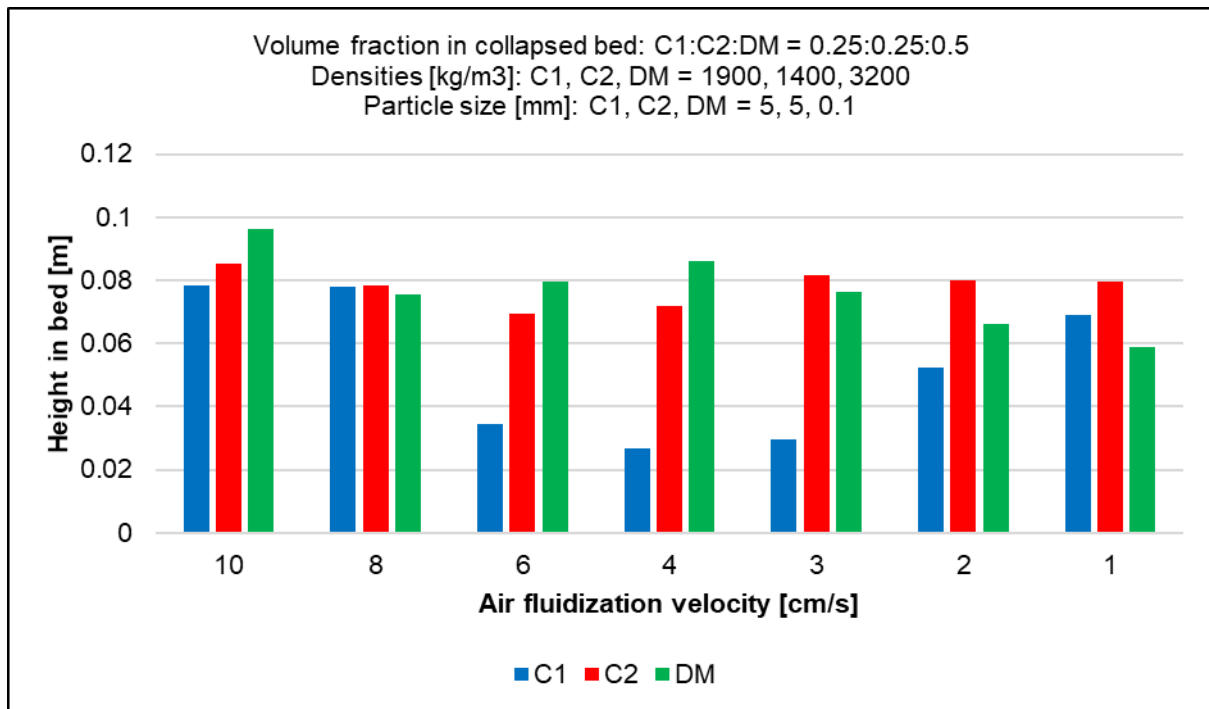


Figure 7.2-13: ADMFB CFD Case 13

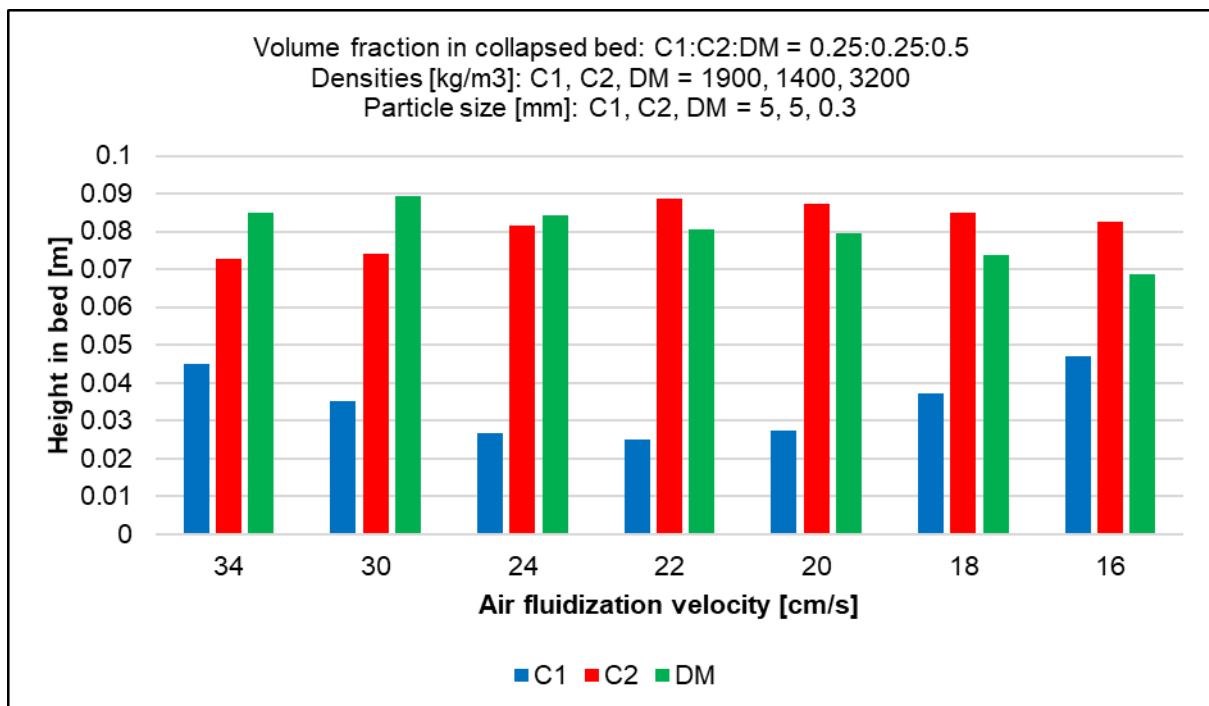


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