

Correlating laboratory and pilot scale reflux classification of fine coal

IGT Smith
20250568

Dissertation submitted in fulfilment of the requirements for
the degree *Magister* in *Chemical Engineering* at the
Potchefstroom Campus of the North-West University

Supervisor: Prof QP Campbell

Co-Supervisor: Prof M le Roux

September 2015

Declaration

I, Izak Gerhardus Theron Smith (the undersigned), declare that the research dissertation, entitled:

Correlating laboratory and pilot scale reflux classification of fine coal

is my own unaided original work. It is herewith submitted in partial fulfilment of the requirements for the degree of Magister in Chemical Engineering to the North-West University of the Potchefstroom campus. It has not been previously submitted for any degree or examination to any other educational institution. All sources have been given recognition and have been referenced.

Signed at _____ on this _____ day of _____ 2015.

IGT Smith

Acknowledgements

Thank you:

- First of all, to our Great Heavenly Father for the support structures and people in my life, making it possible for me to study.
- To my beloved Lee-Mari, who stands beside, motivates and loves me.
- To my parents for their motivation and the opportunity to go to a university.
- To Prof Quentin Campbell and Dr Marco le Roux for their support, technical insight, and mentorship.
- To SAMMRI for the financial support during my post-graduate studies.
- To our Workshop at the School for Chemical and Minerals Engineering at the NWU for support in building the experimental unit.
- To all of the support staff:
 - Sanet Botes for always being very helpful in procuring items, and administering the financial aspects.
 - Nico Lemmer and Gideon van Rensburg for their support in the laboratories.
 - Annieta Bok and Eleanor de Koker.

Abstract

The search for efficient and economical ways to beneficiate fine coal remains an active research area. Recent developments have shown that the reflux classifier can successfully be used on Australian coals, and based on that, a number of pilot plant investigations have been done in South Africa. While pilot scale units are usually used to test the applicability of a new technology on specific coals, a need exists to gather more fundamental data at a laboratory scale in order to save manpower, costs and time. This study has aimed at introducing a way to pre-test material prior to pilot plant trials in the design chain.

The study shows that a laboratory water only reflux classifier can be used as a density fractionator, which accurately produces washability data for coal – this was also investigated by Callen *et al.* (2008). There is also a linear correlation between density cut-point and fluid velocity within the plates. Only when looking at the model proposed in Walton (2011:68), does it become clear that the relationship is indeed slightly curved. Many investigations from laboratory and pilot tests accept the linear relationship, and describe it as slightly curved due to the settling being in the intermediate settling regime (Iveson *et al.*, 2014; Galvin & Lui, 2011).

The separation procedures that produce two products – an overflow and underflow – compare well with fractionation results produced. Thus, fractionation results can generate washability data and predict batch separation operations. The laboratory reflux classifier setup is also dependent on particle size, where individual size ranges achieve e.p.m. values of 0.012 and 0.030, while the combined separation efficiency is 0.039.

It was, however, found that the respective laboratory scale reflux classifier that was designed and built was not suitable for continuous operation. The vertical fluidisation section was not high enough to enable a steady fluidised bed. This was necessary for density separation within the bed and to produce a significant pressure differential. It is also recommended to obtain a PID controller.

Keywords: reflux classification; water only density fractionation; fine coal

Opsomming

Die soeke na meer effektiewe en ekonomiese steenkoolverrykingsmetodes bly 'n aktiewe navorsingsveld. 'n Aantal loodsondersoeke op die reflux klassifiseerder is in Suid-Afrika gedoen, omdat onlangse ontwikkelings in hierdie veld suksesvol toegepas word op steenkoolverryking in Australië. Alhoewel loodsondersoeke normaalweg gedoen word om die toepasbaarheid van nuwe tegnologie te toets, bestaan daar steeds die behoefte om hierdie tegnologie op 'n meer fundamentele wyse te toets en laboratoriumdata te genereer, wat mannekrag, koste en tyd sal spaar. Hierdie studie het gepoog om materiaal in die ontwerpsfase, voor loodsondersoeke gedoen word, te toets.

Daar is bevind dat 'n laboratorium refluxklassifiseerder wel gebruik kan word as 'n digtheidsfraksioneerder, wat akkurate wasbaarheidsdata vir steenkool kan bepaal – dit is ook deur Callen *et al.* (2008) bevind. Ook is daar 'n lineêre korrelasie tussen die digtheidsnypunt en die snelheid van die fluïde tussen die plate. Daar is wel bevind dat die model wat in Walton (2011:68) vir hierdie verhouding voorgestel word effens gekrom is. Heelwat ondersoeke in beide laboratorium en loodsaanlegte aanvaar en bevind die lineêre verhouding, en beskryf dit as effens gekrom omdat die uitsakking in die oorgangsuitsakkingsregime val (Iveson *et al.*, 2014; Galvin & Lui, 2011).

Die skeidings lot prosedures wat twee produkte produseer – 'n oorvloei en ondervloei – vergelyk goed met die verkrygte fraksioneringsresultate. Dus kan wasbaarheidsdata bepaal sowel as die lot skeidingswerking voorspel word. Die laboratorium reflux klassifiseerder is ook afhanklik van die grootte van partikels, waar e.p.m. waardes vir individuele groottereekse tussen 0.012 en 0.030 was, terwyl die gekombineerde skeidingseffektiwiteit 0.039 was.

Dit was egter bevind dat die respektiewe laboratorium reflux klassifiseerder wat ontwerp en gebou was, nie geskik vir kontinue werking was nie. Die vertikale fluidiseeringsdeel was nie hoog genoeg om a stabiele gefluidiseerde bed te ontwikkel nie. Dit was noodsaaklik vir digtheids skeiding in die betrokke gedeelte asook om 'n betekenisvolle druk differensiaal oor die bed te genereer. Dit is ook 'n aanbeveling om 'n PID beheerder aan te skaf.

Sleutelwoorde: reflux klassifikasie; water-digtheidsfraksionering; fyn steenkool

Table of Contents

Declaration.....	i
Acknowledgements.....	ii
Abstract.....	iii
Opsomming.....	iv
Table of Contents.....	v
List of Figures	viii
List of Tables	xii
List of Abbreviations	xiii
List of Symbols	xiv
CHAPTER 1 - Introduction	1
1.1 Background.....	1
1.2 Motivation.....	2
1.3 Scope of Investigation	4
1.3.1 Design and commission of a laboratory-scale reflux classifier	4
1.3.2 Preliminary test work	4
1.3.3 Main experiments.....	4
1.4 Dissertation Outline	5
CHAPTER 2 - Literature Review	7
2.1 Coal.....	7
2.1.1 Origin.....	7
2.1.2 Characterisation	8
2.2 Coal Beneficiation	9
2.2.1 A historical overview	9
2.2.2 Beneficiation methods in fine coal	9
2.3 Fluid Dynamics	13
2.3.1 Flow regime.....	13
2.3.2 Pressure drop and frictional losses	15
2.4 Solid-Liquid Systems.....	16
2.4.1 Rheology.....	16

2.4.2	<i>Forces on a submerged particle</i>	17
2.4.3	<i>Terminal settling velocity</i>	19
2.4.4	<i>Inertial lift</i>	23
2.5	Reflux Classification	23
2.5.1	<i>Influence of inclined plates</i>	26
2.5.2	<i>Applications of the reflux classifier</i>	27
2.5.3	<i>The reflux classifier and coal beneficiation</i>	31
2.5.4	<i>Fundamental and empirical description</i>	33
2.5.5	<i>Related and future research</i>	36
CHAPTER 3 - Experimental		38
3.1	The RC105 at the North-West University	38
3.1.1	<i>Designing and manufacturing of the RC105</i>	40
3.1.2	<i>Process flow diagram and other equipment</i>	44
3.1.3	<i>RC105 operation</i>	49
3.2	Experimental Details and Method	52
3.2.1	<i>Material used</i>	52
3.2.2	<i>Equipment</i>	52
3.2.3	<i>The float-sink setup</i>	54
3.2.4	<i>Experiment layout</i>	55
3.2.5	<i>Sample analysis</i>	58
CHAPTER 4 - Results and Discussion		59
4.1	Fractionation Procedures	59
4.1.1	<i>Tracers</i>	59
4.1.2	<i>Tracers and Coal A</i>	62
4.1.3	<i>Coal A</i>	63
4.1.4	<i>Coal B</i>	67
4.1.5	<i>Fractionation procedure – Summary</i>	67
4.2	Coal Washability	70
4.2.1	<i>Specification and washability: Coal A</i>	70
4.2.2	<i>Specification and washability: Coal B</i>	70
4.2.3	<i>Washability analysis with RC105</i>	71
4.3	Separation Procedures	74
4.3.1	<i>Tracers</i>	74
4.3.2	<i>Coal A</i>	75
4.3.3	<i>Coal B</i>	76
4.3.4	<i>Separation test procedure – Summary</i>	80
4.4	The Influence of Particle Size	81
4.4.1	<i>Mineral content and particle weight</i>	83
4.4.2	<i>Comments on sample splitting technique</i>	85
4.5	Comparing correlations with literature	86

CHAPTER 5 - Closing Remarks	91
5.1 Conclusions.....	91
5.2 Recommendations	93
References	97
Appendices.....	101
APPENDIX A - RC105 Design Drawings.....	102
APPENDIX B - Detailed Calculations.....	106
APPENDIX C - Photos of the RC105 Rig.....	110
APPENDIX D - Partition Curves	118

List of Figures

Figure 1-1: World energy demand by the type of fuel in quadrillion Btu (Doman & Arora, 2013:11)	2
Figure 1-2: Outline of the dissertation	5
Figure 2-1: Muhanga's fine coal dense-medium cyclone beneficiation plant (De Korte, 2014)	12
Figure 2-2: Forces acting on a particle settling within a fluid	19
Figure 2-3: Illustration of a lamella settler (Salah, 2012).....	24
Figure 2-4: Reflux classifier illustration, (a) representing the definition and (b) an example of a reflux classifier design (Nguyentranlam & Galvin, 2004).....	25
Figure 2-5: Solids concentration increase illustration (Galvin & Nguyentranlam, 2002)	27
Figure 2-6: Partition curve for size classification in a reflux classifier (Doroodchi <i>et al.</i> , 2006)	28
Figure 2-7: Particles experiencing fluid superficial velocities proportional to their terminal velocities in the laminar flow regime (Galvin <i>et al.</i> , 2010a)	31
Figure 2-8: Variable descriptions for equations 2.51 to 2.53	33
Figure 3-1: Break-down of the RC105 parts	39
Figure 3-2: Manufacturing of the capsule housing the plates	40
Figure 3-3: 3D model drawing of the 3.5 mm channel capsule without the plates inserted	41
Figure 3-4: Illustrating the plug and the pipe configuration that is fastened to the female PVC union fitting	42
Figure 3-5: 3D model drawing of the fluidisation and feeding chamber	42
Figure 3-6: 3D model drawing of the overflow bucket	43
Figure 3-7: 3D model drawing of the flow distribution section.....	43
Figure 3-8: Photos of the 75 μ m wire mesh, perforated plate and flow distribution part filled with marbles	44
Figure 3-9: Conceptual process flow diagram for RC105.....	45
Figure 3-10: Illustrating reference points for calculating static head	47
Figure 3-11: The completed RC105 unit and descriptions.....	48
Figure 3-12: Batch mode operation PFD	50
Figure 3-13: Batch mode overflow configuration.....	51
Figure 3-14: Photo of the spinning riffler sample divider	53
Figure 3-15: Correct and incorrect cuts performed with the sample cutter	54
Figure 3-16: Float-Sink setup	55
Figure 4-1: Partition curve at 50.0 LPM: Fractionation (Tracers)	59
Figure 4-2: Partition curve at 75.0 LPM: Fractionation (Tracers)	60
Figure 4-3: Correlation: Density cut-point and volumetric flow rate (Tracers: water only)	61
Figure 4-4: Correlation: Density cut-point and volumetric flow rate (Tracers: Hindered settling)	62
Figure 4-5: The amount of coal, average density and cumulative %mass in each flow fraction (Fractionation: Coal A).....	63

Figure 4-6: Correlation: Average density for cumulative tracer and coal A data	64
Figure 4-7: Correlation: fractional density averages for tracers in water only and hindered settling system, and for coal A.....	65
Figure 4-8: Comparing the actual density cut-point and the average density (Fractionation: Tracers – hindered settling)	66
Figure 4-9: The amount of coal, average density and cumulative %mass and %ash in each flow fraction (Fractionation: Coal B)	67
Figure 4-10: Correlation: Fractional average density and volumetric flow rate (Fractionation: Coal)	68
Figure 4-11: Summary of the correlations obtained during fractionation tests (fractional average density)	69
Figure 4-12: Densimetric and cumulative ash float and ash sink curves (Floats-Sinks: Coal A)	70
Figure 4-13: Densimetric curve and calorific value (Floats-Sinks: Coal A)	71
Figure 4-14: Densimetric and cumulative ash float and ash sink curves (Floats-Sinks: Coal B).....	71
Figure 4-15 Densimetric curve and calorific value (Floats-Sinks: Coal B)	72
Figure 4-16: Densimetric and cumulative ash float and ash sink curves (RC105 fractionation: Coal B)	72
Figure 4-17: Comparing the densimetric curve obtained through float-sink analysis and RC105 fractionation: Coal B.....	73
Figure 4-18: Comparing ash curves obtained through float-sink analysis and RC105 fractionation: Coal B.....	73
Figure 4-19: Obtained density cut-points compared to the fluidisation rate and D_{50} correlation (Tracers)	75
Figure 4-20: Partition Curve of the fixed flow rate run at 40 L/min: Coal A.....	75
Figure 4-21: Partition curves for the three fixed flow rate runs: Separation runs (Coal B)	76
Figure 4-22: Cumulative average density for coal A and B, and tracer data	77
Figure 4-23: Comparing the predicted and obtained overflow average density: Separation runs (Coal B)	78
Figure 4-24: Comparing the predicted and obtained yield: Separation runs (Coal B)	78
Figure 4-25: Comparing Tromp curves of the repeat and high $\rho_{B,i}$ runs with the original separation run at 32.5LPM: Coal B	79
Figure 4-26: Correlating density cut-point to initial bed density in batch mode: Coal B.....	79
Figure 4-27: Correlating e.p.m. to initial bed density in batch mode: Coal B.....	80
Figure 4-28: Particle size distribution: Coal B (Feed)	81
Figure 4-29: Float-Sink data for each of the four PSD ranges: Coal B (Feed)	82
Figure 4-30: Comparing the Tromp curves for the three particles size distributions: Original 32.5 LPM run (Coal B).....	83
Figure 4-31: %ash of the particular size range in each float-sink fraction – the PSD is defined by bottom size: Original run at 32.5 LPM (Coal B).....	84
Figure 4-32: Comparing PSD's of sample split products from different sample split instances	85
Figure 4-33: Averaged correlation: Density cut-point and volumetric flow rate.....	87
Figure 4-34: Comparing the actual density cut-points obtained to the new correlation	88
Figure 4-35: Comparing correlations to Walton (2011): $z = 0.0035$ m	88
Figure 4-36: Comparing correlations to Walton (2011): $z = 0.0038$ m	89

Figure 4-37: Comparing correlations to Walton (2011): $z = 0.0032 \text{ m}$	89
Figure 4-38 Fitting Walton's (2011:68) equation to fractionation data	90
Figure 5-1: Water clarifier concept.....	94
Figure 5-2: Water clarifier concept: side view	95
Figure A-1: Detailed drawing of the plates supporting PVC side of the plates section	102
Figure A-2: Detailed drawing of the 3.5 mm channel capsule before cutting to 70°	103
Figure A-3: Detailed drawing of the plate section cut to 1 m and 70° to the horizon.....	103
Figure A-4: Detailed drawing of the fluidisation and feeding port section	104
Figure A-5: Detailed drawing of the overflow bucket.....	105
Figure A-6: Detailed drawing of the flow distribution section	105
Figure C-1: Photo – Entire unit	110
Figure C-2: Photo – Rotameters and piping.....	111
Figure C-3: Photo – Piping	112
Figure C-4: Photo – Inside of fluidisation section	112
Figure C-5: Photo – Fluidisation section, discard and manometer	113
Figure C-6: Photo – Coal capture in semi-batch mode (High quality coal)	113
Figure C-7: Photo – Overflow compartment with plates protruding the plates section	114
Figure C-8: Photo – Gravity feeding of unit during semi-batch mode	115
Figure C-9: Photo – Initial stage of valve and pipe configuration of semi-batch feeding system.....	116
Figure C-10: Photo – Coal slurry tank with agitator motor and shaft.....	117
Figure D-1: Partition curve for water only tracer run at 42.5 LPM (Fractionation)	118
Figure D-2: Partition curve for water only tracer run at 45.0 LPM (Fractionation)	118
Figure D-3: Partition curve for water only tracer run at 47.5 LPM (Fractionation)	118
Figure D-4: Partition curve for water only tracer run at 50.0 LPM (Fractionation)	119
Figure D-5: Partition curve for water only tracer run at 52.5 LPM (Fractionation)	119
Figure D-6: Partition curve for water only tracer run at 55.0 LPM (Fractionation)	119
Figure D-7: Partition curve for water only tracer run at 57.5 LPM (Fractionation)	120
Figure D-8: Partition curve for water only tracer run at 60.0 LPM (Fractionation)	120
Figure D-9: Partition curve for water only tracer run at 62.5 LPM (Fractionation)	120
Figure D-10: Partition curve for water only tracer run at 65.0 LPM (Fractionation)	121
Figure D-11: Partition curve for water only tracer run at 70.0 LPM (Fractionation)	121
Figure D-12: Partition curve for water only tracer run at 75.0 LPM (Fractionation)	121
Figure D-13: Partition curve for water only tracer run at 80.0 LPM (Fractionation)	122
Figure D-14: Partition curve for water only tracer run at 85.0 LPM (Fractionation)	122
Figure D-15: Partition curve for coal and tracer run at 30.0 LPM (Fractionation).....	122
Figure D-16: Partition curve for coal and tracer run at 32.5 LPM (Fractionation).....	123
Figure D-17: Partition curve for coal and tracer run at 35.0 LPM (Fractionation).....	123

Figure D-18: Partition curve for coal and tracer run at 37.5 LPM (Fractionation).....	123
Figure D-19: Partition curve for coal and tracer run at 40.0 LPM (Fractionation).....	124
Figure D-20: Partition curve for coal and tracer run at 45.0 LPM (Fractionation).....	124
Figure D-21: Partition curve for coal and tracer run at 50.0 LPM (Fractionation).....	124
Figure D-22: Partition curve for coal and tracer run at 55.0 LPM (Fractionation).....	125
Figure D-23: Partition curve for coal and tracer run at 60.0 LPM (Fractionation).....	125

List of Tables

Table 1-1: Listing the extent of particle diameter influencing settling velocity influences (Galvin <i>et al.</i> , 2010c) ..	2
Table 2-1: Performance data on a two-stage cleaner spiral circuit (Luttrell <i>et al.</i> , 2007:81)	13
Table 2-2: Parameter <i>K</i> relation to the channel spacing.....	16
Table 3-1: Experimental run break-down	56
Table 4-1: Summary of the cut-point and e.p.m. values (Fractionation: Tracers – water only).....	61
Table 4-2: Summary of the cut-point and e.p.m. values (Fractionation: Tracers – hindered settling).....	62
Table 4-3: Summary of the material and conditions during the fractionation test work.....	68
Table 4-4: Summary of predicted and actual cut-densities (tracer fractionation)	74
Table 4-5: Cut-point and efficiency data: (Separation Runs: Coal B)	76
Table 4-6: Comparing actual product average density and yield to predicted data (Separation Runs: Coal B) ...	77
Table 4-7: Summarising the density cut-point and e.p.m values of each particle size range's Tromp curve:	
Original run at 32.5 LPM (Coal B).....	82

List of Abbreviations

<u>Abbreviation</u>	<u>Description</u>
DR	Discrete run
E.p.m.	Mean probable error
PFD	Process flow diagram
PSD	Particle size distribution
RC	Reflux classifier
RC150	The laboratory RC with an inner cross-sectional area: $0.150 \times 0.150 \text{ m}$
RC300	Identified pilot RC with an inner cross-sectional area: $0.300 \times 0.300 \text{ m}$
RC600	Miscellaneous RC with an inner cross-sectional area: $0.600 \times 0.600 \text{ m}$
ROM	Run-of-mine
SR	Separation run

List of Symbols

<u>Symbol</u>	<u>Description</u>	<u>Unit</u>
A	Area	m^2
a	RC channel width (plate and channel dimension)	m
B	RC channel depth (into page)	m
b	RC channel height (plate and channel dimension)	m
C_D	Drag coefficient	—
$C_{D,S}$	Reduced drag coefficient (suspension of particles)	—
$C_{D,w}$	Reduced drag coefficient (used in accordance with f_w)	—
D	Pipe diameter	m
D_E	Equivalent diameter of conduit	m
D_H	Hydraulic diameter of conduit	m
d_p	Particle size in terms of diameter	m
E	RC throughput advantage	—
F	Force	$kg.m.s^{-2}$
F_B	Buoyancy force	$kg.m.s^{-2}$
F_D	Drag force	$kg.m.s^{-2}$
F_G	Gravitational force	$kg.m.s^{-2}$
f	Friction factor	—
f_f	Fanning friction factor	—
f_w	Wall factor (influencing u_t)	—
g	Gravitational acceleration	$m.s^{-2}$
h	Height	—
K	Parameter (related to rectangular channel dimensions)	—
K_{Ne}	Parameter (u_t for non-spherical particles: Newtonian Regime)	—
K_{St}	Parameter (u_t for non-spherical particles: Stokes' Regime)	—
L	Conduit length	m
L_{ch}	RC channel length	m
n	Parameter (Richardson and Zaki exponent)	—
n_{pl}	Number of plates in RC	—
P	Pressure	Pa
ΔP	Pressure drop	Pa
ΔP_f	Pressure drop due to friction	Pa
Q	Volumetric flow rate of fluid	$L.min^{-1}$
Q_w	Volumetric flow rate of fluidisation water	$L.min^{-1}$
Re	Reynolds number within conduit flow	—
Re_c	Critical Reynolds number	—
Re_p	Particle Reynolds number	—
Re_s	Particle Reynolds number for suspensions	—
Re_{sh}	Shear Reynolds number	—

<u>Symbol</u>	<u>Description</u>	<u>Unit</u>
Re_t	Reynolds number at particle terminal settling velocity (u_t)	—
t	Time	s
U	Fluid velocity within conduit	$m \cdot s^{-1}$
U_R	Fluid velocity relative to particle	$m \cdot s^{-1}$
u	Particle velocity	$m \cdot s^{-1}$
u_p	Fluid velocity experienced by the particle	$m \cdot s^{-1}$
u_s	Sedimentation velocity of suspension	$m \cdot s^{-1}$
$u_{s,st}$	Stokes' sedimentation velocity	$m \cdot s^{-1}$
u_t	Particle terminal settling velocity	$m \cdot s^{-1}$
$u_{t,c}$	Particle critical terminal settling velocity	$m \cdot s^{-1}$
V_p	Particle volume	m^3
W	RC fluidisation bed section width	m
w	RC horizontal channel width ($z/\sin \theta$)	m
$y_{V,sl}$	Solids fraction in slurry by volume	—
z	Normal plate spacing / channel width	m
z	Vertical direction - vector notation (e_z)	—

Greek Letters

<u>Symbol</u>	<u>Description</u>	<u>Unit</u>
γ	Shear rate of the particle	s^{-1}
ϵ	Wall roughness	m
η	Segregation efficiency in RC	—
θ	Channel angle to the horizontal plane	—
μ	Fluid viscosity	$kg \cdot m^{-1} \cdot s^{-1}$
μ_w	Viscosity of water at 25°C	$kg \cdot m^{-1} \cdot s^{-1}$
ρ_F	Fluid density	$kg \cdot m^{-3}$
ρ_p	Particle density	$kg \cdot m^{-3}$
ρ_w	Density of water	$kg \cdot m^{-3}$
ρ_{50}	Relative density at 0.50 partition number (Tromp curves)	$kg \cdot m^{-3}$
ϕ	Sphericity	—

CHAPTER 1 - Introduction

Coal plays a major role in the energy, industrial and consumer sectors of South Africa. Industrially coal is used extensively – ranging from coking coal, in metallurgical processes, to the production of vast amounts of chemicals at Sasol in coal to liquid processes. Of South Africa's electricity 90% is generated from coal-fired power plants and 27% of South Africa's coal was exported as a commodity to foreign countries in 2009. The latter two uses make coal one of the main economic drivers of the country (Eberhard, 2011:2).

Roughly 250 million tons of coal are mined in South Africa each year (IEA, 2010:III.5). Of this run-of-mine coal 15% (37.5 million tons) can be classified as fines ($-500\mu\text{m}$) (England, 2002:179), and 10 million tons are discarded each year, because the coal is too fine ($-150\mu\text{m}$) (Reddick *et al.*, 2007).

Over the last decade a new process has been developed that may prove useful in fine coal beneficiation. The Boycott phenomenon (discovered in 1920 by A.E. Boycott, where settling rates of suspensions drastically increase when the settling vessel is at an incline (Boycott, 1920)) was adopted to develop this new technology – the Reflux Classifier (Nguyentrannam & Galvin, 2004). The technology was developed at the University of Newcastle in Australia by Prof. K.P. Galvin, and the company that oversees the fabrication and distribution of commercial reflux classifiers in South Africa is FL Schmidt Pty Ltd. Reflux classifiers are already being used in countries like Australia, China and Mozambique, but to a much lesser extent in South Africa.

1.1 Background

In a perfect separation scenario, one would recover all of the valuable material at a hundred percent grade. In reality this is impossible – not only does a perfect separation process not exist, but minerals or valuable materials are never fully liberated. Middlings are always present due to incomplete liberation in comminution circuits and have the potential to report either to the concentrate or discard streams (Wills & Napier-Munn, 2006:16).

Another factor causing ineffective separation is particles or material being misplaced due to characteristics inherent to the material or the separation equipment. Ultra-fine material ($-45\mu\text{m}$), for example, has very low settling velocities due to its small size, and these ultra-fine particles are usually transported along with water without being subjected to the separation force of the particular separation process, decreasing the grade or recovery (Wills & Napier-Munn, 2006:214; Honaker *et al.*, 2001).

Fines impair separation efficiency in a number of other ways as well. Gravity separation processes, for instance, rely on particles having differing settling velocities. The settling velocity of a particle is strongly dependent on the size of the particle and the density

difference between the settling medium and the particle. The difference in settling velocities diminishes when particles become too small, causing low separation efficiency. This effect is especially noted in the Stokes' regime, and Table 1-1 lists the influence of a particle's diameter at certain size fractions. Therefore, gravity separation processes are generally used in cases where the particles are larger than 0.1 mm (Galvin *et al.*, 2010c).

Table 1-1: Listing the extent of particle diameter influencing settling velocity influences (Galvin *et al.*, 2010c)

Particle Size	Flow Regime	Size Dependence
- 0.1 mm	Stokes'	d_p^2
- 2.0 mm + 0.1 mm	Intermediate	d_p^1
+ 2.0 mm	Newton's	$d_p^{0.5}$

A wide particle size range poses great problems as well. It influences separation efficiency negatively, because big light particles and small heavy particles may have the same settling velocity.

1.2 Motivation

Figure 1-1 shows how the demand for coal as an energy source increases over the time axis; historically as well as in future projections. From 2010 to 2025 the increase in energy required from coal as a source will have grown by over 50 quadrillion Btu. Thus, more ROM coal needs to be utilised, and as high-grade feedstock is diminishing, better beneficiation methods should be investigated. Too much coal is being discarded, because the assumption exists that it is not economically viable to beneficiate fine coal. Technology has been developed over the years to upgrade fine coal into useful products, but many of these processes are not in use today. These are usually considered as not efficient enough, or expensive to maintain and operate. Hence, fine coal beneficiation remains an active research area.

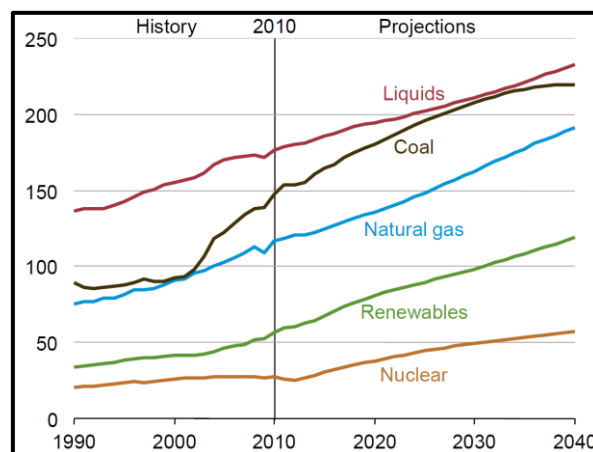


Figure 1-1: World energy demand by the type of fuel in quadrillion Btu (Doman & Arora, 2013:11)

Studies have shown that the reflux classifier is a promising new technology in fine coal beneficiation. The reflux classifier was developed and works for Australian coal fields. Since both Southern African and Australian coals are Gondwanaland coals, the assumption is that the reflux classifier will work on South African fine coal as well. There is no data on the reflux classifier's performance on South African coals, and pilot trials are currently being done in South Africa to test the reflux classifier's performance.

While pilot plants are usually used to test the applicability of a new technology to specific coals, the need exists to gather more fundamental data at a laboratory scale. This study is aimed at introducing a way to pre-test material prior to pilot plant trials in the design chain. This is done by correlating existing pilot test results to laboratory generated data.

The envisioned correlation will save manpower, costs and time. The range of extensive pilot scale trials will be reduced. Fewer runs mean less sample consumption (4 to 6 tons per run). By knowing the planned pilot runs' specifics (from the proposed laboratory scale work), time is saved and manpower is reduced. The designed and built laboratory scale reflux classifier was, however, not suitable for continuous operation. Numerous continuous runs were done, but to no avail, and the focus was shifted to using the unit that was built as a coal fractionator and batch processing system.

Background on washability analyses

The conventional method for generating densimetric curves are by the use of heavy liquids in a series of dense medium baths. These heavy mediums can be organic liquids, dissolved salt solutions or dense powder suspensions. There are a few safety and environmental concerns when using these substances. Heavy organic liquids like tetra-bromo-ethane, or TBE, for instance is very toxic and can cause targeted organ failure (Sciencelab.com, Inc., 2005a) while zinc chloride is corrosive and may permeate through the skin (Sciencelab.com, Inc., 2005b). If handled correctly, mediums used to make-up different densities for the purpose of performing washability analyses can be used safely without incidents. When disposing of spent mediums, the environment should be taken into account, and replenishing the spent medium is also expensive.

A water only fractionation process will eliminate the use of hazardous mediums, and the reflux classifier is such a process where only water is used.

Closing argument

The reflux classifier has the following attributes:

- High throughput, due to increased settling rates within inclined channels.
- Partial elimination of the particle size effect when a wide particle size distribution is treated.
- Operating costs are low.
- Maintaining the unit at optimal operating conditions is simple.
- Adjusting the operating variables when a different coal product is desired is easy.

1.3 Scope of Investigation

1.3.1 Design and commission of a laboratory-scale reflux classifier

A laboratory-scale reflux classifier was designed and built by applying findings from literature (detailed in section 3.1) where similar units were built under Professor Kevin Galvin's guidance at the University of Newcastle in Australia. Extensive effort and time were put into the design of the reflux classifier. The unit was then built with the assistance of the School of Chemical and Minerals Engineering Workshop, after which it was commissioned, implemented and used. A few final adjustments were necessary, and the fully functional unit was completed in due time.

1.3.2 Preliminary test work

Extensive tracer tests were performed in order to correlate density cut-point and fluidisation water volumetric flow rate. Some coal tests were also considered as preliminary tests. These experiments were done on a fractionation basis with the aim of proving that a laboratory-scale reflux classifier is capable of producing repetitive results.

1.3.3 Main experiments

The main experiments involved extensive batch operations, where the density cut-point remained the key variable. These tests involved coal from two different sources and both fractionation and separation procedures (described in section 3.1.3) were followed. In order to correlate pilot unit data and laboratory generated data, bed density within the fluidisation section was considered the primary variable in semi-batch operations, which is discussed in sections 5.1 and 5.2.

Water only coal fractionator

The key area of investigation was that of producing washability data with a small batch operated water only laboratory scale reflux classifier. Water based washability analyses have many advantages over normal float sink baths. More detail is provided in section 4.2, where washability data from float and sink analyses are compared to reflux classification fractionation procedures.

Auxiliary tests

The influence of particle size on separation performance was evaluated by isolating the different particle size ranges of a test. The scope of investigation did not include different particle size range tests, but the responses of large light particles and small heavy particles were evaluated by means of their mineral content.

Correlating laboratory and pilot scale reflux classification

The need was identified to determine if the RC105 (the NWU's reflux classifier unit) can be used to reproduce pilot scale reflux classification data, and the possibility was explored further.

1.4 Dissertation Outline

A brief outline of the dissertation follows to explain the contents and flow. The outline of the dissertation is summarised in Figure 1-2.

The literature review covers the topics of:

- Coal – where coal originated and how it is characterised
- Coal beneficiation – explaining other processes used in coal beneficiation and some historical aspects
- Fluid dynamics – describing the different flow regimes and pressure drop
- Solid-liquid systems – exploring the various implications of particle-particle and liquid-particle interactions on the settling velocity of single and multiple particle systems
- Reflux classification – introducing the process, explaining its role in coal beneficiation and how it separates on density, listing the application thereof and describing it fundamentally and empirically.

The experimental chapter is in two parts:

- One part covers the manner in which the RC105 was designed, built and commissioned; the sizing of the associated process equipment; and the operation thereof.
- The other part describes the experimental method, including other equipment used, materials used and how the data was analysed from sampling to the final data generated.

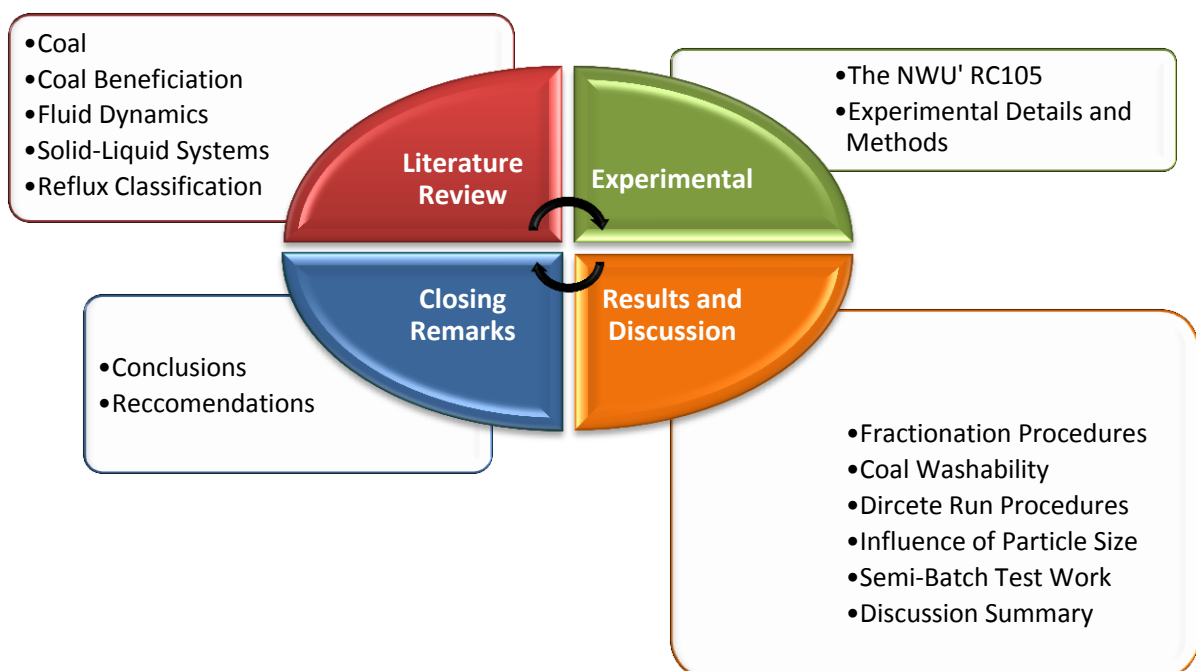


Figure 1-2: Outline of the dissertation

Introduction

Chapter 4 explains the entire results campaign and attempts to draw meaningful conclusions from the fractionation and separation procedure results; and the auxiliary objectives which include the influence of particle size. Whether a laboratory scale reflux classifier can be used for washability analyses was investigated by comparing float and sink washability data with reflux classification fractionation data. The discussion summary ties all of the findings together.

The dissertation closes off with a few remarks which include the conclusions of the investigation, and gives a few recommendations on the topic of:

*Correlating laboratory and pilot scale reflux classification
of fine coal*

CHAPTER 2 - Literature Review

2.1 Coal

2.1.1 *Origin*

Coal is formed when organic material metamorphoses under heat and pressure after having been deposited in the earth's crust. Today's coal fields were formed at a time when two supercontinents existed – Laurasia (presently the northern hemisphere) and Gondwana (presently the southern hemisphere) (Falcon & Ham, 1988).

Laurasian coals were formed from carboniferous swamps in hot and humid coastal areas, whereas Gondwana coals were formed from the Permian swamps, which varied from very cold to hot areas. The vegetation in the Gondwana region varied accordingly, and consisted of the following (Falcon & Ham, 1988):

- Vegetation associated with sub-arctic conditions
- Cold to cool deciduous forests (typically hardwood)
- Warmer woodlands in savannah-like areas
- Swamps infested with reeds

Not only did the vegetation in these regions differ, but Gondwana coal fields were formed subsequently to coal fields from Laurasia – making many of the southern hemisphere coals younger coalfields. Coal field deposits of the Ecca beds in the Karoo basin were formed 200 million years ago (England, 2002:1).

Unlike the Gondwana region, vegetation from Laurasia was more uniform and mainly contained an abundance of ferns and enormous lycopod horsetails and other fleshy-barked trees. These plants are classified as sub-tropical equatorial species and are well-accustomed to water (Falcon & Ham, 1988).

All of these factors contribute to the wide variety of characteristics found in southern hemisphere coals and the big difference between southern and northern hemisphere coals. The European coals are more 'mature' in rank than South African coals, for instance, but due to igneous activity (magma intrusions) bituminous coal and some anthracite are mined in South Africa. The Gondwana coals are also located closer to the surface, and are easier to mine when compared to coalfields in North America and the European countries (Falcon & Ham, 1988).

Although South Africa's coal is located at relatively shallow depths, our coal tends to be more difficult to beneficiate. This is due to the manner in which the plant material was deposited and buried. Much of the organic matter was washed into shallow seas or swamps, which contained mineral matter as well. During the coalification processes this mineral

matter became inherent to the coal. This mineral matter (higher density fraction) impairs calorific value, thus causing South African coals to be of a lesser quality when compared to northern hemisphere coals – where a more ‘stable’ deposition of plant material occurred.

2.1.2 Characterisation

Since coal is so diverse, an appropriate characterisation system is required. There are two levels for characterising coal (Falcon & Ham, 1988):

- Empirical
- Fundamental

Empirical levels

This level of characterisation relates to the chemical and physical properties of the coal, and includes the following analyses (Falcon & Ham, 1988):

- Proximate analysis (inherent moisture, ash content, volatile matter and fixed carbon)
- Ultimate analysis (ultimate chemical components: carbon, hydrogen, oxygen and nitrogen)
- Calorific value
- Swelling index

Fundamental levels

This level of characterisation relates to the main building blocks of coal (the organic and mineral matter) and coal rank (degree of coalification). These analyses are done with a microscope (Falcon & Ham, 1988).

The organic matter content constitutes the different macerals in a specific coal. Macerals vary from reactive (vitrinite and exinite) to less reactive macerals (inertinite). There are numerous other macerals, but they are not that important to South African coals, except liptinite. South African coals vary a lot as far as macerals and organic matter are concerned, because the original plant materials did indeed differ in different areas in the Gondwana region (as explained in section 2.1.1). The mineral matter concerns the inherent mineral content of the coal.

The rank of coal is determined by the degree of coalification that the organic matter has undergone. The coalification steps or stages are as follows (England, 2002:9-10):

- Peat
- Lignite or brown coal
- Sub-bituminous coal
- Bituminous coal
- Lean coal
- Anthracite
- Graphite (carbon only), where no volatile material is left

2.2 Coal Beneficiation

2.2.1 A historical overview

Up until the late nineteenth century, the main methods of coal preparation were crushing and screening – ‘coal cleaning’ was done by hand picking. Only at the very end of the 1800s was wet cleaning considered to be practical in coal cleaning processes (Carris, 2007:4).

The first fine (-1 mm) coal cleaning processes were developed in 1920, amongst others the Deister Overstrom concentrating tables, and subsequent (1940s) equipment included vibratory screens, drums, disc filters and centrifugal dryers. Dense-medium washers (using magnetite as the medium) were developed in the 1930s, but were introduced to fine coal cleaning in the late 1950s (Carris, 2007:5).

The first dense-medium fine coal beneficiation plant was implemented in Belgium in 1957 – it was called Tertre. A second fine coal dense-medium plant was built in Belgium at the Winterslag Mine in 1966. This plant was in operation for about 17 years and performed adequately (De Korte, 2002). By the 1980s, a dense-medium cyclone plant was installed at Greenside Colliery, in South Africa, to clean coal in the range of 0.5 mm × 0.15 mm particle size. This plant operated successfully for 18 years and thereafter none of these plants were in operation anymore (England, 2002:185; De Korte, 2002). This may be attributed to the advent of spirals. In 2007, however, Exxaro installed the first double stage ultra-fine dense-medium cyclone plant at Leeuwpan mine (Lundt & De Korte, 2010).

In the past, spirals were used in many heavy mineral separation processes. It was only in the 1980s, however, that spirals were developed for coal beneficiation, and specifically fine coal beneficiation (England, 2002:179). Spirals then became the process of choice in fine coal beneficiation (De Korte, 2002). Recent developments also show that spirals may be very effective in cleaning coal in the 1 mm × 0.15 mm size range when the parameters are monitored correctly in the new configuration of the two-stage cleaner spiral circuits (Luttrell *et al.*, 2007:85).

2.2.2 Beneficiation methods in fine coal

Gravity separation is used extensively in treating coal, tin ores and many other industrial minerals. It is also the most used method in concentrating iron and tungsten ores. With the development of froth flotation, utilisation of gravity separation for some mineral processing areas has declined (Wills & Napier-Munn, 2006:225).

In the coal cleaning industry, the principal method of coal beneficiation remains gravity separation. Conventional beneficiation methods are also water-based processes (Macpherson, 2011:23) and the following processes used in fine coal beneficiation are going to be discussed in this section – note that oil agglomeration and froth flotation separate particles according to the particle’s physical properties and not on a gravitational basis:

- Oil agglomeration
- Froth flotation
- Dense-medium cyclones
- Spirals
- Teetered bed separators

Coal beneficiation processes for larger coal particles include jigs (process equipment includes the alljig, Batac and Apic jig) and dense medium separators (these processes include float and sink separations, dense-medium baths and vessels that employ a centrifugal force) (England, 2002:139, 150, 156, 163).