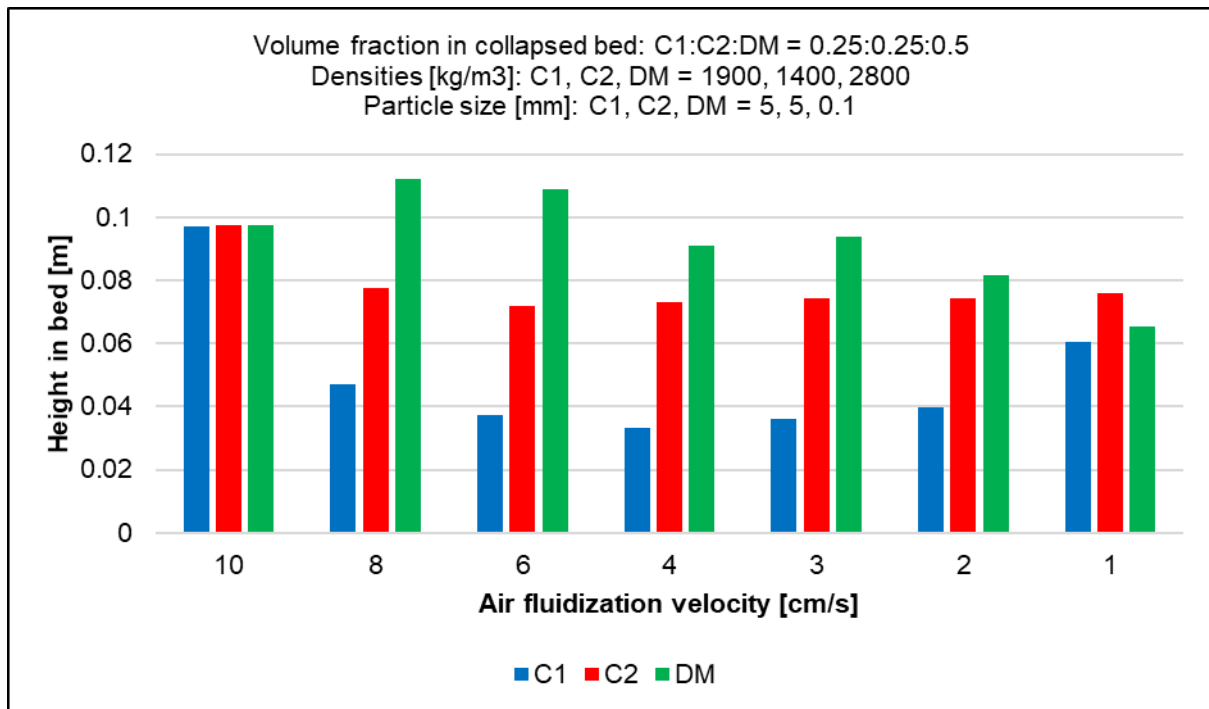


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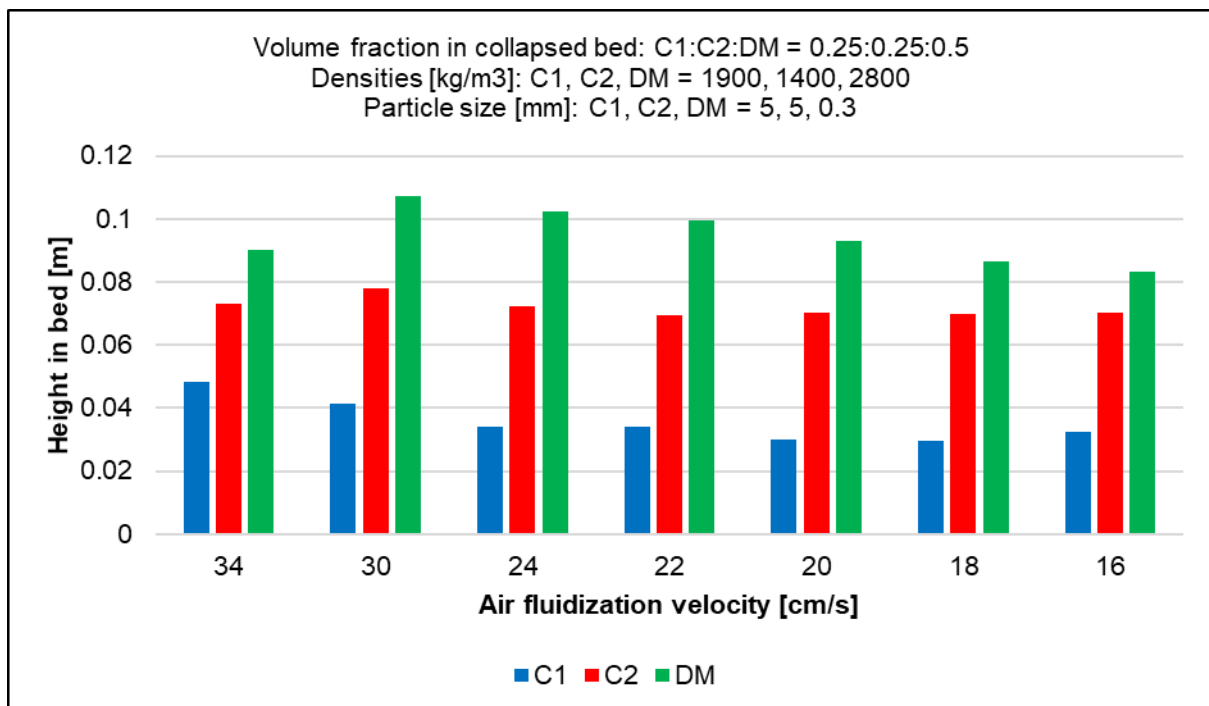