Oil agglomeration

Oil agglomeration consists of adding an immiscible liquid (e.g. hydrocarbon/fuel oil) to a suspension containing fine coal. With suspension agitation, distribution of the oil over oleophilic (hydrophobic) particle surfaces occurs and promotes the forming of liquid bridges between particles. Hence, agglomerates are formed. By adding froth flotation agents, controlling of the oleophilicity for particular minerals is possible (Wills & Napier-Munn, 2006:13).

Apart from laboratory use, oil agglomeration is not used in ultra-fine ($-150\ \mu\text{m}$) coal beneficiation (Wills & Napier-Munn, 2006:13). The main use of oil agglomeration in coal beneficiation, is fine coal ($-500\ \mu\text{m}$) beneficiation with the auxiliary effect of fine coal dewatering (England, 2002:184).

In order to obtain good dewatering and upgrading of the coal, 5% to 10% oil addition is required. This method of fine coal beneficiation is not used in South Africa, because it is not economically viable to use even 5% oil, with the price of oil in South Africa as high as it is (England, 2002:184).

Froth flotation

Froth flotation processes are a physico-chemical method and exploit the difference in surface properties of the particles being segregated. Three mechanisms are distinguished in which particles may report to the overflow:

- particle becomes attached to an air bubble
- particle entrainment
- physical entrapment of a certain particle

The first mechanism is when true flotation occurs and determines the recovery of the valuable mineral or particle. The degree to which the latter two occur, influences the separation efficiency of the process (Wills & Napier-Munn, 2006:267).

It was initially believed that South African coals are not very compatible with froth flotation, with one exception being the KwaZulu-Natal coking coals (England, 2002:180). However,

after test work had been done at the University of Cape Town, these plants were gradually being utilised. Also, it was confirmed that ultra-fines from the Witbank coalfields beneficiated by using flotation methods were technically feasible, and that roughly 12% of South Africa's coal processing plants utilised flotation as a beneficiation method for upgrading ultra-fines (Reddick *et al.*, 2007).

Froth flotation is a successful process in fine coal beneficiation and is widely used all over the world to beneficiate fine and especially ultra-fine coal ($-150\ \mu\text{m}$), but it will never produce a 7% ash product (England, 2002:184). Non-selective separation occurs and gangue reports to the product launder, while particles smaller than $45\ \mu\text{m}$ report to the froth concentrate in direct proportion to water flow rate, with no regard to whether it is product or not (Kohmuench *et al.*, 2007:97).

This process is also relatively expensive and consumes electricity, and chemicals in the form of collectors (oils such as paraffin) and frothers (alcohols such as methylisobutylcarbinol, but better frothers have been developed) (England, 2002:181). Magnetite consumption in dense-medium beneficiation is more economical (England, 2002:184).

Dense-medium cyclones

The first fine coal dense-medium cyclone washer was built in Belgium (De Korte, 2002). Thereafter many fine coal dense-medium cyclone beneficiation plants were built in the United States, Australia and South-Africa. However, none of these are in operation anymore (De Korte, 2014).

When treating $-4\ \text{mm}$ coal in $800\ \text{mm}$ diameter and greater cyclones, the efficiency deteriorates significantly (England, 2002:164) – meaning that smaller cyclones are needed to effectively beneficiate fine coal. Also, recovering the finer magnetite with magnetic drum separators is difficult, and losses of the ultra-fine magnetite occur. Low ash coal could also not be achieved in these operations at normal pressure (9 times cyclone diameter, which was $45\ \text{kPa}$ for the cyclones at Greenside). Only when the pressure was increased to $120\ \text{kPa}$ was a 7% ash product attained (England, 2002:185). This is in line with an investigation piloted by P.J. van der Walt (Van der Walt, 2003), where he demonstrated that with increased vortex finder acceleration (higher feed rate or pressure) in small diameter ($150\ \text{mm}$) cyclones, an e.p.m. of 0.02 for dense-medium operation can be achieved. He was also of the opinion that a coarser medium in a bigger diameter cyclone operated at lower pressure (12D) would yield the same efficiency for fine coal.

An interest in fine coal beneficiation with dense-medium cyclones was rekindled due to a change in an international coal specification, causing the de-commissioning of some fine coal ($-1\ \text{mm}$) spirals plants – which could not deliver to the required specification. A recently implemented fine coal dense-medium plant at Muhanga Colliery has proved that Van der Walt (2003) was correct, and it is one of the only two fine coal dense-medium cyclone plants currently in the world – where the Leeuwpan Colliery fines cyclone circuit is by-passed due

to product specification changes. Figure 2-1 shows a photo of the fine dense-medium coal plant at Muhanga. An efficiency test was done by Coaltech, and it has shown that the 50 tph plant produces a 28 MJ/kg product at e.p.m. values of 0.053 (which was the lowest *reported* e.p.m. for this plant) (De Korte, 2014).



Figure 2-1: Muhanga's fine coal dense-medium cyclone beneficiation plant (De Korte, 2014)

Spirals

Low capital and operating costs and ease of operation are among the most compelling advantages of spirals. However, there are disadvantages as well. Low efficiencies have been reported (e.p.m values in the order of 0.15) (England, 2002:180) and factors such as beaching and material loss, are a reality. Spirals also have a low throughput, because in order for the spiral's separation mechanisms to operate efficiently, the water film within the spiral must be very thin (Macpherson, 2011:13).

There have been new advances in this area, possibly due to a better understanding of the flow pattern (now called the counter-rotation flow pattern) that the material experience within a spiral (Luttrell *et al.*, 2007:76). The efficiency data of a two-stage cleaner spiral circuit is presented in Table 2-1, and it portrays excellent results. Beaching is still not entirely eliminated in these new developments, and the low throughput experienced remains the same. When a top-size of 3 mm is fed into a spiral circuit (compared to the normal -1 mm particles), the beaching effect is also dramatically increased by these larger particles (De Korte & Bosman, 2007).

Table 2-1: Performance data on a two-stage cleaner spiral circuit (Luttrell *et al.*, 2007:81)

SG Class	Clean [%]	Refuse [%]	Partition [%]
Float – 1.6	97.30	1.40	100.0
1.6 – 1.8	1.65	0.96	54.41
1.8 – Sinks	1.03	97.63	1.36

Teetered bed separator

An increasing interest is shown in fluidisation equipment, like a teetered bed separator (TBS), as an alternative option to spirals (De Korte & Bosman, 2007). The TBS is comprised of a fluidised tank that is operated continuously by feeding the tank in the middle, while the overflow is captured and the high density product (or discard) is withdrawn from the bottom. It is allowed to form an autogenous bed to promote phase inversion, and the light particles are rejected from the higher density suspension to the upper part and consequently removed via the overflow (Walton, 2011:28, 29).

TBS equipment is generally used to treat particles in the -2 mm +0.25 mm size range in the coal industry. It is also operated above the minimum fluidisation, which maximises density separation effects and minimises mixing (Walton, 2011:28, 29). TBS equipment also promotes higher throughputs and process the same amount of tonnage that an array of spirals can achieve. In this fluidising equipment, it is also easier to control the material comprised within the unit, making it possible to change the effective density cut-point (De Korte & Bosman, 2007). The downside of gravity separation processes is, however, their dependence on particle size, and this is greatly aggravated within fluidisation equipment like the teetered bed separator.

2.3 Fluid Dynamics

To describe fluid and particle behaviour in solid-liquid process equipment, one must revise some fundamentals of fluid dynamics. These were also used in designing the laboratory scale reflux classifier unit. The reflux classifier is described empirically as well as fundamentally, and some important fluid dynamic and solids-liquid principles were applied.

2.3.1 Flow regime

There are three flow regimes – they are laminar flow, turbulent flow and transitional flow. Transitional flow describes the flow of a fluid reaching the turbulent flow regime that has emerged from laminar flow. When describing fluid flow and especially flow involving particle systems, it is crucial to know in what flow regime the system is classified.

The Reynolds number relates to inertial forces and viscous forces (Welty *et al.*, 2008:128), and is used to characterise the flow of a fluid. The flow is laminar when viscous forces dominate and as the fluid velocity increases for a fixed system, inertial forces become more dominant. When only inertial forces dominate, the turbulent flow regime is reached (Kelly & Spottiswood, 1989:63).

Equations 2.1 and 2.2 (Kelly & Spottiswood, 1989:63) illustrate the logic in deriving the Reynolds number (Re) for flow in circular pipes (equation 2.3); where ρ_F is the fluid density, U the fluid velocity, D the diameter of the pipe and μ_w the viscosity of water. The transition from laminar to turbulent flow occurs when the Reynolds number (also referred to as the critical Reynolds number, Re_c) is $2100 < Re_c < 5000$ (Tilton, 1997:6-10). It is then also generally accepted that laminar flow occurs when the Re is below 2100 and turbulent flow occurs when Re exceeds 5000.

$$Re = \frac{\text{Momentum transfer by turbulence mechanism (inertial forces)}}{\text{Momentum transfer by viscous mechanism (viscous forces)}} \quad 2.1$$

$$Re = \frac{\rho_F U^2}{\mu_w U/D} \quad 2.2$$

$$Re = \frac{DU\rho_F}{\mu_w} \quad 2.3$$

The Reynolds number for other conduit shapes can also be determined. When the flow is turbulent, the hydraulic diameter (D_H) of the specific geometrical shape is calculated and can then replace D in equation 2.3. The hydraulic diameter is four times the conduit cross-sectional area divided by the wetted perimeter. Equation 2.4 is used to calculate the hydraulic diameter of a rectangular channel; where a is the width and b the height of the rectangular conduit. The critical Reynolds number also differs from circular conduits and for rectangular channel the following holds: $1900 < Re_c < 2800$. (Tilton, 1997:6-12)

$$D_H = \frac{ab}{2(a+b)} \quad 2.4$$

Inaccurate results are expected when the method of hydraulic diameter is used while the flow is laminar, because the geometrical shape affects the resistance to flow in such a manner that it cannot solely be described by the cross-sectional area and the wetted perimeter. In this case the equivalent diameter D_E is calculated, but D_E cannot replace D in equation 2.3 for reasons described in section 2.3.2 (Tilton, 1997:6-12).

2.3.2 Pressure drop and frictional losses

When a fluid is flowing through a conduit, one is normally concerned about pressure drop and flow rate. Pressure drop is directly linked to frictional loss, and head loss due to a height difference, fittings and valves.

Equations for frictional loss are derived by solving a pipe system by means of dimensionless analysis, where two of the dimensionless numbers are the Reynolds number and the friction factor (f). Just as in the case of the Reynolds number (section 2.3.1) the friction factor can be expressed as a ratio of momentum transfers, equation 2.5. The dimensionless number that was obtained for the pipe system is called the Fanning friction factor (f_f), and is given in equation 2.6 (Kelly & Spottiswood, 1989:66; Tilton, 1997:6-10).

$$f = \frac{\text{total momentum transfer}}{\text{momentum transfer by turbulence}} \quad 2.5$$

$$f_f = \frac{D \Delta P}{2\rho u^2 L} \quad 2.6$$

For fluid flows in the laminar and turbulent flow regimes, equations 2.7 and 2.8 can be used to determine the friction factor. Equation 2.7 is called the Hagen-Poiseuille equation and may be derived from the Navier-Stokes equations. It may also be rewritten in the form of the volumetric flow rate, equation 2.9. Equation 2.8 is called the Blasius equation (Kelly & Spottiswood, 1989:66; Tilton, 1997:6-10).

$$f = \frac{16}{Re} \quad ; Re \leq 2100 \quad 2.7$$

$$f = \frac{0.079}{Re^{0.25}} \quad ; 4000 \leq Re \leq 10^5 \quad 2.8$$

$$Q = \frac{\pi \Delta P D^4}{128\mu L} \quad ; Re \leq 2100 \quad 2.9$$

In turbulent flows, wall roughness (ϵ) begins to play an important role and should be considered in the determination of the friction factor. The Colebrook formula (equation 2.10) gives a good estimate for the friction factor in turbulent flows where the pipe isn't smooth, but an equation by Churchill (equation 2.11) is explicit in f and holds for smooth as well as rough pipes (Tilton, 1997:6-10).

$$\frac{1}{\sqrt{f}} = -4 \log \left(\frac{\epsilon}{3.7D} + \frac{1.256}{Re\sqrt{f}} \right) ; Re > 4000 \quad 2.10$$

$$\frac{1}{\sqrt{f}} = -4 \log \left(\frac{0.27\epsilon}{D} + (7/Re)^{0.9} \right) ; Re > 4000 \quad 2.11$$

For turbulent flow and other conduit shapes the hydraulic diameter (D_H) can again replace D in equation 2.6. The hydraulic diameter may only be used for turbulent flows. For laminar flows the equivalent diameter is used to determine the pressure drop over a conduit, and D_E may only substitute D in equation 2.9 and not in equations 2.3 and 2.6, and hence not in equation 2.7 (Tilton, 1997:6-12). This is because the velocity of the fluid is not equal to the volumetric flow and area relation when the area is determined by the equivalent diameter (equation 2.12 shows this inequality) (Tilton, 1997:6-12).

$$U \neq \frac{4Q}{\pi D_E} \quad 2.12$$

The equivalent diameter for rectangular channels can be determined by means of equation 2.13; where a is the channel width, b the channel height and K a parameter related to the spacing of the rectangular channel and given in Table 2-2 (Tilton, 1997:6-12).

$$D_E = \left(\frac{128ab^3}{\pi K} \right)^{1/4} \quad 2.13$$

Table 2-2: Parameter K relation to the channel spacing

$a/b =$	1	1.5	2	3	4	5	10	∞
$K =$	28.45	20.13	17.49	15.19	14.24	13.73	12.81	12.00

2.4 Solid-Liquid Systems

2.4.1 Rheology

To determine the viscosity of coal slurries is complex, and this section will only provide a brief discussion of the topic of rheology and propose a rough approximation for the viscosity of a slurry.

Normally particle suspensions that are comprised of spherical particles bigger than 50 μm are considered to be Newtonian liquids. Slurries containing particles of irregular shapes start to exhibit Bingham plastic non-Newtonian behaviour even at particle sizes as small as

50 μm . Equation 2.14 gives an estimate for the viscosity of slurry containing irregular shaped particles; where $(y_{V,sl})$ is the volumetric fraction of the solids in the slurry (Kelly and Spottiswood, 1989:65).

$$\frac{\mu}{\mu_w} = 1 + 2.5y_{V,sl} + 14.1y_{V,sl}^2 + 0.00273 \exp(16.0y_{V,sl}) \quad 2.14$$

2.4.2 Forces on a submerged particle

Before liquid-particle systems can be described, the forces that are acting on a particle submerged in a fluid need to be identified. These forces are gravity, drag and buoyancy.

Gravity and buoyancy

When a particle is submerged in an infinite motionless fluid, the density of the fluid relative to the density of the particle dictates whether a particle will sink or float. The resultant force acting on an irregularly shaped particle in the described system can be expressed by equation 2.15; where (P_i) represents the pressure exerted on the top and bottom surfaces (dA) of the particle in the vertical direction z (the vector notation is used to indicate that the top and bottom surface may not be in line with the horizontal plane); ρ_P the particle density; g the gravitational acceleration and h the height of the particle. The difference in pressure $(P_1 - P_2)$ can be substituted by $\rho_F gh$ as in equation 2.16; where ρ_F is the fluid density (Welty *et al.*, 2008:24).

$$d\mathbf{F} = (P_1 - P_2) dA \mathbf{e}_z - \rho_P gh dA \mathbf{e}_z \quad 2.15$$

$$d\mathbf{F} = \rho_F gh dA \mathbf{e}_z - \rho_P gh dA \mathbf{e}_z \quad 2.16$$

After integration it is noticed that the resultant force is a function of the particle volume (V_P), the gravitational acceleration and the difference in density of the fluid and the particle (equation 2.17). It is also noted that the resultant force can be split into two separate forces – the forces of gravity and buoyancy – and that these forces work in opposite directions (Welty *et al.*, 2008:24).

$$\mathbf{F} = (\rho_F - \rho_P)gV_P \mathbf{e}_z \quad 2.17$$

For a spherical particle equation 2.18 gives the gravitational force (F_G) acting on the particle and is in the downward direction; where d is the particle diameter.

$$F_G = \left(\frac{\pi d^3}{6}\right) \rho_P g \quad 2.18$$

Buoyancy is a force that works in the opposite direction than that of gravity and equation 2.19 gives the buoyancy force (F_B) for a particle submerged in a fluid.

$$F_B = \left(\frac{\pi d^3}{6} \right) \rho_F g \quad 2.19$$

Drag

Drag is due to the behaviour of the relative motion of the particle and the fluid. Whether a particle is moving relative to the fluid or a fluid is moving relative to the particle, the fluid will exert a force on the particle called the drag force (F_D). The drag force is proportional to the rate of momentum transferred by the fluid (Darby, 2001:341). Equation 2.20 is used to describe the drag force; where C_D is the drag coefficient, A the projected area of the particle in the direction of flow, ρ_F the density of the fluid and U_R the relative velocity of the fluid to the particle.

$$F_D = C_D A \frac{\rho_F U_R^2}{2} \quad 2.20$$

The drag coefficient is dependent on the particle Reynolds number described by equation 2.21; where U is the fluid superficial velocity, μ the fluid viscosity and d_p the particle diameter. For particle Reynolds numbers below 2 the drag coefficient is linear to the Reynolds number and is determined by equation 2.22 (Svarovsky, 2000:528). This is also called the Stokes' flow regime where the flow can be described as creeping flow. When $10^3 < Re_p < 10^5$ (Newtonian flow regime), equation 2.23 is used to determine the drag coefficient (Darby, 2001:343).

$$Re_p = \frac{\rho_F U d_p}{\mu} \quad 2.21$$

$$C_D = \frac{24}{Re_p} ; Re_p < 2 \quad 2.22$$

$$C_D = 0.44 ; 10^3 < Re_p < 10^5 \quad 2.23$$

In the transition from laminar to turbulent flow, the determination of the drag coefficient is more complex. Wadell has proposed the following correlation (equation 2.24) to determine drag coefficient very accurately for all particle Reynolds numbers up to 2×10^5 (Darby, 2001:344), but for simplicity the following relation (equation 2.25) is also proposed that was developed by Vance (1965) (Walton, 2011:17).

$$C_D = \left(0.632 + \frac{4.8}{\sqrt{Re_p}} \right)^2 \quad 2.24$$

$$C_D = \frac{18.5}{Re_p^{0.6}} \quad 2.25$$

2.4.3 Terminal settling velocity

Figure 2-2 illustrates the forces acting on a suspended particle as described in section 2.4.2. When the gravity force is greater than the buoyant and drag forces the particle will free fall or settle out from the fluid. As the particle accelerates, the drag force will increase and cause the particle to decelerate until a constant settling velocity is reached. The drag force equals the gravitational force, which is described as the terminal settling velocity (Darby, 2001:347).

An equation can be derived for the terminal settling velocity by constructing a momentum balance (equation 2.26) over this system; ρ_P is the density of the solid or particle. The term du/dt is set to zero for a constant settling velocity and the solution is given in equation 2.27, which gives the terminal settling velocity (Darby, 2001:347).

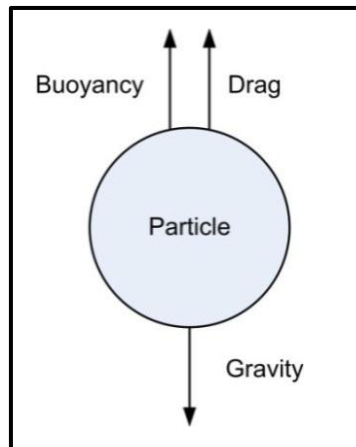


Figure 2-2: Forces acting on a particle settling within a fluid

$$\frac{g(\rho_s - \rho_F)\pi d_P^3}{6} - \frac{C_D \rho_F \pi d_P^2 u^2}{8} = \frac{\pi d_P^3 (\rho_s + \rho_F/2)}{6} \frac{du}{dt} \quad 2.26$$

$$\text{When } du/dt = 0 ; \quad u_t = \sqrt{\frac{4g \Delta \rho d_P}{3\rho_F C_D}} \quad 2.27$$

The drag force can then be expressed in terms of the terminal settling velocity of a single particle, equation 2.28.

$$C_D = \frac{4g \Delta \rho d_P}{3\rho_F u_t^2} \quad 2.28$$

Since the drag force is dependent on the Reynolds number (as seen in section 2.4.2) and directly dependent on the terminal settling velocity, the terminal velocity should be determined by incorporating knowledge of the flow regime. For the Stokes' regime equation 2.22 is used to arrive at equation 2.29, and for the Newtonian flow regime equation 2.23 is used to arrive at equation 2.30 to determine the settling velocity of a single spherical particle (Darby, 2001:347; Rhodes, 2008:31, 32). For the intermediate regime, the terminal velocity of a spherical particle is given by equation 2.31, where this equation is also derived by introducing equation 2.25 to equation 2.28 (Walton, 2011:18).

$$u_t = \frac{d_p^2 \Delta \rho g}{18\mu} \quad ; Re_p < 2 \quad 2.29$$

$$u_t = 1.74 \sqrt{\frac{d_p \Delta \rho g}{\rho_F}} \quad ; 10^3 < Re_p < 10^5 \quad 2.30$$

$$u_t = 0.153 \frac{d_p^{8/7} \Delta \rho^{5/7} g^{5/7}}{\rho_F^{2/7} \mu^{3/7}} \quad ; 2 < Re_p < 10^3 \quad 2.31$$

It is then also noted that the terminal settling velocity of a particle is proportional to the square of the particle diameter and dependent on the fluid viscosity in Stokes' regime, while it is only proportional to the square root of the particle diameter in the Newtonian flow regime (Rhodes, 2008:32).

Irregularly shaped particles

Particle shape affects the drag coefficient much more in the Newtonian regime than it does within the Stokes'. A particle settling in laminar conditions will be rotated so that the elongated part of the particle is parallel to the direction of motion, while a particle settling in the Newtonian regime presents its maximum area perpendicular to the fluid flow (Rhodes, 2008:33).

One way of describing an irregularly shaped particle by using only one parameter, is to describe it in terms of the particle's sphericity (φ). Sphericity is defined as the surface area of a sphere (that has the same volume as the particular particle) divided by the surface area of the particle (Rhodes, 2008:33; Tilton, 1997:6-52).

Equations 2.32 and 2.33 are used to calculate the settling velocity of an irregularly shaped particle in the Stokes' and Newtonian regimes respectively. Equations 2.34 and 2.35 are used to determine the parameters K_{St} and K_{Ne} respectively (Tilton, 1997:6-52).

$$u_t = K_{St} \frac{d_p^2 \Delta \rho g}{18\mu} \quad 2.32$$

$$u_t = \sqrt{\frac{4d_p \Delta \rho g}{3K_{Ne}\rho_F}} \quad 2.33$$

$$K_{St} = 0.843 \log\left(\frac{\varphi}{0.065}\right) \quad 2.34$$

$$K_{Ne} = 5.31 - 4.88\varphi \quad 2.35$$

Wall effects

The settling velocity of a particle in the vicinity of a solid boundary is invariantly less than that of a particle settling in an infinite fluid. The wall factor (f_w) can be used as the correction factor for this occurrence by multiplying the calculated terminal velocity in an infinite fluid with the wall factor. The wall factor can be determined with equation 2.36 and is only valid for a particle that is settling according to Stokes' law (Rhodes, 2008:34).

$$f_w = \left(1 - \frac{d_p}{D}\right)^{2.25} ; Re_p \leq 0.3 ; d_p/D \leq 0.97 \quad 2.36$$

In the calculations of the terminal velocity of particles settling in closely spaced inclined channels, Walton (2011) uses an equation proposed by Munroe (1888) to determine the factor for wall effects. This equation is given in 2.37, and describes the reduction in the drag coefficient (equation 2.38) when small particles are settling in a vertical pipe with diameter D .

$$f_w = \left(1 - \left(\frac{d_p}{D}\right)^{3/2}\right)^{-2} \quad 2.37$$

$$C_{D,w} = f_w C_D \quad 2.38$$

Hindered Settling

In sedimentation or fluidisation processes, particle-particle interactions become important – due to high solids concentrations in the suspensions. In these processes hindered settling occurs, because of particle-particle interaction and the upward displacement of water due to particles occupying that particular volume. The settling velocities of the individual particles (or the required flow velocity to carry the particles to the overflow in fluidised systems) subsequently decrease as the solids concentration increases (Galvin *et al.*, 2010c; Kelly & Spottiswood, 1989:76, 77).

To determine the sedimentation velocity of a suspension of particles (u_s), the infamous Richardson and Zaki equation (equation 2.39) is normally used. Where n is dependent on the particle type and size (typical values of n in coal suspensions are 7.50 for coal particles +75 μ m -90 μ m and 6.48 for coal particles +300 μ m -355 μ m (Kelly & Spottiswood, 1989:79), $y_{V,sl}$ is the volume fraction of the solids in the slurry and u_t signifies the terminal settling velocity of a single particle in an infinite fluid, and is determined by using any of the equations 2.29, 2.30, 2.32 or 2.33 depending on the particle shape and flow regime (Kelly & Spottiswood, 1989:79; Richardson & Zaki, 1954).

$$u_s = u_t(1 - y_{V,sl})^n \quad 2.39$$

According to Zigrang and Sylvester (1981), the following correlations (equation 2.40 where $C_{D,S}$ and Re_s are given by equations 2.41 and 2.42 respectively) were developed by Barnea and Mednick for the determination of particle settling velocities in solid-liquid systems, and were published in 1975. It is also noted that when $y_{V,sl}$ equals zero in equations 2.40 and 2.42, they will transform to the equations applicable to a single particle settling in an infinite fluid (equations 2.24, 2.28 and 2.21).

$$C_{D,S} = \left(0.63 + \frac{4.8}{\sqrt{Re_s}}\right)^2 \quad 2.40$$

$$C_{D,S} = \left(\frac{4d_p \Delta \rho g}{3\rho_F u_s^2}\right) \left(\frac{1 - y_{V,sl}}{1 + y_{V,sl}^{1/3}}\right) \quad 2.41$$

$$\sqrt{Re_s} = \sqrt{\left(\frac{u_s d_p \rho_F}{\mu \exp[5y_{V,sl}(1 - y_{V,sl})/3]}\right)} \quad 2.42$$

Since these equation cannot be expressed explicitly in the form of u_s (as seen in equation 2.43, where a and b are given by equations 2.44 and 2.45 respectively – note a and b do not represent channel dimensions, only substitutes), Zanker proposed nomographs as a successful solution. Equation 2.43 was obtained by substituting $\sqrt{Re_s}$ from equation 2.42 into equation 2.40 and combining the result with equation 2.41 (Zigrang & Sylvester, 1981).

$$u_s = a - b\sqrt{u_s} \quad 2.43$$

$$a = \frac{2}{0.63\sqrt{3}} \sqrt{\frac{\Delta \rho g d_p (1 - y_{V,sl})}{\rho_F (1 - y_{V,sl}^{1/3})}} \quad 2.44$$

$$b = \frac{4.8}{0.63} \sqrt{\frac{\mu \exp[5y_{V,sl}(1 - y_{V,sl})/3]}{d_P \rho_F}} \quad 2.45$$

Zigrang and Sylvester (1981) then found a solution for equation 2.43 through some mathematical manipulation by rewriting equation 2.43 in the quadratic form (equation 2.46), of which an explicit solution for u_s exists and is given by equation 2.47, where c is given by equation 2.48 (Zigrang & Sylvester, 1981).

$$u_s^2 - (2a + b^2)u_s + a^2 = 0 \quad 2.46$$

$$u_s = c - \sqrt{c^2 - a^2} \quad 2.47$$

$$c = \frac{1}{2}(2a + b^2) \quad 2.48$$

2.4.4 Inertial lift

The phenomenon where a particle is dragged along a surface plane and suddenly separates from the plane wall is called inertial lift. Factors that directly influence inertial lift are the particle shear rate and sedimentation velocity. The ratio of the shear Reynolds number (Re_{sh} ; equation 2.49) to the sedimentation Reynolds number (Re_s ; equation 2.50) gives an indication when this phenomenon will occur (King & Leighton, 1997); γ is the shear rate of the particle, $u_{s,st}$ is the Stokes' sedimentation velocity (particle settling velocity) of the particle and r_p the particle radius.

When a fluid is moving over a plane wall, a particle (initially at rest) will start to be transported along the wall surface. The transition from non-slip to slip translation across the wall surface occurs when the ratio of Re_{sh}/Re_s is in the order of 1 (King & Leighton, 1997). When the ratio Re_{sh}^2/Re_s is nearing 4 (King & Leighton, 1997), the particle will tend to separate from the plane wall and the effect of inertial lift is imminent.

$$Re_{sh} = \frac{\gamma \rho_F r_p^2}{\mu} \quad 2.49$$

$$Re_s = \frac{u_{s,st} \rho_F r_p}{\mu} \quad 2.50$$

2.5 Reflux Classification

The reflux classifier is a relatively new technology, designed to separate particles on the basis of either density or size. It can be described as a lamella settler attached to a fluidised bed (Nguyentranlam & Galvin, 2001). A lamella settler is used in solid-liquid operations,

either to thicken slurries or to clarify a water stream (Svarovsky, 2000:172). Figure 2-3 illustrates how a lamella settler functions. A liquid stream containing solids enters the unit, the solids or particulate matter settle onto the inclined plates and slide down to the sludge draining hopper below. Clear or near clear water leaves the unit via the overflow. The effectiveness of this unit forms its basis on the Boycott phenomenon (Nguyentranlam & Galvin, 2004).

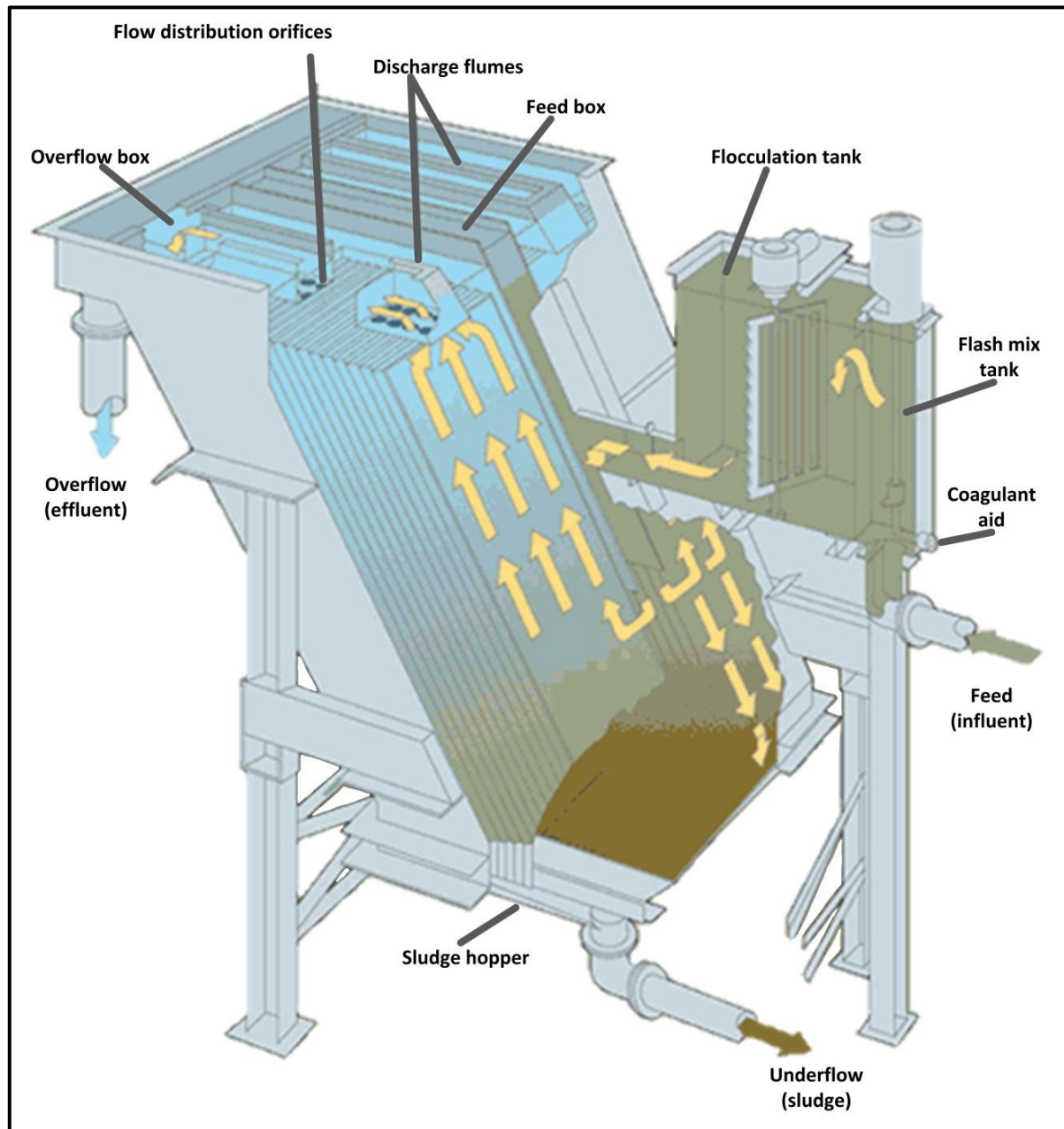


Figure 2-3: Illustration of a lamella settler (Salah, 2012)

A.E. Boycott discovered that when a settling vessel is maintained at an angle, the particle settling rate dramatically increases. He discovered this effect (now called the Boycott phenomenon), when blood corpuscles were to be settled out. He had set the settling tube at an incline, and noticed that the settling velocity increased – at an angle of 67.5° it was

observed that after 20 hours, 42% had settled out, compared to 4% when the tube was at a 90° angle to the horizon (Boycott, 1920). Thus, the principal characteristic that promotes effective sedimentation in the Lamella settler is the parallel inclined plates. This ensures that the settling distance is dramatically reduced and increases the settling rate of the particles.

The compacted array of inclined plates is a contributing to efficiency. It increases the effective settling area to space occupied ratio. A 5.5 m X 3.7 m and 5.2 m high Lamella thickener has an effective settling area of 230 m², which is the equivalent of a 17 m conventional thickener. The vertically projected area of all the plates combined is considered to be the effective settling area of the Lamella thickener (Kelly & Spottiswood, 1989:329).

A reflux classifier can then be defined as a set of inclined plates above a fluidised bed (Figure 2-4). The representation numbered (a) explains the definition mentioned, and (b) shows how an entire reflux classifier design may look. Here the feed is charged into the middle of the unit as indicated (Figure 2-4), and two inclined channel sections are incorporated. Normally only one inclined channel is used while feeding into the middle of the fluidised section – only to indicate the versatility in design considerations. While studies performed prior to 2006 used a channel angle of 60° to the horizon (Galvin & Nguyentranlam, 2002; Doroodchi *et al.*, 2004), the optimum angle of 70° was determined by Laskovski *et al.* (2006).

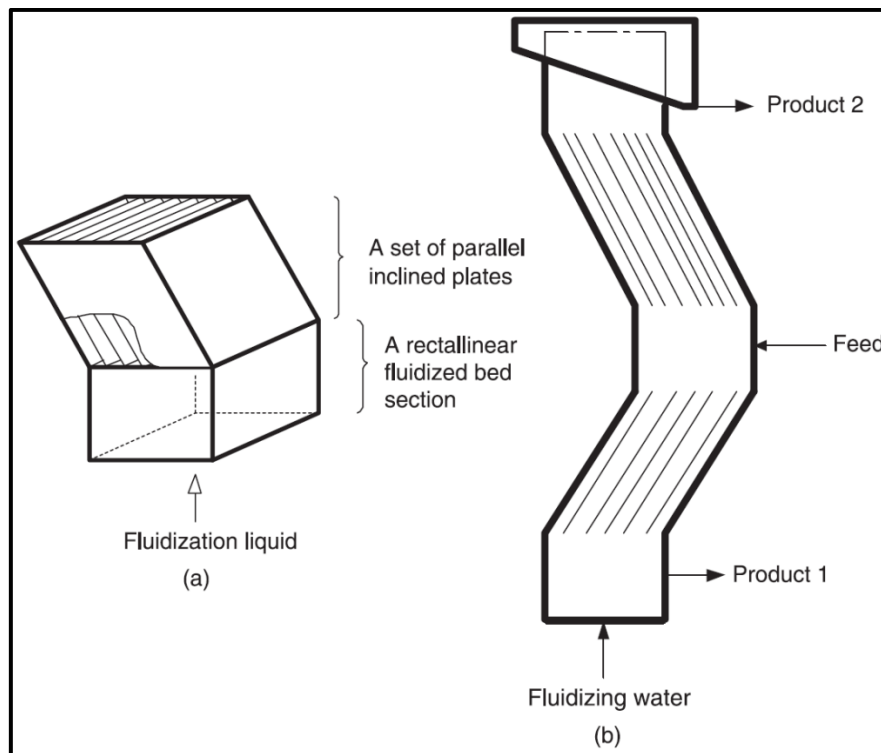


Figure 2-4: Reflux classifier illustration, (a) representing the definition and (b) an example of a reflux classifier design (Nguyentranlam & Galvin, 2004)

As in the case of the lamella settler, the effective settling area of the reflux classifier equals the vertically projected area of all the inclined plates within the classifier. Thus, the reflux classifier also promotes a very high material throughput. The parallel inclined plates within the reflux classifier have the same effect of increasing the settling velocity of the particles.

2.5.1 Influence of inclined plates

Not only does a set of inclined plates improve settling times (which in turn promote higher throughputs in addition to the greater effective settling area), it also improves the separation efficiency by inducing the reflux action and making it possible to maintain a high solids density within the fluidised section at high liquid fluidisation rates.

The recycle action

A study (Zhang & Davis, 1990) was done to examine the effect when the underflow in a single inclined settler is partially recycled. Without the underflow recycle, the separation efficiency of the inclined settler was poor. Insufficient recovery of the fines fraction was reported (with no coarse particles within the overflow), while a mixture of fine and coarse particles (although with little enrichment of coarse particles) was obtained within the underflow.

With the addition of the partial recycle stream, a reduction of the fines fraction was noted in the underflow and the recovery of fine particles increased substantially. The recycling of the underflow definitely promotes a more refined separation (Zhang & Davis, 1990).

When the solids bed is fluidised in the reflux classifier system, the higher hydraulic loading permits the slower settling particles to be forced through the inclined channels so that they may report to the overflow (Galvin *et al.*, 2002). The faster settling particles will settle onto the inclined plates, and slide down to the fluidised section below in the form of a concentrated sediment. A recycle action (or “reflux”) develops. This reflux action is the same as performing the partial recycle of the underflow in the study of Zhang and Davis, 1990. Thus, in principle, the internal reflux action that is developed will also promote a sharper separation to be experienced (Galvin *et al.*, 2005). It is noted that the self-recycling effect helps to prevent particles being misplaced during the separation process, which improves the quality of the segregation (Callen *et al.*, 2008).

High solids concentration

The recycled rejects form a high solids concentration within the fluidised section. This higher density suspension is an autogenous dense medium (Galvin *et al.*, 2002).

K.P. Galvin and G. Nguyentranlam (2002) performed an experiment to test the influence of inclined plates on the solids concentration in a fluidised bed. A set of parallel plates was lowered into the fluidised bed, while the water fluidisation rate was kept constant (Figure 2-5). The fluidisation rate was set low enough to keep the particles from moving past the inclined plates (Galvin & Nguyentranlam, 2002).

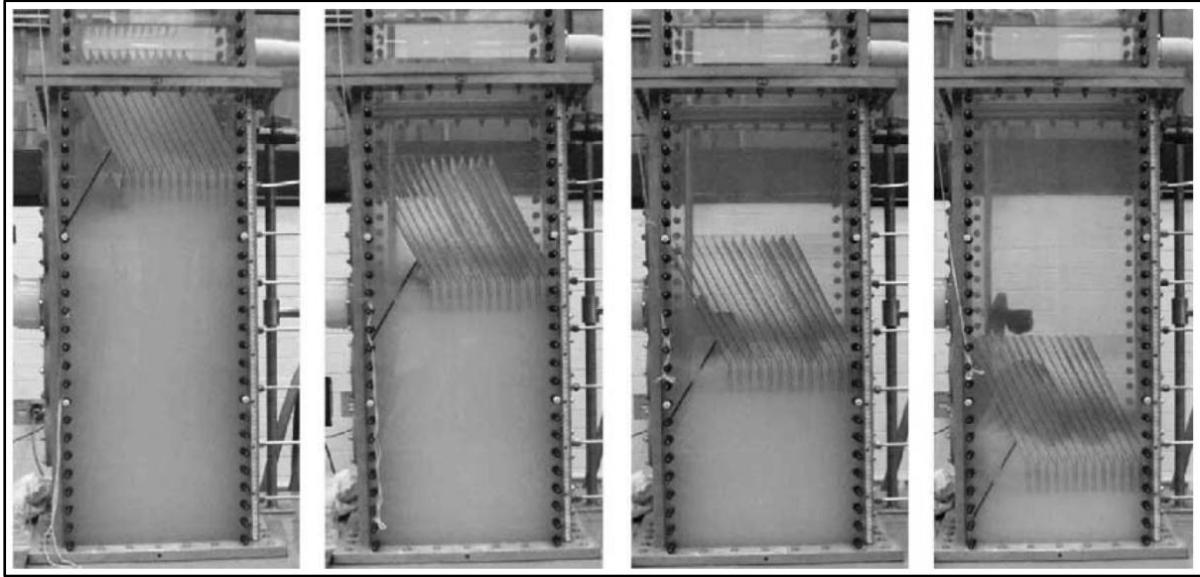


Figure 2-5: Solids concentration increase illustration (Galvin & Nguyentranlam, 2002)

From Figure 2-5 it can be observed that the volume of the fluidising section was reduced when the set of inclined plates was lowered into the fluidised bed. The solids or suspension concentration increased and no particles had been forced through the inclined channels. A fluidisation rate of $6.4 \times 10^{-4} \text{ m}^3 \cdot \text{s}^{-1}$ was achieved, which corresponds to a water flow velocity of $6.4 \times 10^{-3} \text{ m} \cdot \text{s}^{-1}$. The Richardson and Zaki equation was used to determine the average settling characteristics of the suspension, and it was found that the terminal settling velocity of the particles was $3.0 \times 10^{-3} \text{ m} \cdot \text{s}^{-1}$ (Galvin & Nguyentranlam, 2002), less than half the water flow velocity achieved.

In a normal fluidised bed the solids in the bed will expand when the liquid fluidisation rate is increased – diluting the solids concentration in the bed. The reflux classifier system, on the other hand, allows for high suspension concentrations to form, even at fluidisation velocity well beyond the terminal velocities of the particulates in the system. This high solids density (dense medium) that is created also improves the separation efficiency of the equipment, while higher throughputs are obtained with increases in fluidisation rates (Galvin & Nguyentranlam, 2002).

2.5.2 Applications of the reflux classifier

Due to the attainment of a high solids concentration at high fluid superficial velocities and sharp separation qualities, various applications for the reflux classifier exist. Some of these applications include the following:

- Particle size classification and slimes removal
- Generation of coal washability data
- Crystallisation and adsorption
- Density separation

Particle size classification and slimes removal

If the densities of the particles are the same, particle classification can be achieved in a reflux classifier on the basis of particle size. Doroodchi *et al.* (2006) performed tests on silica and obtained good separation efficiencies. A sharp particle size separation was obtained and with a fluidisation rate increase, the cut-point was also shifted to a bigger particle size (Figure 2-6).