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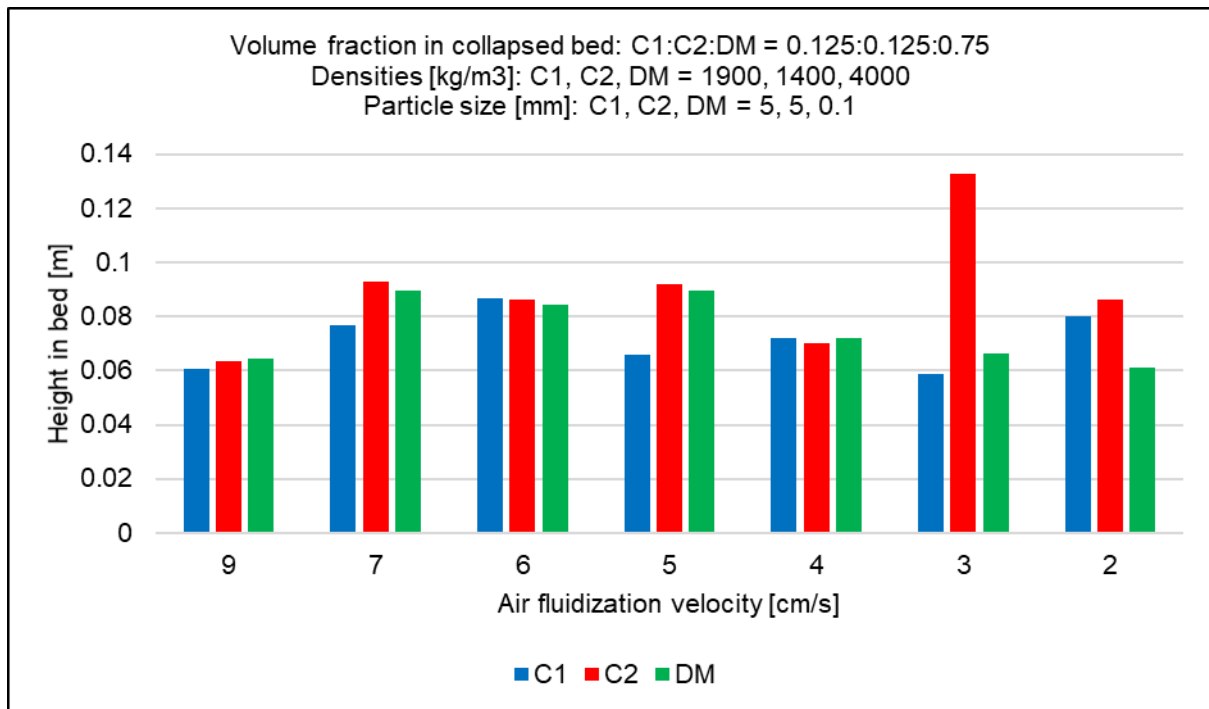