It must be noted that better size classification occurs when the bed within the fluidised section of the reflux classifier is dilute or absent (due to expulsion of low density particles from an autogenous dense-medium bed) and that no preferential re-suspension of large low density particles occurs. Both of these cases are advantageous in gravity separation processes, but detrimental in size classification (Walton, 2011:28; 57).

Since the reflux classifier may be used in the classification of fines, slimes removal is also a lucrative application of this technology. Screens have the disadvantage of becoming blinded when too fine material ($-50\ \mu\text{m}$) is classified, and hydro-cyclones on the other hand can effectively classify material smaller than $25\ \mu\text{m}$ down to $5\ \mu\text{m}$, but have the disadvantage that 10% to 20% of the fine material is lost to the underflow, due to fine material leaving the system in direct proportion to the water (Walton, 2011:135).

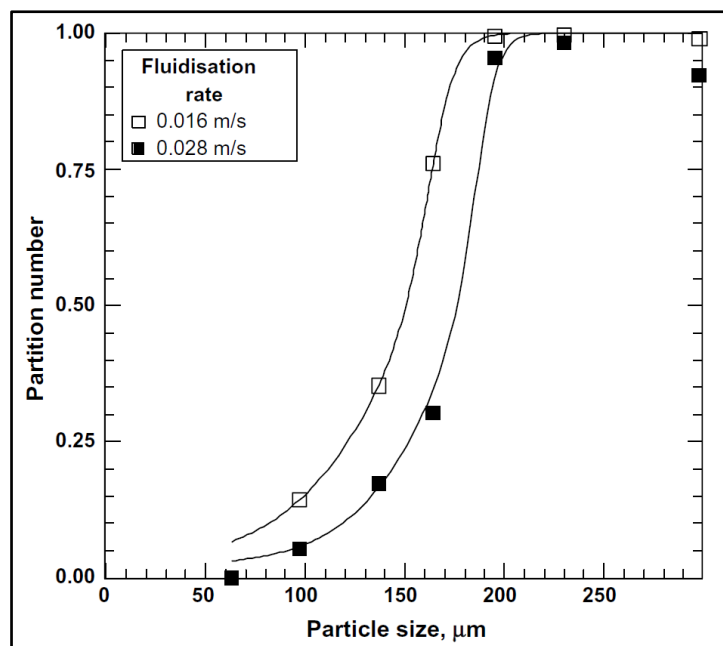


Figure 2-6: Partition curve for size classification in a reflux classifier (Doroodchi *et al.*, 2006)

Due to the fact that the fluidisation water enters from the bottom of the reflux classifier and the underflow is also drawn through the bottom, the disadvantage experienced by a hydro-cyclone is overcome within the reflux classifier arrangement and no slimes are misplaced through the underflow (Walton, 2011:135). When the fluidisation rate is such that only the slimes make it through to the overflow, the assumption is that all of the slimes will be separated from the material.

Should the need exist to de-slime a coal product stream, a two stage reflux classifier system can be implemented – the second unit is used to de-slime the product of the first one. It was observed that the second unit needs to have twice to four times the area of the first unit to process the load, because the fluidisation velocity is significantly lower. The presence of an autogenous bed will also mar the separation efficiency and it would be advisable to have only the superficial channel velocity as the principle elutriation factor (Walton, 2011:153).

Generation of coal washability data

The motivation behind water-based methods to determine washability data is to eliminate the hazardous organic compounds used in traditional float-sink analyses. A study conducted by Callen *et al.* (2008) investigated the possible use of three water-based methods to replace the standard float sink analysis. The Boner jig, the Mintek MDS jig and the reflux classifier were considered for assessment in this study. The conclusion was that it is possible to determine the washability data with these water-based methods, but slight discrepancies were noted at relatively low ash levels (Callen *et al.*, 2008).

For the size range of -50 +16 mm, the Mintek jig proved most efficient and the Boner jig proved adequate for the size range of -16 +0.25 mm, but better results were obtained when this size range was divided into two size ranges of -16 +2 mm and -2 +0.25 mm. The reflux classifier was successful over the whole range and proved exceptionally effective in the size range of -0.25 +0.045 mm, and even produced a higher yield than the laboratory flotation tests. Double fractionation is required, however, when the reflux classifier is used, which entails the determination of the particle size distribution (Callen *et al.*, 2008).

Crystallisation and adsorption

In crystallisation (or adsorption) processes it is highly desirable to increase the rate of mass transfer between the fluid and the particles (or the rate of molecular adsorption onto the particles). When crystallisation or adsorption is performed in a fluidised bed, two situations will increase the rate of mass transfer or rate of molecular adsorption (Nguyentranlam & Galvin, 2004):

- High liquid fluidisation rate (for vigorous mass or molecular interactions)
- High solids density (increased chance for mass or molecular interactions)

When the liquid fluidisation rate is increased in a normal fluidised bed, the bed will become 'diluted', which in turn, jeopardises the crystallisation or adsorption performance. With the aid of the inclined plates in the reflux classifier, the solids density remains high as no solids will leave through the inclined plates until a certain liquid fluidisation rate is reached. Thus, high fluidisation rates at high solid concentrations can be achieved, which will be beneficial in obtaining high crystallisation or adsorption performances (Nguyentranlam & Galvin, 2004).

Density separation

Both the recycle action and high suspension density are advantageous in separating particles on the basis of density. There are also other modifications to the reflux classifier and attributes that promote successful segregation of particles according to their density. The particle settling rate (Galvin *et al.*, 2002), which governs the separation in gravity separation processes, is dependent on the density of the suspension medium and the particle, as well as the particle size. The following are ways to suppress the size dependence within the reflux classifier.

With an increase in the aspect ratio for the inclined plates (plate length over plate spacing), an increase in effective sedimentation area is obtained (which provides a throughput advantage) and a decrease in the separation size is also obtained (Laskovski *et al.*, 2006; Zhou *et al.*, 2006). Laskovski *et al.* (2006), realised that this effect has an asymptotic limit, which is dependent on the particle's Reynolds number, and is due to particle re-suspension when the plate spacing becomes relatively narrow. Laskovski *et al.* (2006) attributed this particle re-suspension to shear induced lift forces that occur when channels are narrowly spaced, but their experimental work was not enough to include this effect in their generalised model (Walton, 2011:58; 59) – as derived in section 2.5.4.

According to Laskovski *et al.* (2006), this preferential re-suspension of big low density particles in the channel is advantageous in gravity separation and is due to the use of longer and more narrowly spaced channels. It was also believed that this effect is limited to the aspect ratio in a way that too long and narrowly spaced channels will have diminishing efficiencies in separating particles according to their density, and narrow spaced channels were not pursued (Laskovski *et al.*, 2006; Walton, 2011:8).

It was only later that the influence of closely spaced inclined channels was investigated with the aim of improving gravity separation of fine material. Galvin *et al.* (2009) found that a set of inclined plates which were *closely spaced* helped in the suppression of the size dependence – they used a plate spacing as little as 1.77 mm. These closely spaced channels promote the following conditions (Galvin *et al.*, 2009; Galvin *et al.*, 2010c):

- a stable laminar flow;
- a lift force that is shear induced;
- an elutriation velocity directly proportional to the particle diameter near the upward facing inclined channel surface (Figure 2-7).

If the shear rate is high enough, particles can be conveyed up through the channel without experiencing significant mechanical and lubrication friction forces. The laminar flow that is developed, promotes an almost linear local velocity near the inclined surface. Thus, a particle near the inclined surface will experience a local fluidising velocity directly proportional to its diameter near the upward surface of the inclined plate (Galvin *et al.*, 2009). Figure 2-7 illustrates that the particles will experience liquid velocities proportional to

their terminal settling velocity – big particles experience a greater liquid velocity than small particles (Galvin *et al.*, 2010a).

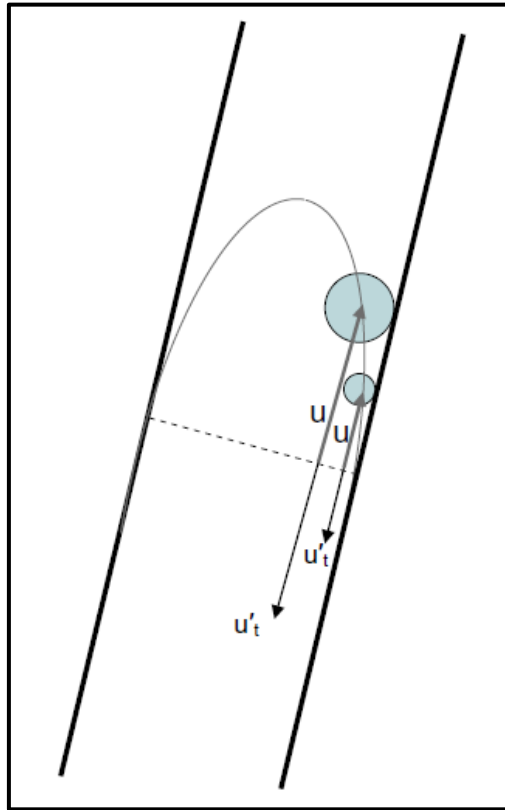


Figure 2-7: Particles experiencing fluid superficial velocities proportional to their terminal velocities in the laminar flow regime (Galvin *et al.*, 2010a)

Galvin and Liu (2011) examined the hydrodynamics within the closely spaced inclined plates theoretically. They combined equations describing the terminal velocity, fluid flow and shear induced inertial lift. These calculations agreed with experimental data obtained and confirmed the possibility to elutriate particles according to their density. Experimental work by Walton *et al.* (2010) also confirmed that more closely spaced channels improve fractionation. They tested various arrays of channels with plate spacings of 18.00, 9.00, 4.20 and 1.77 mm (Galvin & Liu, 2011; Walton *et al.*, 2010).

2.5.3 The reflux classifier and coal beneficiation

Classification by density can be applied to the beneficiation of fine coal.

A pilot-scale trial was performed to test the effectiveness of using the reflux classifier to beneficiate fine coal. A -2 mm feed was used and first-rate gravity separation was obtained. The study of Galvin *et al.* (2002) also compared the variation in D_{50} with particle size and the throughput obtained with a conventional teetered bed separator (TBS). The results generated showed that a significantly smaller variation in cut-point with particle size was observed and the solids throughput was substantially more than will be obtained in a TBS configuration (Galvin *et al.*, 2002).

This pilot test was done with normally spaced inclined channels. The cross-sectional area was $0.6 \times 0.6 \text{ m}$ and the unit was 3.5 m high. A throughput of $45 \text{ t.m}^{-2}.\text{h}^{-1}$ was obtained. A product ash of 9% and combustible recovery of 96% was achieved – 1% below the maximum attainable combustible recovery (Galvin *et al.*, 2002). Since the pilot trial, development of the technology at full-scale has been progressed by Ludowici MPE in association with the University of Newcastle (Galvin *et al.*, 2005).

The performance of the reflux classifier at full scale was then evaluated (Galvin *et al.*, 2005). It was found that the separation efficiency was equivalent to the pilot unit and that the variation in cut-point and EP values was similar. A PID controller was installed to control the cut-point by varying the discard flow rate. Significant changes had been made to the water fluidisation rate, but a remarkably small effect was noted (Galvin *et al.*, 2005). Thus, the cut-point can be varied by either changing the fluidisation rate or the discard flow rate.

With the discovery that closely spaced channels suppress the particle size dependence significantly, tests were done on fine coal beneficiation. Channel spacings of 1.77 mm and 4.20 mm were used on a feed finer than 0.5 mm and 2.0 mm respectively. Remarkable increases in separation efficiency and less variation in cut-point with particle size was obtained. The Écart Probable Moyen (e.p.m.) of 0.14 for wider channels was reduced to 0.06 over the size range 2.00 mm X 0.25 mm beneficiated in closely spaced channels (Galvin *et al.*, 2010 a).

A study was also undertaken to test the potential of the reflux classifier to beneficiate coarser coal particles. Wider channels proved adequate, and variation in cut-point with particle size proved to be at a level that is basically independent of particle size. Particle sizes greater than 2 mm beneficiated in these classifiers proved very successful and EP values of less than 0.05 were reported and improved to 0.03 as top sizes of 8 mm were approached. The required fluidisation velocity is great, however, and a nominal of 10 m/h per mm top-size is required for sufficient fluidisation water (Galvin *et al.*, 2010 b).

In coal processing one is dealing with widely spaced channels at channel spacings of 9 mm and larger and narrowly spaced systems smaller than 9 mm (Galvin *et al.*, 2010c). To summarise the requirements for coal beneficiation (Galvin *et al.*, 2010 c), wider channels are used for particle sizes greater than 2 mm. The argument put forward by Laskovski *et al.* (2006) holds for these wider channels. When particles smaller than 2 mm are to be beneficiated with the reflux classifier, closely spaced inclined channels are required. These narrow configurations help in suppressing the size effect experienced in gravity separations, because the flow is laminar and the inertial lift effect is developed. A normal rule of thumb for fine coal beneficiation is that the channel spacing should be a minimum of three times the particle top-size being beneficiated, usually in a spacing range between 2 mm and 6 mm (Galvin *et al.*, 2010c).

2.5.4 Fundamental and empirical description

Quantification of the throughput advantage

Callen *et al.* (2007) used a kinematic approach, where the particle capture efficiency is increased when the horizontally projected settling area increases, to determine the throughput advantage. The throughput advantage (E) for the reflux classifier can then be defined as the horizontally projected settling area of the plates divided by the cross sectional area of the fluidised section (equation 2.51). Figure 2-8 describes the variables used, where L_{ch} is the channel length; B the channel depth into the page; n_{pl} the number of channels; w the horizontal channel spacing; θ the channel angle to the horizontal plane and W the width of the fluidised section.

$$E = \frac{n_{pl} \times (wB + BL_{ch} \cos \theta)}{WB} \quad 2.51$$

For a single channel the width of the fluidised bed section (W) is the same as the horizontal width (w) for a single channel, and the normal channel spacing (z) is incorporated by substituting w in the second term with $z/\sin \theta$ in equation 2.52 to arrive at equation 2.53. Equation 2.53 holds for $0^\circ < \theta < 90^\circ$ and can be used in the case where either a single channel or an array of channels is present providing the plate thickness is very small relative to the channel spacing, because equation 2.53 does not account for plate thickness (Walton, 2011:51).

$$E = \frac{wB}{wB} + \frac{BL_{ch} \cos \theta}{wB} \quad 2.52$$

$$E = 1 + \left(\frac{L_{ch}}{z} \right) \cos \theta \sin \theta \quad 2.53$$

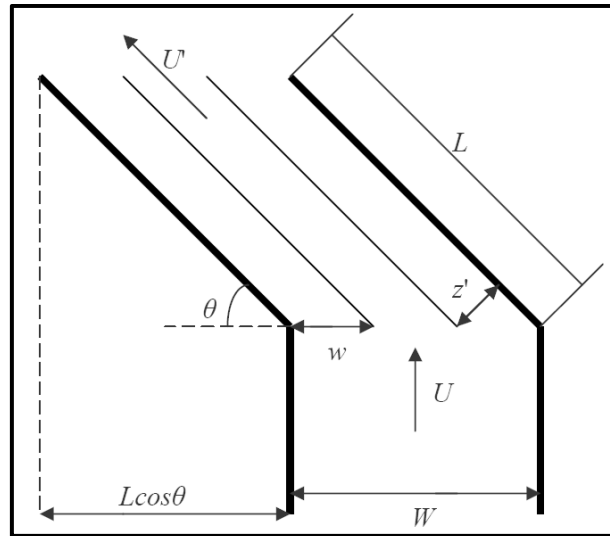


Figure 2-8: Variable descriptions for equations 2.51 to 2.53

The throughput advantage signifies the segregation efficiency as the segregation rate increases by the factor E . It should also be noted that as the aspect ratio is increased, E would have no upper bound. It is of course unreasonable to expect equation 2.53 to hold indefinitely for the aspect ratio (Galvin *et al.*, 2010c).

The segregation efficiency

Both Callen *et al.* (2007) and Laskovski *et al.* (2006) used equation 2.53 in describing the throughput advantage. During their studies they came across a phenomenon where low density particles were preferentially re-suspended when the channel spacing was reduced as already described in section 2.5.2. Hence, both of these authors introduced the concept of segregation efficiency (η) to describe the system, which can be defined as the ratio of the particle settling velocity that would be just enough in order for the particles to escape to the overflow launder (u) to the theoretical critical terminal settling velocity ($u_{t,c}$), divided by the throughput advantage. Equation 2.54 describes this relation and u is equal to the fluid velocity (U) of the system. Substituting equation 2.53 into 2.54 produces equation 2.55.

$$\eta = \frac{U/u_{t,c}}{E} \quad 2.54$$

$$\eta = \frac{U/u_{t,c}}{1 + \left(\frac{L_{ch}}{Z}\right) \cos \theta \sin \theta} \quad 2.55$$

Elutriation velocity

Laskovski *et al.* (2006) established an empirical relationship for the segregation efficiency by using Buckingham's Π -theorem. The number of dimensionless groups required to describe the system sufficiently was determined and then applied to the obtained experimental data. All terms that had exponents less than 0.05 were omitted and the resulting expression is given in equation 2.56, where Re_t is the particle Reynolds number at terminal settling condition. Equations 2.55 and 2.56 can then be combined to produce equation 2.57.

$$\eta = \frac{1}{1 + 0.133 \cos \theta \sqrt[3]{Re_t} \left(\frac{L_{ch}}{Z}\right)} \quad 2.56$$

$$\frac{U}{u_{t,c}} = \frac{1 + \cos \theta \sin \theta \left(\frac{L_{ch}}{Z}\right)}{1 + 0.133 \cos \theta \sqrt[3]{Re_t} \left(\frac{L_{ch}}{Z}\right)} \quad 2.57$$

Zhou *et al.* (2006) then used equation 2.57 to determine the optimum angle of inclination, which was 70° to the horizontal plane when the aspect ratio is between 100 and 200. When the limit of the aspect ratio (L_{ch}/Z) is taken, as it approaches infinity, and the determined

angle of inclination substituted in equation 2.57, one would arrive at equation 2.58 (Galvin *et al.*, 2010c).

$$\frac{U}{u_{t,c}} = \frac{7.5}{\sqrt[3]{Re_t}} \quad 2.58$$

The required fluid velocity to carry a particle through an infinitely long array of inclined channels can be determined with equation 2.58. Since equation 2.58 is independent of the spacing of the channel, it was experimentally verified for a broad range of plate spacings (Walton, 2011:56).

Narrowly spaced channels

The kinematic elutriation model from Laskovski *et al.* (2006) assumed that when a particle has settled onto an upward facing plate, it would slide down to the fluidised section below and will not have the chance of being elutriated through the overflow. In case of narrowly spaced channels it was noted that this is not the case and particles can be re-suspended due to shear induced lift forces (Walton, 2011:63).

Galvin *et al.* (2009) managed to derive a new expression (equation 2.60) to determine the required fluid velocity that will just be able to carry a particle to the overflow by taking the following factors into consideration:

- Wall effects determined with equation 2.37 to calculate the reduced drag coefficient $C_{D,w}$.
- Laminar flow is assumed at very narrowly spaced channels and a parabolic fluid velocity is obtained.
- The fact that bigger particles near the surface of the inclined plate will experience a larger portion of the hydraulic loading than smaller particles as described in section 2.5.2.

Galvin *et al.* (2009) also quantified the latter effect (described in in section 2.5.2 and Figure 2-7) in equation 2.59, where u_p is the fluid velocity experienced by the particle.

$$u_p = \frac{3}{2}U \frac{d_p}{z} \left(2 - \frac{d_p}{z}\right) \quad 2.59$$

$$\frac{U}{u_{t,c}} = \frac{z}{3d_p} \quad 2.60$$

Equation 2.60 is then used to calculate the fluid flow velocity required. Equations 2.61, 2.62 and 2.63 calculate the flow velocity required to just be able to elutriate a spherical particle

through very narrowly spaced channels to the overflow in the Stokes', Newtonian and intermediate particle settling regimes respectively (Walton, 2011:68).

$$U = 0.019z \frac{d_p \Delta \rho g}{\mu} \quad ; Re_p < 2 \quad 2.61$$

$$U = 0.617z \sqrt{\frac{\Delta \rho g}{\rho_F d_p}} \quad ; 10^3 < Re_p < 10^5 \quad 2.62$$

$$U = 0.0479z \frac{d_p^{1/7} \Delta \rho^{5/7} g^{5/7}}{\rho_F^{2/7} \mu^{3/7}} \quad ; 2 < Re_p < 10^3 \quad 2.63$$

It is also noted how the particle size dependence is reduced when looking at equations 2.61, 2.62 and 2.63. The most significant effect is noted in the intermediate settling regime where the particle size dependence is reduced from $d_p^{8/7}$ to $d_p^{1/7}$ (Walton, 2011:68).

It is believed that the particle settling regime is just outside the bounds for the Stokes' settling regime and Walton *et al.* 2010 have shown that equation 2.63 can be successfully applied in determining the fluid flow velocity required.

2.5.5 Related and future research

Gas fluidised bed

When inclined plates are introduced to a gas fluidised bed, a two-fold positive effect is obtained. The system ensures the capture of the finest particles, which reduces the amount of fines lost and the amount of dust scrubbing cyclones or filters needed. This is especially helpful in the pharmaceutical industry. Secondly, classification by size can be achieved at a sharper separation (Callen *et al.*, 2007).

According to Macpherson *et al.* (2010), this process can also be implemented to beneficiate coal on the basis of density when sand is used as dense medium and the equipment is subjected to vibration. A study was conducted on tracer density in a sand dense medium. The system was fluidised with air and two vibration settings were tested. The process was able to beneficiate coal in the size range of -8 +1 mm. It was also noted that the slip velocities were proportional to particle diameter, and that dense and low-density particles had negative and positive slip velocities respectively (Macpherson *et al.*, 2010).

An extension was done on the study conducted in Macpherson *et al.* (2010). Here it was concluded that on varying the underflow rate the density cut-point could be varied and a wide density cut-point could be achieved in the air-sand vibratory fluidised bed. E.p.m.

values of 0.06 were also achieved. The addition of the inclined plates improved the stability of the gas fluidised system and minimised effects caused by upsets in the feed. By increasing the sand underflow rate, a decrease in cut-point was achieved, and an increase in fluidisation air, increased the density cut-point. A significant decrease in the size dependence was also noted (Macpherson *et al.*, 2011).

Split fluidisation

A modification to the water-based reflux classifier has been noted to improve the separation ability according to density. This modification is called split fluidisation. A portion of the fluidisation water is introduced to the side of the fluidising section, while the rest is fluidised from the bottom. This ensures that less mixing occurs in the bottom of the fluidising section, and a higher quality underflow stream is obtained. This in turn improves the separation efficiency to a great extent (Doroodchi *et al.*, 2006).

CHAPTER 3 - Experimental

3.1 The RC105 at the North-West University

This section explains the design methodology and commissioning of the experimental unit (the laboratory scale reflux classifier – referred to as the RC105) depicted in Figure 3-1, and also describes the operation of the unit. The unit was built in the School of Chemical and Minerals Engineering workshop at the Potchefstroom campus of the North-West University. Refer to Appendix A to view the detailed design drawings and Appendix B for detailed calculations – references to the appropriate sections within the Appendices are provided.

Unit deliverables

The objective of the study was to investigate whether the small laboratory reflux classifier could be used as a water only density fractionator. An auxiliary objective was to test the particle size dependence on the separation efficiency.

The following were considered:

- The unit would be operated both in continuous and batch mode.
- The unit would be square with a cross-sectional area of 100 X 100 mm.
- The plates section would be 1.0 m in length at an angle of 70° to the horizontal, while the plates (1.15 m in length) protruded sufficiently through the inclined section.
- The rule of thumb for plate spacing was used, and a 3.5 mm spacing was used to process a feed with a top size of 1180 µm.
- It was essential to recycle the fluidisation water, and that no fines would accumulate within the system.
- The coal would also be fed continuously by slurry.
- The option to run fractionation tests and separating test was available.

Material specification

The main body of the unit was constructed using clear 10 mm thick PVC sheets. Clear PVC has the following advantages:

- One can observe the interactions within the unit.
- PVC is easily cut and milled without cracking or shattering the sheets.
- PVC is much more durable, where Perspex is extremely brittle.
- PVC can be cemented together very effectively and firmly by using Tangit PVC-U Cement.

When deciding on the material for the plates, factors such as strength, weight, cost and corrosion were taken into account. In the end weight and corrosion resistance outweighed the other factors, and 0.9 mm thick aluminium plates were used.

The plates were fixed in 5 mm deep slots cut into the sides of the PVC sheets, and the unsupported distance stretched only 105 mm, reducing possible bending of the plates.

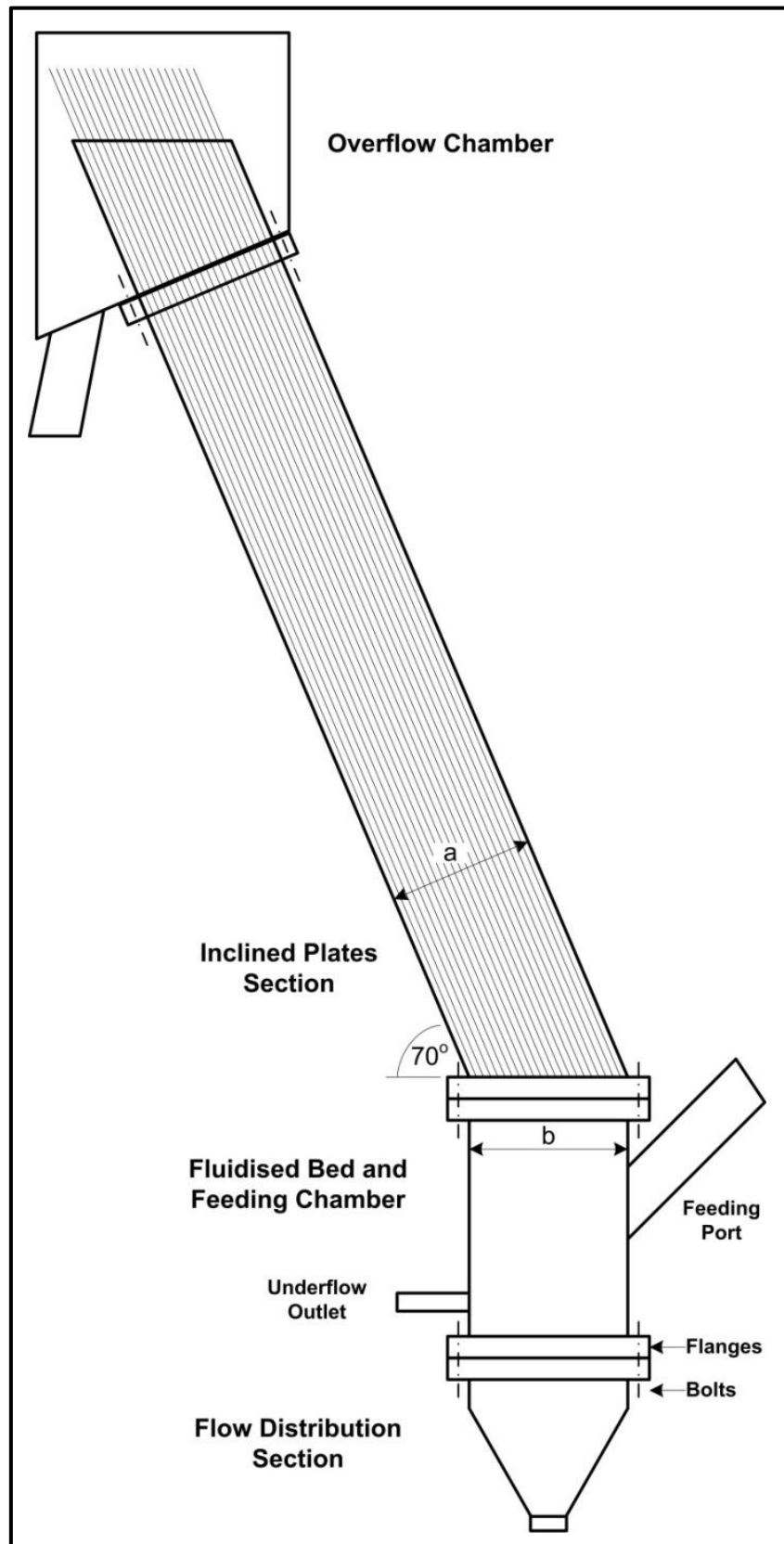


Figure 3-1: Break-down of the RC105 parts

3.1.1 Designing and manufacturing of the RC105

The laboratory reflux classifier unit was designated as the RC105, because its inner open cross sectional area was 0.105 m X 0.105 m. The RC105 consisted of four parts which were bolted together by flanges. Gaskets were placed between the flanges to prevent water from seeping through. Detailed designs of the parts and units are given in Appendix A. These parts are displayed in Figure 3-1 and they are:

- The inclined plates section
- The fluidised bed and feeding chamber
- The overflow chamber
- The flow distribution section

The inclined plates section

The channels within the capsules were designed to be 1.0 m in length, at an angle of 70° to the horizontal. To manufacture the PVC body of a capsule, a rectangular unit of 1.25 m in length was built, and the ends cut to 1 m at 70° (this is displayed in Figure 3-2 and Figure A-2). Before the PVC sheets were glued together, 5 mm deep slots were cut into the plate supporting sides with a 1 mm thick slot cutter (Figure A-1). Flanges were glued to both ends of the unit (Figure 3-1).

While the open space distance across the plate within the capsule was 105 mm, the plate width was 115 mm, designed to slide into the slots. The 5 mm deep slots extended down the whole length of the capsule. The plate length was kept at 1.15 m so that the plates protruded from the capsule by 15 cm, which is also depicted in Figure 3-1. This was to ensure that the overflow discharged through the side openings between the plates.

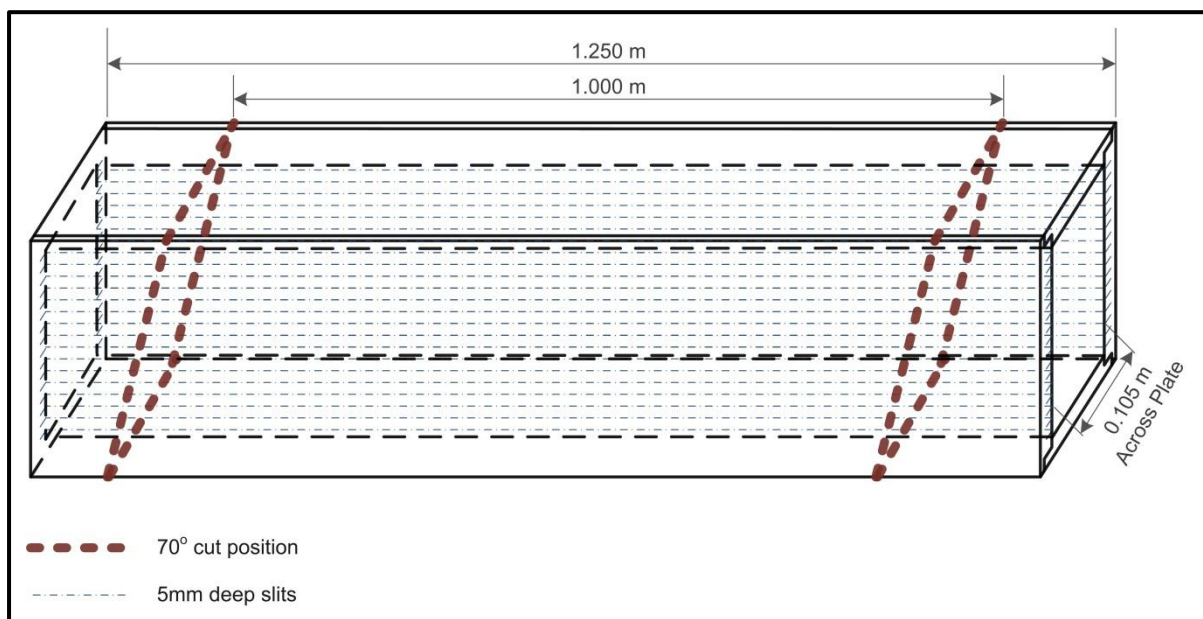


Figure 3-2: Manufacturing of the capsule housing the plates

Figure A-3 shows the detailed drawing of the completed plates section without the plates, and Figure 3-3 shows a 3D model drawing.

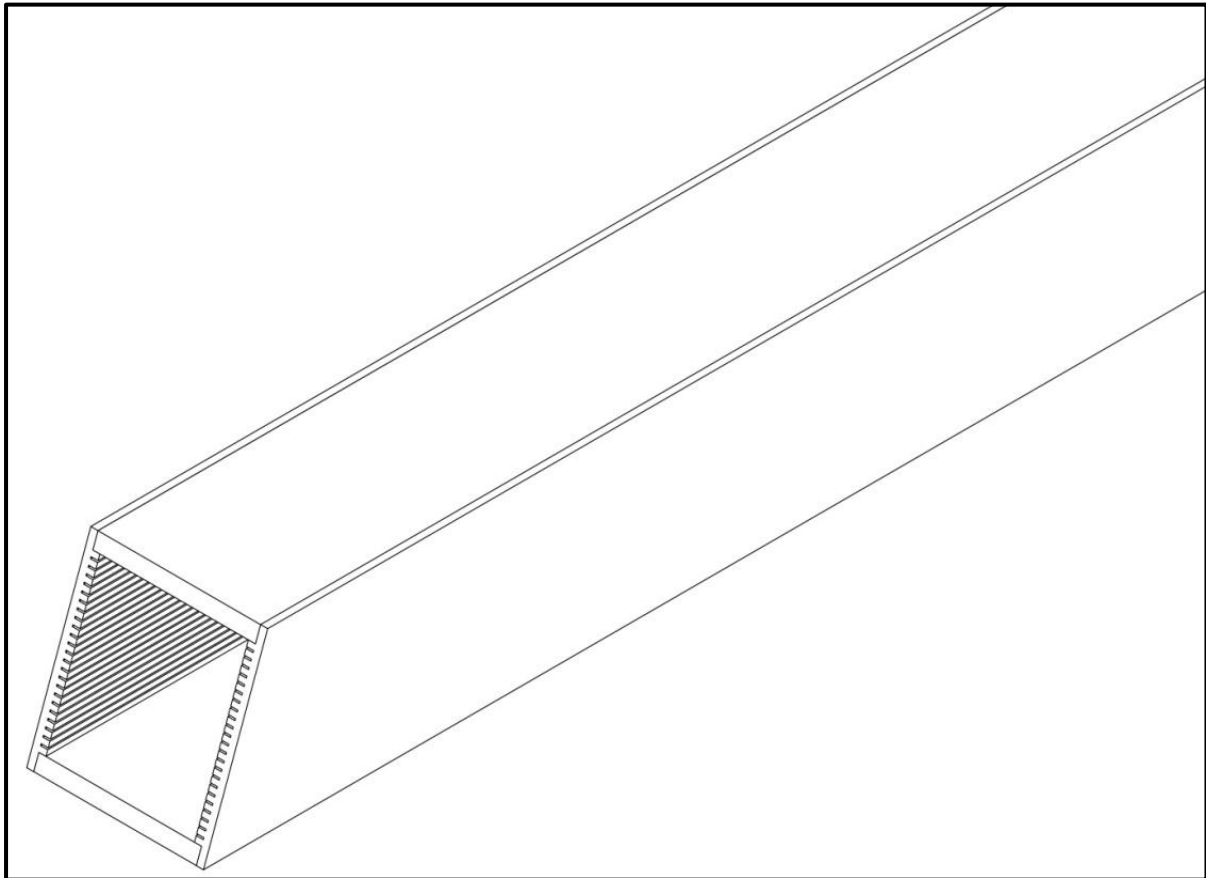


Figure 3-3: 3D model drawing of the 3.5 mm channel capsule without the plates inserted

The fluidised bed and feeding chamber

Due to the following factors, the diameter of the feed port and pipe was designed to be relatively large:

- coal would be fed in the form of slurry
- the feed flow rate would be varied
- fluidisation water would push up into the feed port

The underflow pipe diameter was designed to be as small as possible in order to minimise material by-pass to the underflow (in this case the underflow was the discard). Also, there would be sufficient head within the fluidised bed chamber to deliver an appropriate volumetric flow to the underflow regardless of the diameter (a higher flow velocity was expected). The inner diameters of the feeding port and underflow pipes were 50 mm (2" pipe standard) and 6.4 mm (¼" pipe standard) respectively.

The underflow outlet hole was fitted with a 6.4 mm male PVC pipe fitting. The feed port entered at a 45° angle and a 50 mm PVC pipe was glued to the hole, with a modified 50 mm

PVC female union fitting glued to the end. This was the configuration for continuous operation, while a plug was fitted here for batch operation (Figure 3-4).



Figure 3-4: Illustrating the plug and the pipe configuration that is fastened to the female PVC union fitting

Another small hole was made in the middle of the front side of the chamber where a manometer was installed to measure the fluidised bed density. It turned out that this was not successful due to very small pressure differentials, and thus no pressure measurement was available. Refer to unit operation in section 5.2 in Chapter 5 for recommendations regarding proper solids concentration control.

The cross-sectional area of the fluidised bed section was manufactured as 105 mm x 111.4 mm to facilitate the inclined plate section. The four sides were glued together including the bottom and top flanges, as depicted in Figure 3-1. The height of fluidised bed and feed chamber was 200 mm (refer to section 5.2 regarding fluidised bed section height). Refer to Figure A-4 in Appendix A for detailed design drawings, and to Figure 3-5 for a 3D model drawing.

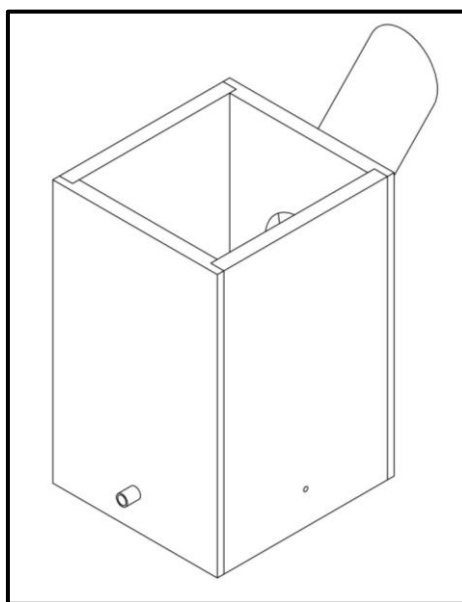


Figure 3-5: 3D model drawing of the fluidisation and feeding chamber

The overflow chamber

The overflow bucket was designed to be big enough to extend above the extended plates. The bottom part was made longer where a 6 mm hole was cut. Here a PVC pipe of 300 mm in length was glued to the overflow bucket to direct the overflow into the overflow capturing system, which is described in section 3.1.3. Figure 3-6 shows a 3D model drawing of the overflow bucket. Refer to Figure A-5 for detailed design drawings.

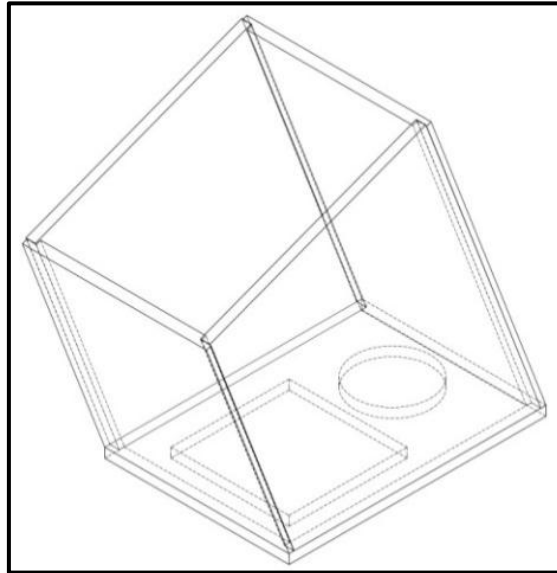


Figure 3-6: 3D model drawing of the overflow bucket

The flow distribution section

The PVC pieces for the distribution section were cut and milled to the right sizes (refer to Figure A-6 for detailed design drawings) and then glued together.

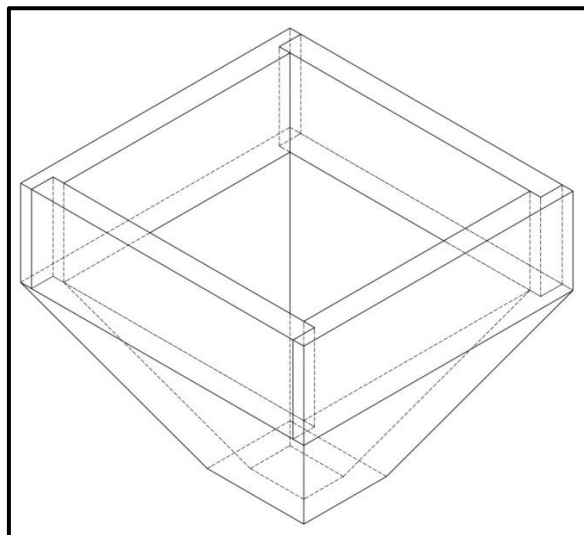


Figure 3-7: 3D model drawing of the flow distribution section

The flow distribution section tapered down to a 25 mm x 25 mm square hole at an angle of 70° . The pyramid form was glued to a 40 mm high rectangular PVC box to which the flange

was glued (Figure 3-7 shows a 3D model drawing of the flow distribution section). Glass marbles were used as a flow distribution medium (Figure 3-8: bottom right).

At the bottom a 25 mm PVC pipe fitting was glued to the flow distribution section, with an 1180 μm mesh to keep the marbles from entering the fluidisation water pipe. A 75 μm mesh, supported with a perforated plate, was placed between the flow distribution section and the feed chamber to ensure that no particles would flow back through the fluidisation water pipe. Figure 3-8 shows the perforated plate, 75 μm wire mesh and the flow distribution part filled with marbles.



Figure 3-8: Photos of the 75 μm wire mesh, perforated plate and flow distribution part filled with marbles

3.1.2 Process flow diagram and other equipment

A process flow diagram (Figure 3-9) was conceptualised to understand the process options and required equipment. It was decided that the fluidisation water would be recycled directly from the overflow into the fluidisation water tank from where it was pumped through the unit. The solids were captured, and measures were taken to prevent fines from accumulating in the fluidisation water tank. Since a peristaltic pump was too expensive, it was decided to feed the coal slurry gravitationally for the continuous feed operation.