Figure 7.2-17: ADMFB CFD Case 17

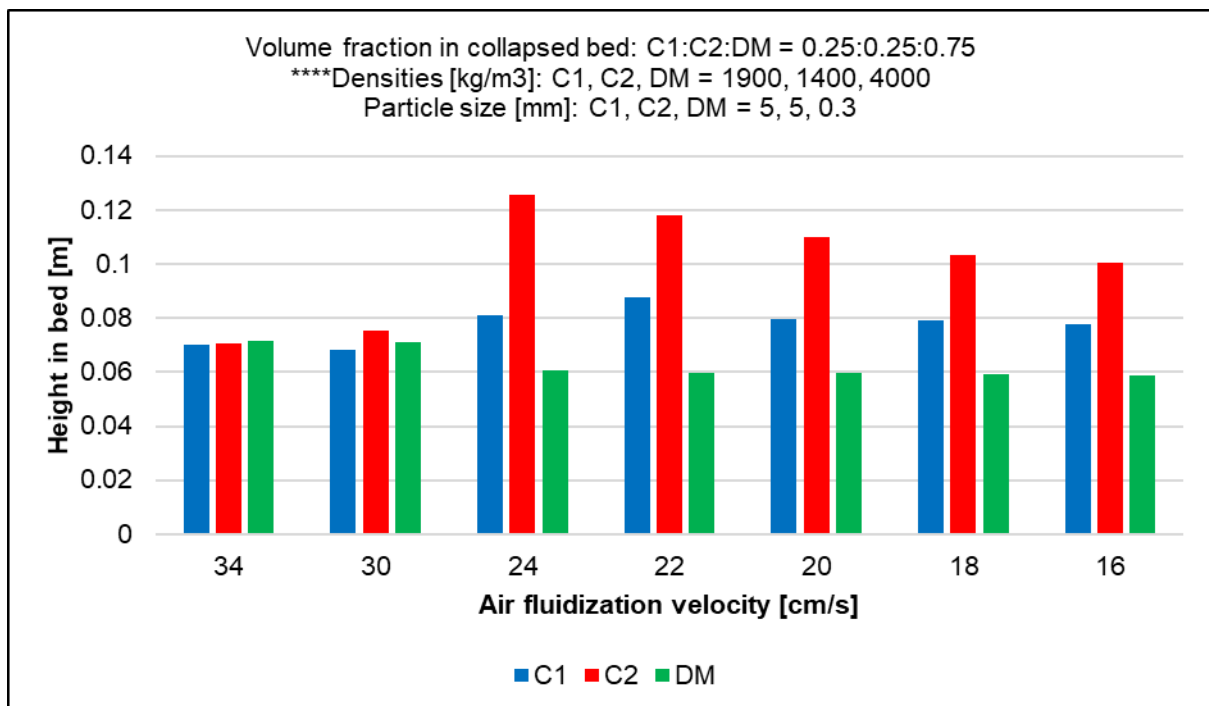


Figure 7.2-18: ADMFB CFD Case 18

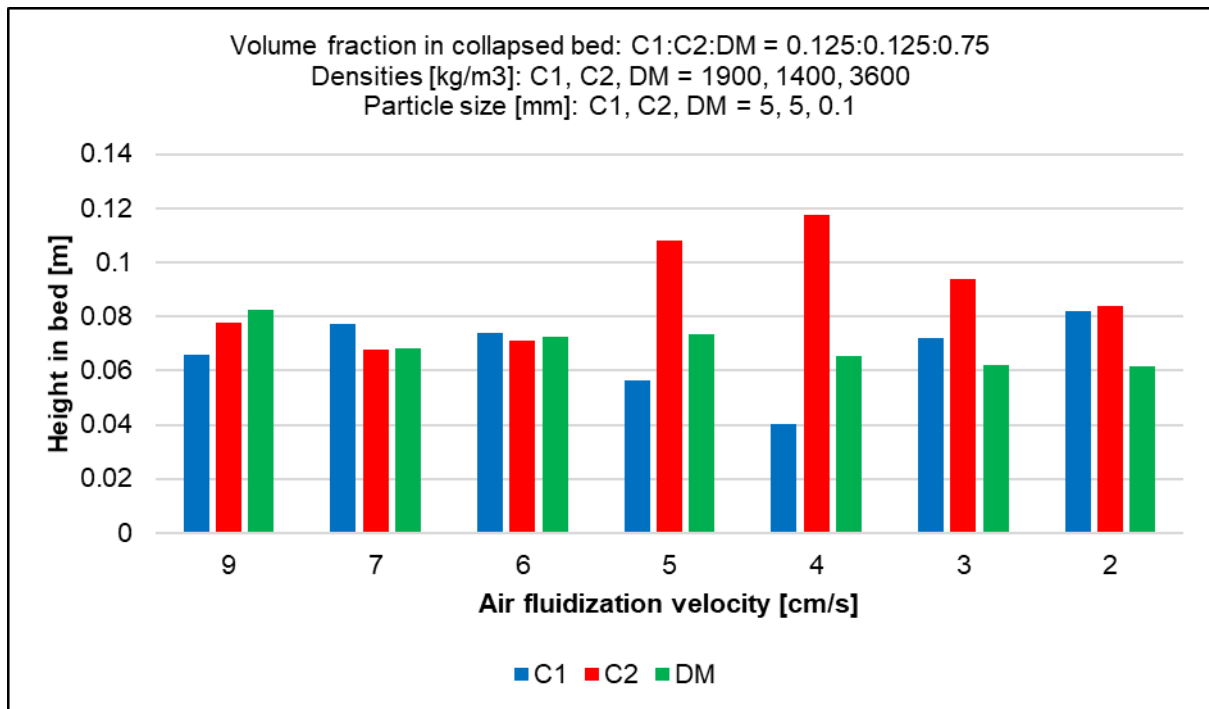


Figure 7.2-19: ADMFB CFD Case 19

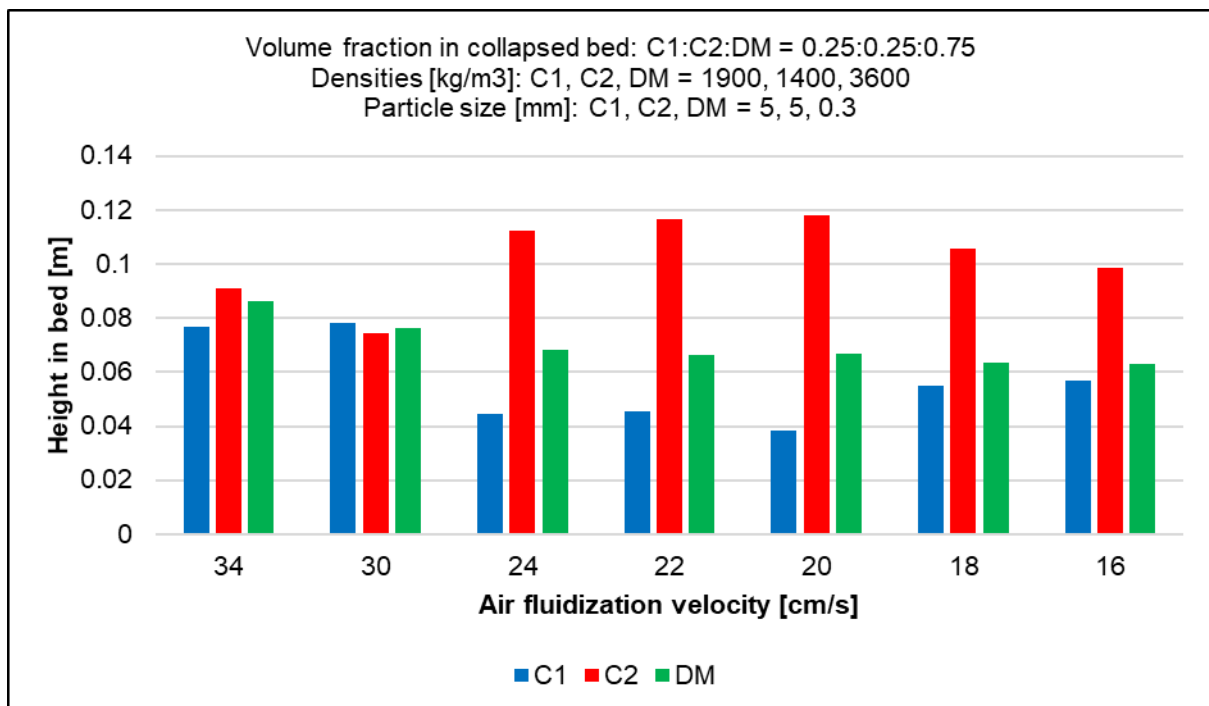


Figure 7.2-20: ADMFB CFD Case 20

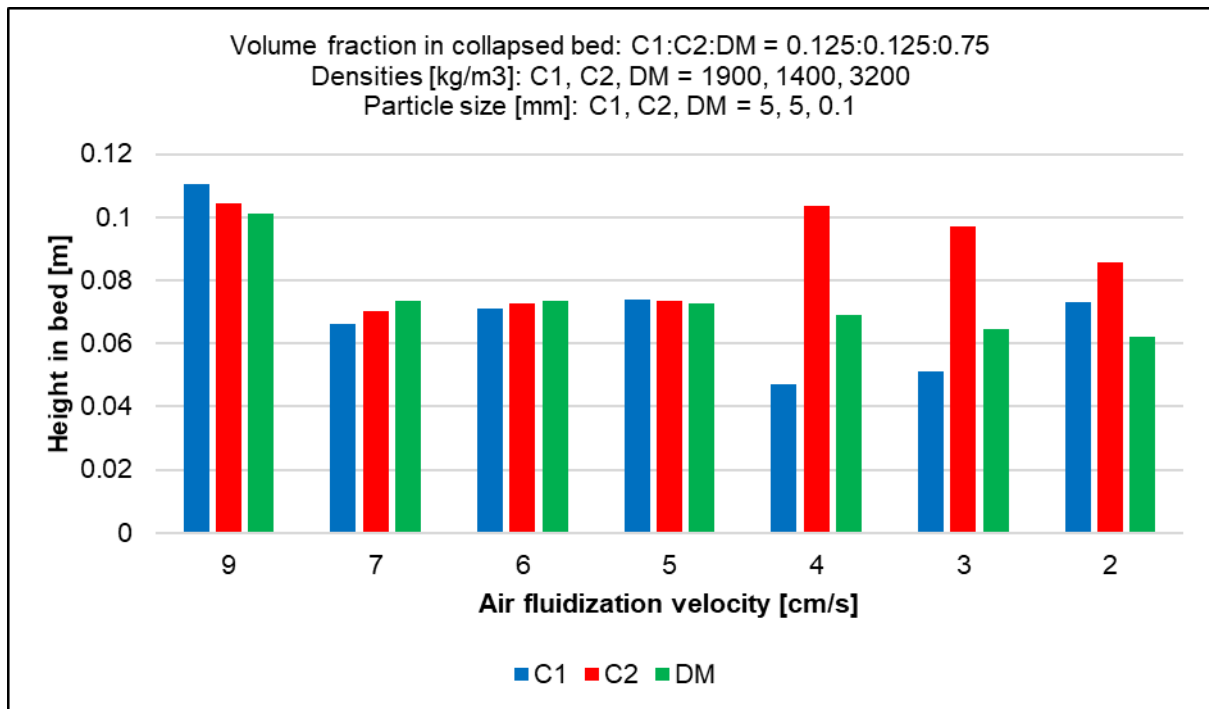


Figure 7.2-21: ADMFB CFD Case 21

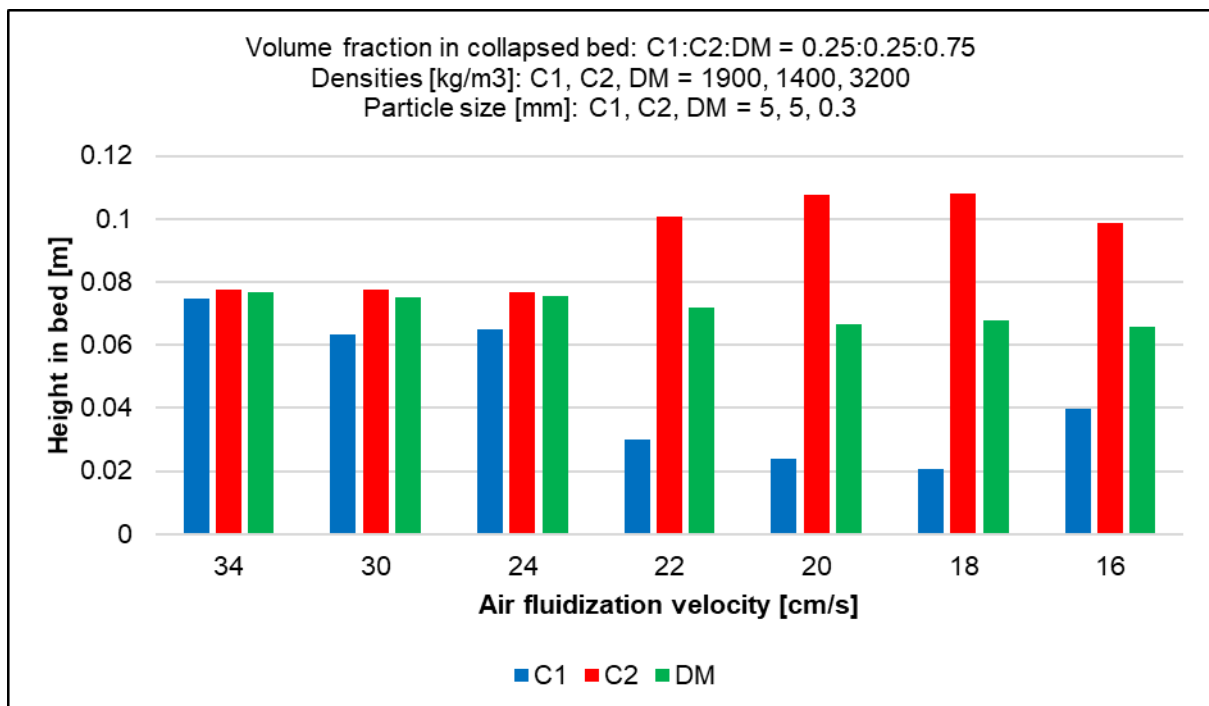


Figure 7.2-22: ADMFB CFD Case 22

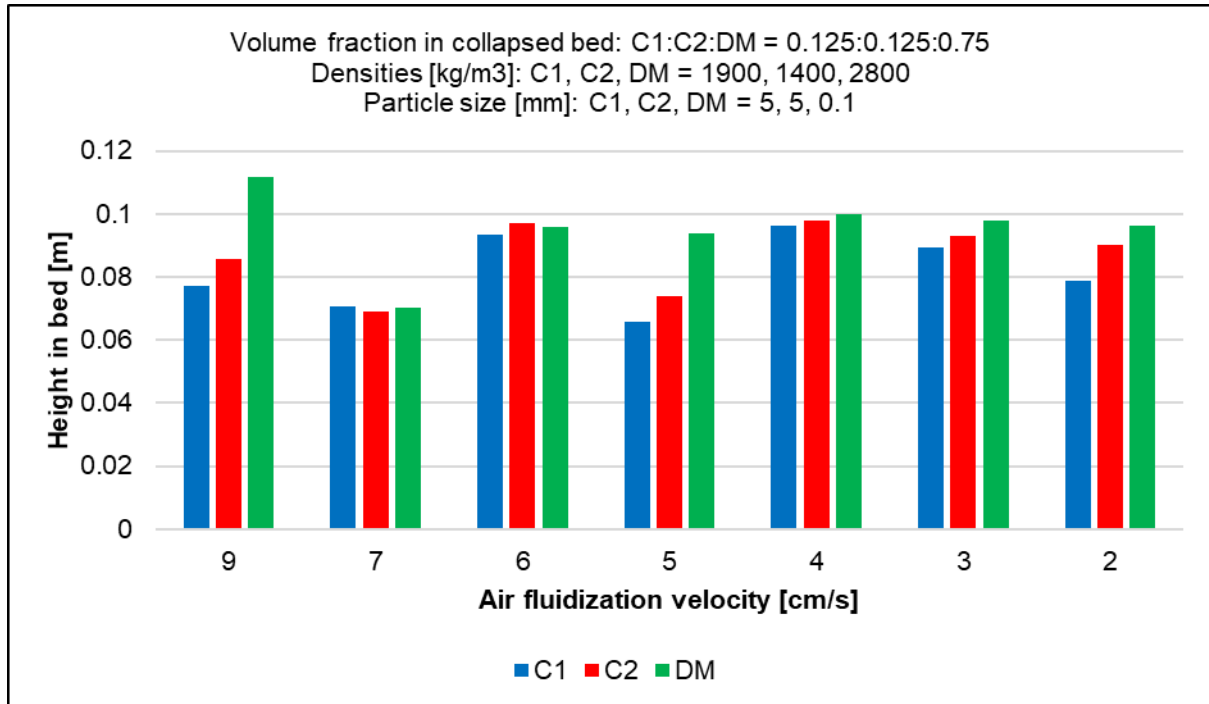


Figure 7.2-23: ADMFB CFD Case 23

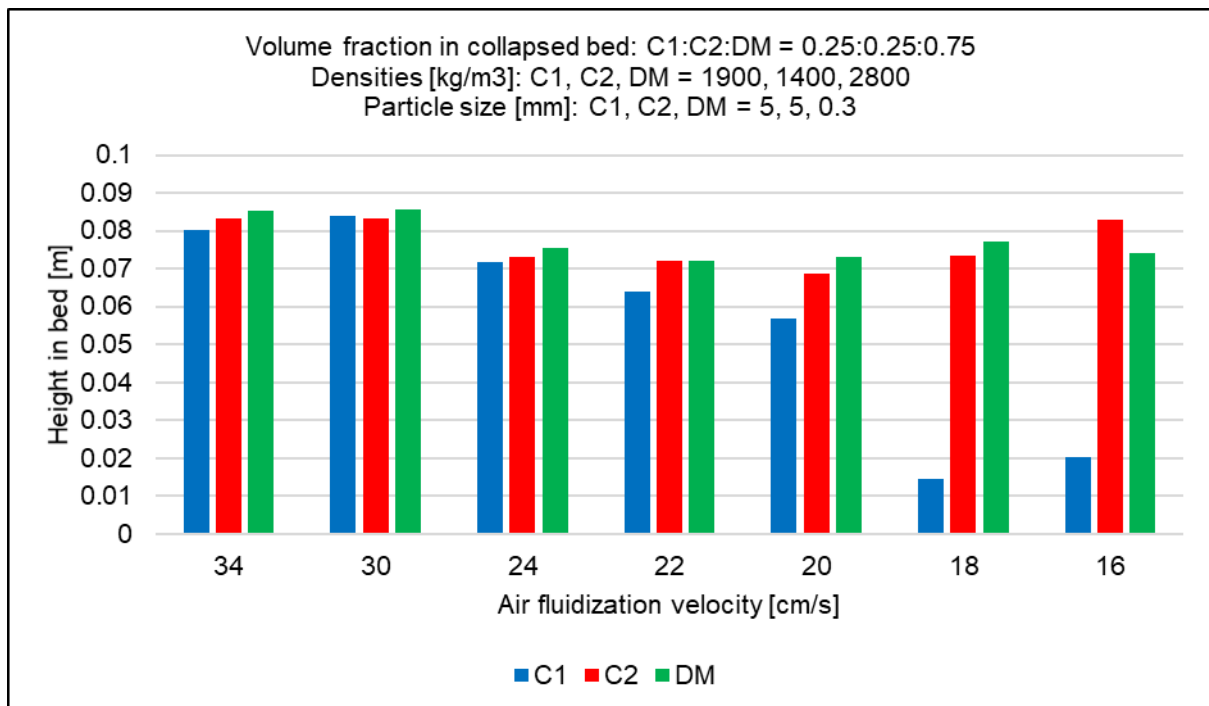


Figure 7.2-24: ADMFB CFD Case 24

CHAPTER 8

Conclusion, Recommendations and Contributions