Additional equipment included:

- Fluidisation water pump
- Fluidisation water holding tank
- Slurry holding tank
- Rotameter
- Agitator within the coal slurry tank
- Particle retention system, capturing overflow particles

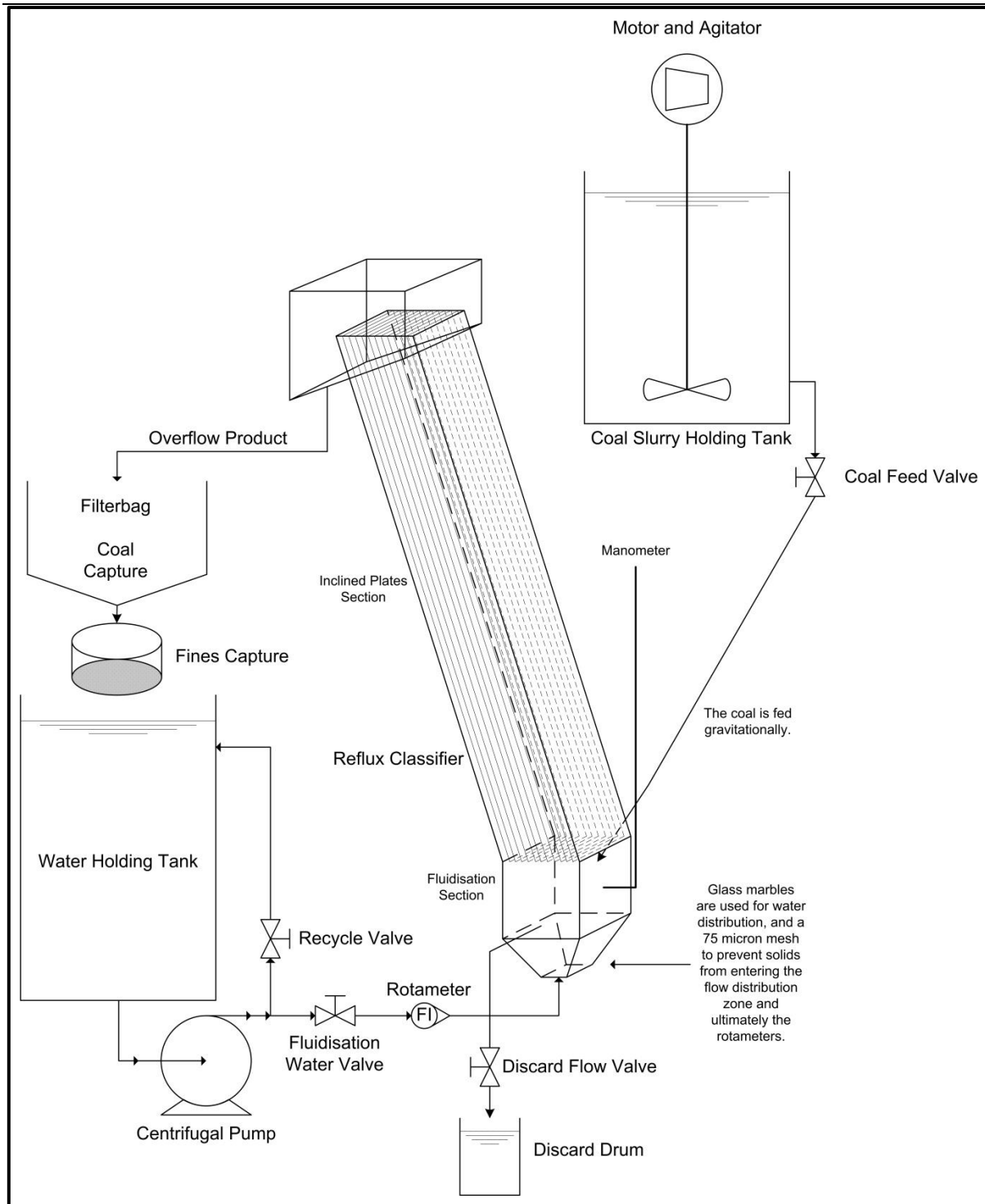


Figure 3-9: Conceptual process flow diagram for RC105

Determination of required fluidisation water

During the initial stages of the design, a throughput number obtained from the literature (Galvin *et al.*, 2002) was used to size the system. The unit used in this study was a RC600 (meaning that the inner cross-sectional area of the unit was 0.6×0.6 m), and the solids throughput that was obtained was $45 \text{ t} \cdot \text{m}^{-2} \cdot \text{h}^{-1}$. For this particular coal a 10% ash product was achieved at a yield of 80%, and the relative density cut-point was 1.6.

In order to determine the required fluidisation water volumetric flow rate a few assumptions were made (for detailed calculation, refer to section B.1 in Appendix B):

- The solids concentration in the plates section of any unit is 20%.
- The plates occupy 30% of the plates section's volume within the RC600.
- The coal has an average solids density of 1.5 (80% was lighter than 1.6).

Based on the above assumptions, the superficial fluid velocity required to transport this amount of solids through the RC600 was calculated to be 0.033 m/s. When scaling process units two requirements must be satisfied: the two units must be geometrically similar and the Reynolds number has to be the same for both unit operations. To obtain the same performance from the RC105 for the same feed coal, the superficial fluid velocity must give the same Reynolds number as the one calculated for the RC600, which was calculated to be 858, as seen in section B.1. Using the dimensions of the RC105, and also the assumption of 20% solids within the plates section, the volumetric flow rate required was 41.3 L/min, giving a superficial fluid velocity of 0.078 m/s.

To facilitate possible future work on the RC105 using higher density materials, the required pump power was doubled. The chosen pump (0.75 kW pump) could deliver a volumetric flow rate of 100 L/min. Two rotameters (high flow and low flow) were installed.

Determination of the head required

The head required was calculated for two reasons:

- Determination of the pump duty
- Determination of the coal slurry holding tank height (coal would be fed gravitationally)

There are two types of head that should be accounted for: frictional head loss, and static head loss. In order to determine the frictional head, the flow regime should be known. The literature suggested that the flow within narrowly spaced channels is laminar (section 2.5.4), and a calculation showed that at a flow rate of 50 L/min within the 3.5 mm channels, the Reynolds number was 1034. Also, before the transitional flow regime is entered ($Re_{\text{Transitional}} > 1900$ for rectangular conduits), a maximum volumetric fluidisation rate of 91.4 L/min was possible. The detailed calculations are given in section B.2 of Appendix B.

Thus, the flow was shown to be laminar within the plates over the entire experimental flow regime, and the equivalent diameter could be used in accordance with the volumetric flow equation which incorporates pressure drop (equation 2.9).

$$Q = \frac{\pi \Delta P D^4}{128 \mu L} ; Re \leq 2100 \quad 2.9$$

The equivalent diameter for the 3.5 mm channels was determined as 0.011 m, using equation 2.13. The pressure drop due to friction within one 3.5 mm channel was calculated as 0.185 kPa. The total pressure drop within the plates section with 24×3.5 mm channels was calculated as 4.44.

These calculations considered only water, since it was argued that water would be the solids conveying medium and only water would be pumped through the system. Refer to section B.3 in Appendix B for detailed calculations. All calculations were done for a volumetric flow rate of 99 L/min to incorporate a 10% over design.

When determining the pressure drop due to static head, on the other hand, the fluid relative density was used. In the fluidisation section high solids concentrations were expected, and a maximum RD of 1.3 was used. Within the plates section an RD of 1.1 was used as a maximum, although it was expected that the RD would be in the order of 1.02.

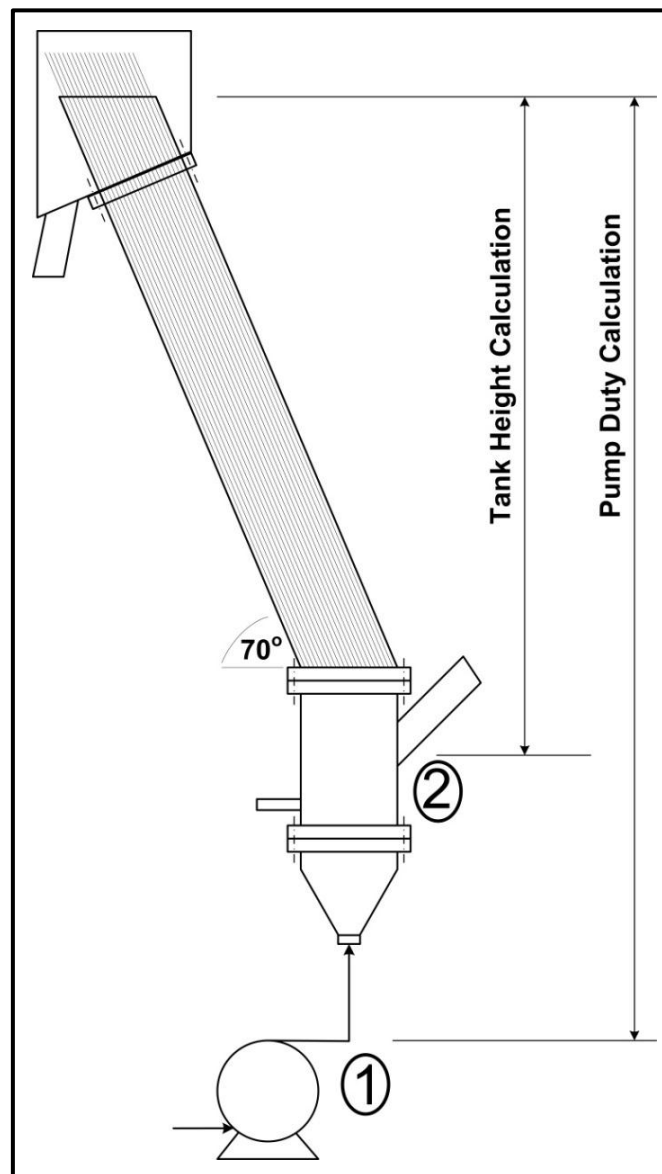


Figure 3-10: Illustrating reference points for calculating static head

Figure 3-10 shows the reference points used to calculate the static head. Reference point 1 was used to consider the pump calculations, but it was found that it was not necessary to calculate the pump duty required. During the early stages of commissioning, a centrifugal pump was selected to deliver 100 L/min flow, and this pump could deliver a head of at least 30 m at the specified flow rate.

The static head required for feeding the coal gravitationally was calculated to determine the required height of the coal slurry tank. Details of the calculation are given in section B.3 of Appendix B. The static head from reference point 2 was then calculated as 11.42 kPa. Thus, after adding the frictional head, the total head that had to be overcome at reference point 2 to ensure slurry flow into the system was 15.82 kPa for a plates section comprised of 24 × 3.5 mm channels.

In determining the height of the slurry tank required the slurry relative density was also needed. At 25% solids by volume, using 1.6 as the coal solids specific gravity, the feed slurry had a density of 1150 kg/m³. The height of the slurry holding tank was then 1.40 m from the inlet head. Detailed calculations are given in section B.3 of Appendix B. Figure 3-11 illustrates the completed unit with all the components.



Figure 3-11: The completed RC105 unit and descriptions

3.1.3 RC105 operation

The RC105 could be operated in two modes: batch or continuous. Two types of test procedures were also considered: fractionation tests with increasing fluidisation flow rates, and separation procedures at predetermined fixed flow rates in discrete separation tests. Only separating run procedures could be performed in continuous mode, while both fractionation and separating run procedures could be performed in the batch mode.

Fractionation procedure

Fractionation procedures involve fractionating the material by incrementally increasing the volumetric flow rate. Before the flow rate was adjusted for the next flow rate, the product was removed and designated as the product obtained at that specific flow rate fraction. The procedure was repeated using increasing flow rates until the entire sample was completely elutriated from the unit, and the remaining material that did not report to the overflow at the highest flow rate was designated as the discard. A product distribution could then be obtained over a range of volumetric flow rates.

Separating run procedure

Separation runs are separate runs at pre-determined flow rates. The material was charged to the unit and the system was started up. The flow rate was increased to a predetermined fixed flow rate and kept at this flow rate until no more solids were observed in the overflow. Thus, one product and one discard sample were obtained.

Batch operation

The following start-up procedure was used for each batch run performed (Figure 3-12 shows the locations of the valves and other process elements):

- The required amount of coal (about 500 g) was charged via the feed port.
- The fluidisation water tank was filled.
- The fluidisation water feed valve was closed.
- The recycle valve was set at 90% open.
- The pump was switched on.
- As the water feed valve was opened to obtain the predetermined volumetric flow rate, the recycle valve was closed by the same amount to ensure that the total open percentage between the two valves remained about 90.

When the desired volumetric flow rate was reached the process was allowed to run until no more solids were observed in the overflow (for both fractionation and separating run procedures). In the case of separating run procedures the process had then ended and the shutdown procedure could be followed, while for fractionation procedures the volumetric flow rate was increased at this stage – the process was ended when no solids were observed in the overflow at the maximum flow rate.

The shut-down procedure was as follows:

- The discard valve was opened to capture the discard.
- The fluidisation section was finally flushed until all of the discard particles were removed.
- The pump was switched off.

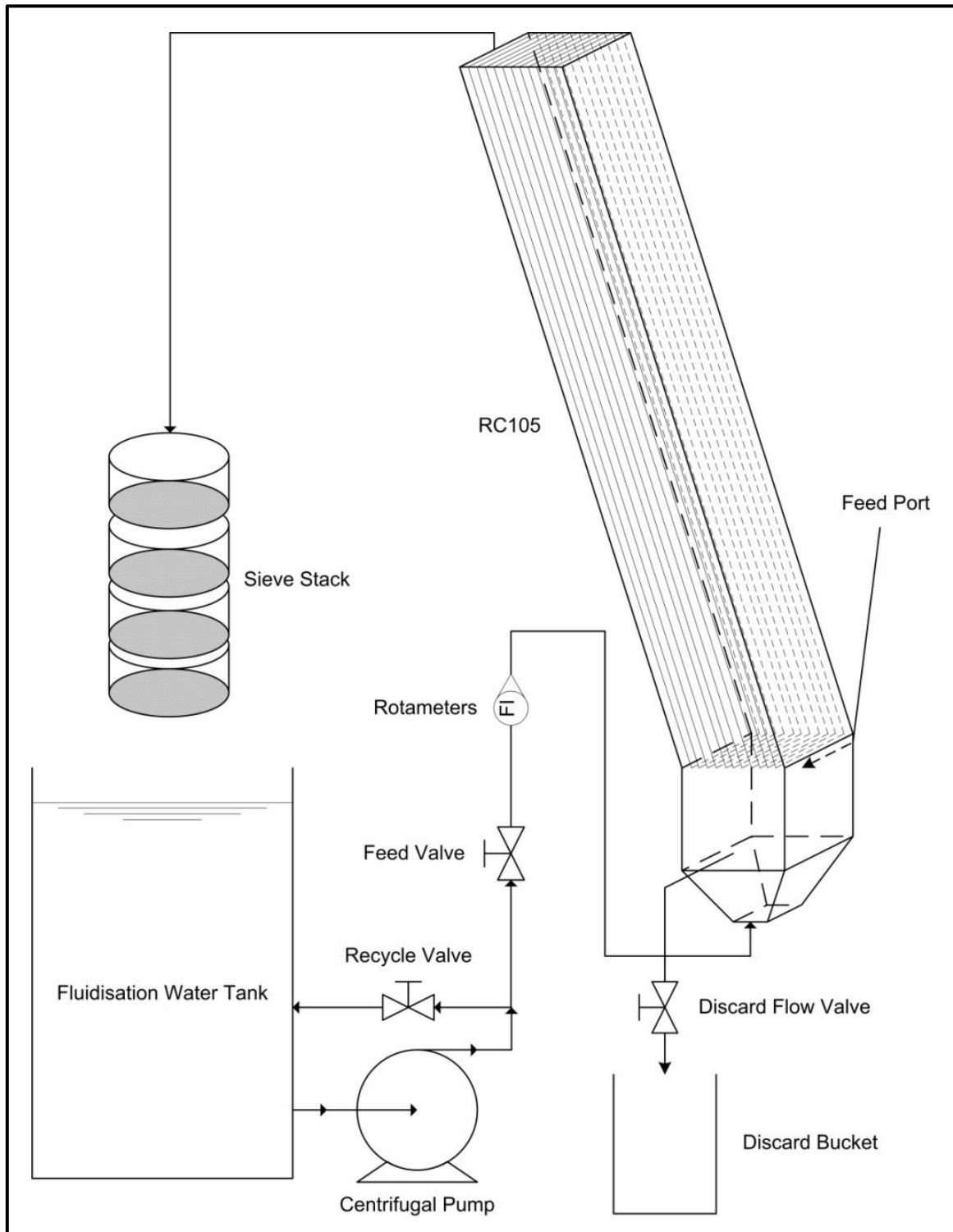


Figure 3-12: Batch mode operation PFD

Experimental

The overflow particles were separated from the water using a sieve stack, and the overflow water was recycled to the fluidisation water tank. The sieves were stacked as follows from bottom to top:

- 45 μm standard
- 75 μm wet
- 850 μm wet
- 2400 μm standard

The standard sieves were the normal 50 mm deep sieves, while the wet sieves were the 250 mm deep sieves. The 2400 μm sieve was used as a safeguard to ensure that the impact of the water did not damage the other sieves. Figure 3-13 shows the concept, where the 75 μm and 2400 μm sieves are visible.



Figure 3-13: Batch mode overflow configuration

The sieve stack split the coal overflow into two fractions: the +75 μm -850 μm and +850 μm fractions. This measure ensured that the 75 μm sieve was not overloaded with material and that water could freely run through. The 45 μm sieve prevented any generated fines from entering the water tank and hence the water circulating system.

During fractionation procedures the fluidisation water flow was diverted to recycle when no more solids were observed in the overflow, and the sieves' contents were transferred to a bucket. The sieves stack was returned to its position on the fluidisation water tank, and the flow increased to the next fluidisation flow.

For the tracer-only tests, the tracers were collected on the 850 μm sieve.

3.2 Experimental Details and Method

3.2.1 Material used

To compare the results generated in this study to previous pilot tests using a RC300 on various South African coals, the same feed material was selected as for the pilot tests. Two coals from different coal fields (coal A and coal B) were used, and the particle size ranges for the pilot unit feed samples were -1400 +500 μm and -1400 +0 μm , as is.

For the tracer tests, 1 mm cubic magnetic tracers were acquired from Tenova Minerals and Mining. The tracer specific gravities ranged from 1.2 to 2.0 in increments of 0.1. The amount of tracers charged for each test was 20 per SG class, and 190 tracers were charged per test.

3.2.2 Equipment

The equipment used during this study was the RC105 (see the details in section 3.1). Other equipment that was used included the following:

- Gyratory crusher
- Sieves stack (dry and wet)
- Sample splitter
- Vacuum oven (drying of material at 100°C)
- Proximate TGA (% ash and moisture)
- IKA C5000 Bomb Calorimeter (calorific value)
- Quanta Chrome Stereo Pycnometer (sample density deduction)

Comminution and screening

The size range of interest was -1180 μm and +75 μm (all of the runs were done using this particle size distribution), but the material as received was in the size range -1400 +0 μm . The +1180 μm particles were crushed in the rotary crusher. Dry sieves were used to screen larger particles out for re-crushing and to do PSD analyses on dry material, while the wet sieves (250 mm deep sieves) were used for de-sliming the feed or to do PSD analyses on wet material. Section 3.1.3 explains how the sieves were used during RC105 operation.

Sample splitting

Sample splitting forms a vital part of the study, because it allows one to obtain a smaller representative sample of the whole. Figure 3-14 shows an illustration of the rotary sample splitter that was used during this study, referred to as the spinning riffler. This type of sample divider has been shown to have the least sampling error, and introduce the least operator bias (Allen and Khan, 1970; Allen, 2003:49,230).

The across-cut method was used to cut samples into smaller portions – this method introduces the least bias. Regardless of the amount that was needed, at least two opposing containers had to be either removed or retained. These options are illustrated in Figure 3-15, where the red signifies discarded/removed samples and green the retained/selected samples. If the sample obtained was still too large, it was cut again.



Figure 3-14: Photo of the spinning riffler sample divider

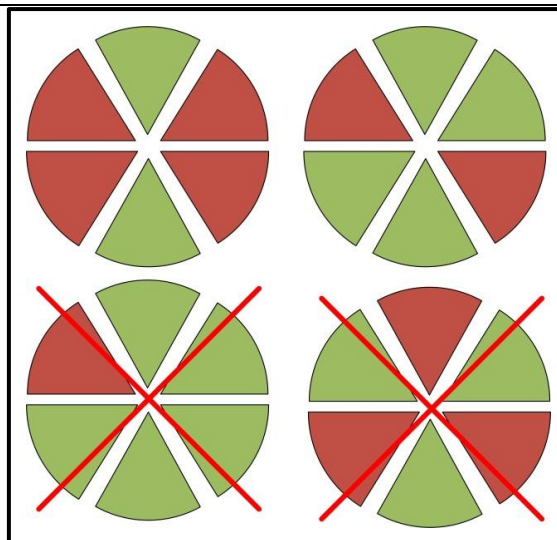


Figure 3-15: Correct and incorrect cuts performed with the sample cutter

3.2.3 The float-sink setup

The float-sink setup comprised a bank of four separating funnels and a Büchner funnel. Solutions of zinc chloride were made up in the SG range of 1.2 to 1.7 in increments of 0.1, and two solutions of each SG class were made up. These were accurately tested by means of glass tube hydrometers, and the densities of the solutions were verified continuously throughout the float-sink operations. Four float-sink procedures ran parallel.

The float-sink procedure was as follows:

- Half of the correct SG solution was poured into the separating funnel first, followed by the coal sample.
- The remaining coal was washed into the separating funnel with the other half of solution.
- The coal was allowed to separate into floats and sinks.
- The sinks were drained from the separating funnel into a Büchner funnel with filter where the solution was captured within the Büchner flask.
- The filter paper containing the sinks fraction was removed.
- The floats were captured on a new filter paper until the entire funnel contents were drained.
- The Büchner flask was then emptied (recovering the solution without dilution).
- The remaining floats within the separating funnel were washed out with water, also using the Büchner funnel.
- The empty separating funnel was charged again, and procedures (a) to (h) repeated.

The zinc chloride solution for the next float-sink procedure was used to rinse the coal from the filter paper when charging the coal. This procedure ensures that the zinc chloride solution is never contaminated and that the solution can be re-used. Figure 3-16 illustrates the float-sink setup comprising four separating funnels and the Büchner funnel.