Conclusions, recommendations and contributions on the research enclosed in this thesis is summarized in Chapter 8. The conclusions were constructed in response to the research objectives listed in Chapter 1, Section 1.2. Recommendations are also included in this chapter regarding ADMFB technology for future reference and further research on the topic.

8.1 Conclusion

The following conclusions were reached throughout the research and is summarized in response to the objectives in Chapter 1, Section 1.2:

- Coal remains an important energy-resource base worldwide as it is considered abundant, affordable and reliable (WEC, [2017-2022](#)), currently meeting up to 77% of South Africa's energy generation (DOE, 2022). The coal-based energy demand in South Africa's is unlikely to change in the next 2 decades because of the lack of appropriate alternative energy sources. The industry is currently investigating dry methods of coal beneficiation in the hopes of finding an alternative that can produce acceptable separation efficiencies. ADMFB system proves to be an effective solution to eradicate ash-forming mineral fractions from ROM coal to improve coal quality and alleviate application issues as well as reduce carbon footprint.
- For +5.6mm-13.2mm, it was found that the specific ADMFB tested was able to separate the coal successfully according to density. More dense coal particles were displaced to the bottom and the less dense coal (better quality) moved to the top layers of the bed. The best overall cumulative products obtained by discarding the bottom bed layer (cut point height of 50mm from the top of the bed) ranged between 13.0% to 17.6% (ash yield) with a corresponding yield of 82.21% to 83.93%, respectively. These results were all obtained for ADMFB experimental runs that employed no dense medium (DM 0:1) and no addition of vibration (VD). The ash and mass yield results showed that the ADMFB can produce low cumulative ash yields at a reasonably low mass production.

PSD influences the separation the ADMFB as larger particle sizes generated a higher ash rejection compared to the smaller particle size ranges. Comparable ash values were obtained for all bed layers of the PSD's investigated. A reduction in PSD had effectively led to a decrease in minimum fluidization velocity. This reduction is because of the reduced lower buoyant force required to suspend the particles.

Coal beds with magnetite yielded mediocre ash rejection results. The results of the top layers of DM 2:1 and DM 1:1 coal bed differed marginally, with somewhat lower ash values observed for the DM 0:1 coal bed. Moreover, high ash values were found in the lower section of the DM 0:1 coal bed, which was indicative of better de-stoning capabilities. Additionally, by adding magnetite to a coal bed, a significant reduction in minimum fluidization velocity was seen.

When vibration was added to the system, no drastic improvement in the results throughout the bed was observed, however an increase in the ash values in the

bottom bed layer was observed. Finally, it was concluded that the separation performance of the ADMFB with vibration (VA) was directed by the particle segregation in the two bottom layers. Vibration and increased quantities of magnetite, resulted in a densely packed bed, which hindered the movement of coal particles.

- For +2.0mm-5.6mm coal particles, low cumulative ash yields were obtained at high mass yields; however, the ash yields increased for -2mm coal. Addition of vibration and dense medium have, in some cases, improved the separation efficiency of the ADMFB.

Adequate results were obtained for larger PSD's down to and including +2.0mm-2.8mm coal particles, after which the separation performance decreased significantly. Density separation was not obvious and irregular for -2mm coal particles and maintaining a stable fluidization state proved to be challenging, especially when dense medium was added. The coal particles separated vertically along the bed height due to differences in particle and bed density, while PSD proved to have a notable impact on the degree of separation. An air fluidization velocity of between 1.1 to $1.4U_{mf}$ was shown as the best performing velocity, which yielded the maximum ash differential between the top and bottom layers of the bed for all the PSD's tested.

For +2.0mm-5.6mm coal PSD's, low cumulative ash yields were attained at high mass yields, however the ash yields increased for -2mm coal. Vibration and dense medium have enhanced the separation efficiency of the ADMFB in some cases. The -2.0mm particles experienced stronger particle-particle interactions, elevated levels of bubbling and back mixing than that of the +2.0mm particles, which provides reasoning for the poor performance of the small particle sizes. The feed ash values attained for the -2.8mm ranges were less than that obtained for the +2.8mm ranges, leading to the conclusion that most of the valuable macerals have been liberated and hence carrying less mineral matter. Nonetheless, the -2.8mm beneficiated product still adhered to the local thermal coal market demands.

- During this study, CFD was employed to improve the understanding of the dense medium (DM) selection process for best particle segregation. The bed consisted of monodispersed ash, coal and dense medium (DM) particles. The fluidized DM particles provide a bed density that acts as a cut density in separating coal from ash on the principle of buoyancy. During this study, segregation of particles in a ADMFB were varied through a parametric study involving DM particles size, DM material density, superficial gas velocity and the ratio of DM to coal and ash. The CFD results showed that segregation was successfully obtained from different combinations of

fluidization velocity and DM properties. Moreover, the properties of a DM bed were investigated where the DM is represented by a binary mixture of heavy and light particles, which in this case was a combination of magnetite and fine coal. By varying fluidization velocity as well as the coal to magnetite ratio, it was found that a homogenous bed could be obtained at low fluidization velocity for a low DM composition ratio. Lastly, an ADMFB was simulated using a binary mixture of fine coal and magnetite as the DM to separate coal from gangue particles. The simulations showed the range in fluidization velocities and DM properties that could be effective to separate high- and low-density coal particles.

8.2 Recommendations for future works

The following recommendations were deduced from the study:

- Thorough insight into the characteristics of coal samples as well as the capability to predict quality and degree of beneficiation for ADMFB utilizing a densimetric measurement systems such as RhoVol technology from De Beers can be beneficial to the utilization of ADMFB in industrial applications.
- An economic evaluation for the ADMFB process is strongly recommended. This economic evaluation should be comprehensive and include the first stage (mining the ROM coal) up to the final stage of converting energy to electricity. The evaluation should encompass considerations of waste management and related costs.
- The current setup at North-West University of Potchefstroom requires re-designing with a screw feeder for the coal and a tailing discharge section. It was found that placing the dense medium and coal particles in the ADMFB ahead of introducing air negatively affects the movement of particles located at the bottom of bed by forming channels through the particle bed. Therefore, it is recommended to consider fitting a continuous feeder mechanism to the top of the ADMFB, such as a screw feeder with adjustable feeding rate to add coal particles to an already fluidized bed medium. Addition of a tailing discharge section removing freely entered particles to a second chamber at the bottom of the bed. This will facilitate particles discharging without interfering with the process and causing operational issues.
- A combination of drying and separation in continuous operation is also proposed. Due to the air blowing through the bed and the porous nature of the coal, it is possible to reduce a large amount of moisture from the coal in a short time.
- Further investigation with modelling tools such as Neptune is required to evaluate fluidization dynamics of the ADMFB.