Risk assessment

There are several hazards involved with using heavy liquids to fractionate ore for washability considerations. Not only is most of the heavy liquids hazardous to human beings, but disposing of the liquids is problematic due it then being hazardous to the environment too.



Figure 3-16: Float-Sink setup

3.2.4 Experiment layout

Table 3-1 summarises the particular tests performed on the various materials used during the experimental campaign. The materials used were the tracers and the two coal samples from different origins (coal A and coal B). Three types of runs were performed, and they were:

- fractionation procedures in batch mode
- separation procedures in batch mode

Fractionation procedures were performed on the tracers, coal A and coal B, and the purpose of the fractionation procedures in batch mode was:

- To investigate the possibility of using the small scale laboratory reflux classifier as a water only fractionator for coal washability analyses.
- To correlate density cut-point and volumetric flow rate.

Separation procedures were done at fixed flow rates in discrete runs, and on tracers, coal A and coal B. The purpose of the separation procedures was:

- To investigate whether the laboratory scale reflux classifier could be used to separate material based on density in *batch mode*.
- To investigate whether the density cut-point and fluidisation flow rate correlation could be used to predict the separation process.

Table 3-1 also shows the auxiliary runs that were performed, namely:

- One repeat of a separation run in batch mode.
- One separation procedure at higher initial bed density.
- Size fractionation of the float-sink analysis products of one separation procedure.
- Ash analysis of the different size fractions obtained.

The type of analysis that was performed on each of the resultant samples is explained in Table 3-1 as well. Mostly when coal was fractionated, the Quantachrome Stereo Pycnometer was used to determine the average density of the fractionated samples and ash content, while float-sink analyses were done on the overflow and underflow samples obtained from the separation procedures.

Table 3-1: Experimental run break-down

Material	Run Type	Purpose	Analyses
Tracers	Fractionation	$\rho_{50} - Q_W$ Correlation	Tracer count
	Fractionation	$\rho_{50} - Q_W$ Correlation (Coal A incl.)	Tracer count
	SR@45.0LPM	Compare to correlation	Tracer count
	SR@57.5LPM	Compare to correlation	Tracer count
	SR@67.5LPM	Compare to correlation	Tracer count
Coal A	Fractionation	$\rho_{50} - Q_W$ Correlation (Tracers incl.)	Pycnometer
	SR@40.0LPM	Compare to correlation	Sink-Float, %Ash
Coal B	Fractionation	$\rho_{50} - Q_W$ Correlation	Pycnometer, %Ash
	SR@32.5LPM	Compare to correlation	Sink-Float
	SR@42.5LPM	Compare to correlation	Sink-Float
	SR@60.0LPM	Compare to correlation	Sink-Float
	SR@32.5LPM	Repetition	Sink-Float
	SR@32.5LPM	Repeat at higher $\rho_{B,i}$	Sink-Float
	PSD's on	SR@32.5LPM sink-floats products	%Ash

The abbreviation SR signifies a separation run

ρ_{50} – density cut-point

$\rho_{B,i}$ – initial bed density

Q_W – volumetric flow rate of fluidisation water

Experiments involving the tracers

Tracer tests were done on the RC105 to construct cumulative partition curves. The overflow product of one fractionation test was combined with the preceding fractionation products to obtain a cumulative overflow product at a specific flow rate. The tracers left in the underflow at each fractionation run were determined by difference.

From these curves a correlation between density cut-point and volumetric flow rate could be obtained. Discrete runs followed the fractionation procedures and three flow rates were identified. The objective was to predict a certain density cut-point using the correlation, and a low (45.0 LPM), medium (57.5 LPM) and high (67.5 LPM) flow rate were chosen to fit the correlation that was obtained through the fractionation procedure.

To investigate how the reflux classification performance would be influenced by a wider particle size range and higher solids concentration, tracers were charged along with a coal sample for a fractionation run – coal A was used for this experiment. The rationale behind this was that the tracers were limited to no more than 180 particles and at a very narrow particle size range (+900 μm -1 100 μm), while coal runs are in the order of many particles at particle size ranges much wider (in the order of more than 10 times the bottom size). The tracer only condition was an infinite dilute system, while the coal plus tracer runs were considered multiple particle systems where particle-particle interactions were prominent.

Coal fractionation

Fractionation procedures were done on coal A and coal B to correlate density cut-point and volumetric flow rate. The overflow samples taken at each flow rate were dewatered and dried, and the mass determined. The average densities of these samples were determined using the pycnometer, while the ash content of the fractionated product samples of coal B was determined using a standard TGA method.

Coal separating procedure

Three flow rates (32.5 LPM, 42.5 LPM and 60 LPM) were chosen for the discrete runs to verify the predicted product yield and cut-point for coal B. These flow rates were chosen to achieve certain significant yields ($\approx 50\%$, $\approx 70\%$ and $\approx 85\%$) which were read off from the populated graph from the fractionation run done on coal B. The test at one flow rate was repeated, and an extra run was done at the same flow rate using a different initial solids concentration – 32.5 LPM was chosen, because it produced a good solids split of near 50%. One separation run was performed on coal A at 40 L/min for the purpose of producing a 10% ash product.

Float and sink analyses

Feed samples of both coal A and coal B were analysed by means of float and sink analyses as described in section 3.2.3. Not only is it good practice to obtain washability data for the feed samples, but these curves were also compared to fractionation data. An auxiliary outcome

was to prove that a laboratory scale reflux classifier can be used to produce washability data through fractionation.

The overflow and underflow samples from the discrete runs were also subjected to float-sink analyses to determine the partition curves for each of the runs. The bulk of the semi-batch runs overflow and underflow samples were treated in the mineral laboratory, and the five batch mode discrete runs product and discard samples were sent to Witlab Laboratories in Witbank for float-sink analysis.

Particle size

Particle size played an integral role in this study since the performance of the reflux classifier is claimed to be less dependent on the size of the particles. Thus, a PSD analysis was done on the product and discard samples of one of the discrete runs performed in batch mode on coal B – note that this was only an auxiliary test, and not the point of the main investigation. The samples were split into seven density fractions at Witlab, and each density fraction was split into four size ranges:

- -355 μm
- +355 -600 μm
- +600 -850 μm
- +850 -1180 μm

The objective was to determine where small heavy particles and big light particles reported to. The density of a particle is also related to its mineral content, and ash analyses were done on each fraction (provided sufficient sample mass could be recovered).

3.2.5 Sample analysis

After the samples were obtained from each run, they were dewatered using a Büchner funnel and dried in an oven at 100°C; the dry samples were then weighed. A complete mass balance was needed for each batch operation to reconstitute the feed. After the samples were weighed, they were subjected to float-sink analysis. Coal fractionation product samples were analysed for average density.

In some cases the recovered samples were too large, and they were split by means of the sample splitter. The samples that had undergone float-sink analyses were properly washed and rinsed to remove the excess zinc chloride solution. Each of these density fraction samples were then analysed for ash and calorific value.

CHAPTER 4 - Results and Discussion

The main experiments of the experimental stage were preceded by preliminary test work. The preliminary test work mainly consisted of extensive tracer tests and test work involving coal A. The main experiments included batch test work involving coal B. The purpose of the semi-batch work was to simulate bed density build-up within the fluidised section.

4.1 Fractionation Procedures

To test the relationship between the density cut-point and volumetric flow rate (and, hence, the superficial fluid velocity between the channels), correlation runs were done. Initially, density tracers in water were used. Coal/tracer mixtures, and later coal only fractionating tests were also done. Coal/tracer mixtures tests involved coal A only, while coal only tests involved both coal A and coal B.

4.1.1 Tracers

The fractionation test was performed in increments of 2.5 LPM from 42.5 LPM to 70 LPM and in increments of 5.0 LPM from 70 LPM to 90 LPM (15 flow fractions were obtained). The partition curves for the 50 LPM and 75 LPM flow fractions are displayed in Figure 4-1 and Figure 4-2 respectively as examples of the partition curves (the entire set of partition curves is provided in section D.1 of Appendix D. The last three flow rates served no real purpose and only flow rates of up to 75 LPM have been included in the discussion. This is due to the fact the end tails for these partition curves did not reach partition numbers of 0.50 at an SG of 2.0.

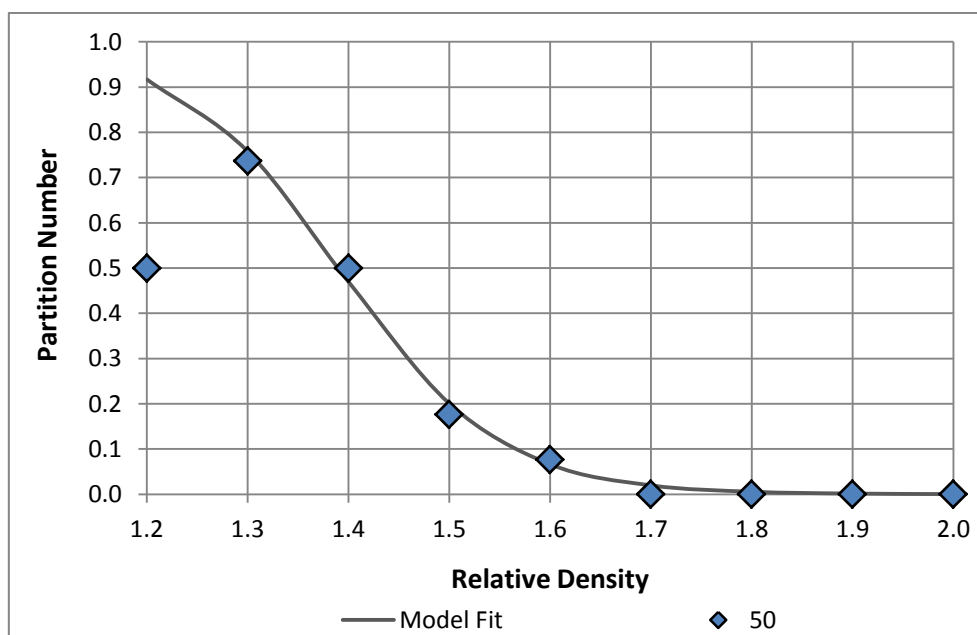


Figure 4-1: Partition curve at 50.0 LPM: Fractionation (Tracers)

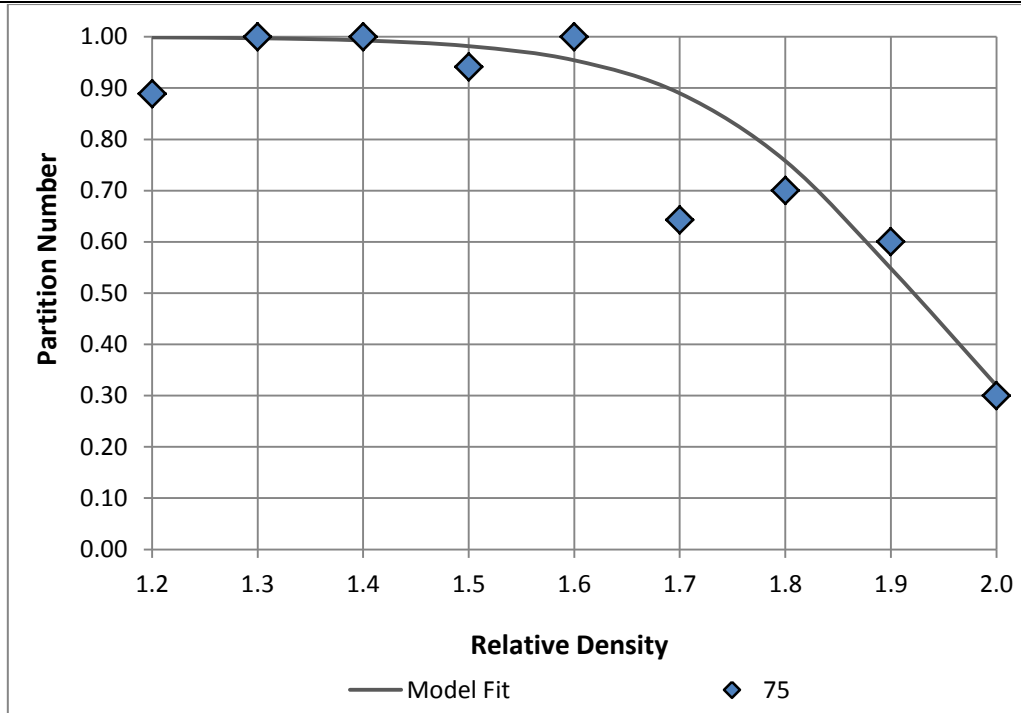


Figure 4-2: Partition curve at 75.0 LPM: Fractionation (Tracers)

The tracers recovered in the overflow after each flow rate run were mathematically combined with the previous run, yielding cumulative recovery data. The tracers in the underflow were determined by the difference between the number of feed tracers and the accumulated number of overflow tracers. The partition numbers for each density class could then be obtained for a specific flow rate tested within the fractionation procedure.

A model proposed by Wills and Napier-Munn (Wills and Napier-Munn, 2006:264) was used to determine the density cut-point and e.p.m. of a particular separation stage (equation 4.1). These were determined by minimising the sum of the squared errors using Solver in MS Excel, where the error was measured by the difference in the calculated partition number and the experimental partition number. Both the experimental partition numbers and the modelled partition curve were plotted on Figures 4.1 and 4.2 to confirm that the model fits the data.

$$P_i = 1 - \frac{1}{1 + \exp \left[\frac{\ln 3 \cdot (\rho_{50} - \rho_i)}{e.p.m} \right]} \quad 4.1$$

The density cut-point for each cumulative fractionation run was plotted against the fluidisation flow rate of the particular run in Figure 4-3, to determine the relationship of fluidisation rate and cut-point. It was found that the relationship was linear with an R^2 of 95%. This had been expected, because in dilute laminar systems Stokes' law governs particle behaviour, and the particle's terminal velocity is directly proportional to the particle density. Note that all the tracers were the same size (± 1 mm) and shape (cubes).

Table 4-1 gives the cut-point and respective e.p.m. values obtained for each flow rate, and considering the particle size competitive separation efficiency is noted. This is due to the lack of particle-particle interactions, since only a maximum of 180 tracers were present in the system. It is believed that high solids concentrations will improve on separation efficiency.

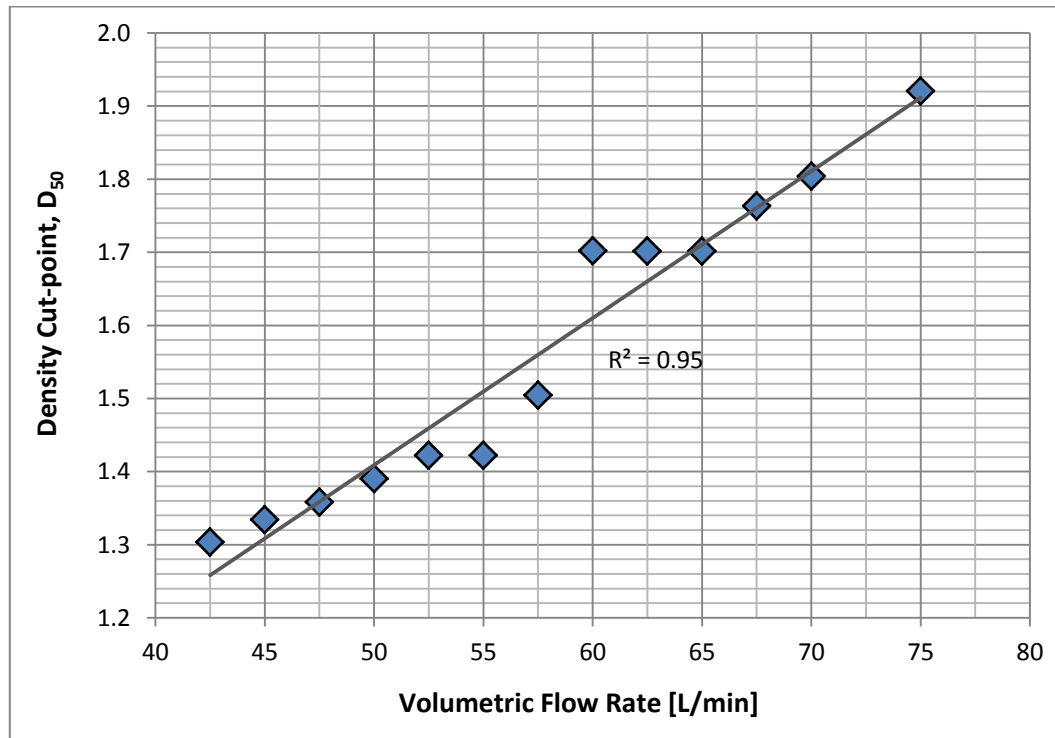


Figure 4-3: Correlation: Density cut-point and volumetric flow rate (Tracers: water only)

Table 4-1: Summary of the cut-point and e.p.m. values (Fractionation: Tracers – water only)

Flow Rate	Cut-point	e.p.m.
42.5	1.30	0.039
45.0	1.33	0.048
47.5	1.36	0.065
50.0	1.39	0.087
52.5	1.42	0.094
55.0	1.42	0.094
57.5	1.50	0.042
60.0	1.70	0.094
62.5	1.70	0.109
65.0	1.70	0.109
67.5	1.76	0.068
70.0	1.80	0.066
75.0	1.92	0.116

4.1.2 Tracers and Coal A

The solids concentration was increased by charging a sample of coal A along with tracers for a fractionation test. The same test work procedures were followed as for the tracer only tests. Partition curves for the various flow rates were generated and are given in section D.2 of Appendix D. The conditions during this fractionation procedure differed, as the new tracers were in the presence of many coal particles. Thus, the tracers were no longer subjected to Stokes' settling regime and were within a hindered settling regime in the plates section. The new relationship between fluidisation flow rate and density cut-point is compared to the relationship associated with a water and tracer only diluted system (Figure 4-4). Table 4-2 summarises the e.p.m. values and density cut-points obtained at each flow rate for the hindered settling system.

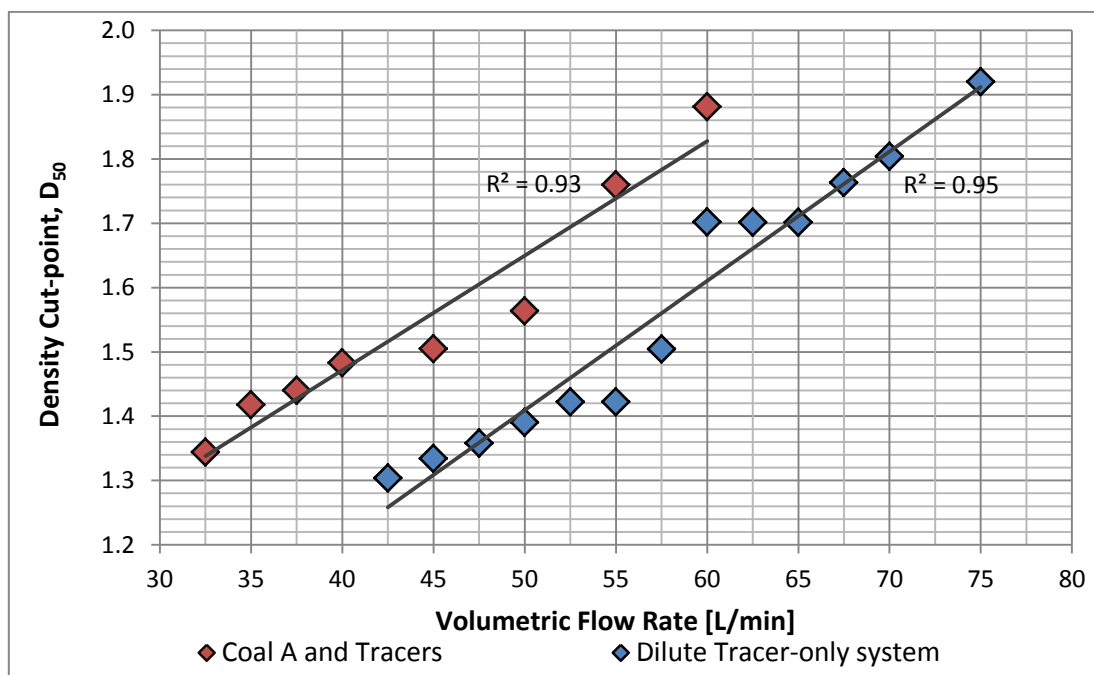


Figure 4-4: Correlation: Density cut-point and volumetric flow rate (Tracers: Hindered settling)

Table 4-2: Summary of the cut-point and e.p.m. values (Fractionation: Tracers – hindered settling)

Flow Rate	Cut-point	e.p.m.
32.5	1.34	0.050
35.0	1.42	0.018
37.5	1.44	0.023
40.0	1.48	0.033
45.0	1.50	0.036
50.0	1.56	0.059
55.0	1.76	0.060
60.0	1.88	0.035

The tracers required a lower volumetric fluidisation rate to be carried over to the overflow, because the terminal settling velocity associated with hindered settling is lower than in the case with the Stokes' settling regime. Interestingly the slope of the relationship remains the same. It is also evident that the separation efficiency has increased by the increase of solids concentration, because lower e.p.m. values are reported in Table 4-2.

4.1.3 Coal A

Typical data that is of interest when fractionation procedures are performed on coal sources are:

- The mass of coal obtained in each fraction
- The %ash of each fraction
- The average density of each fraction

This data was generated for coal A. An average density for each overflow fraction was determined by means of the Quanta Chrome Stereo Pycnometer. Float-sink analyses on each fractionated sample could not be done, due to the small amount of sample captured. The data at each flow rate is summarised in Figure 4-5, and comprises the average density, mass recovery for each flow fraction, and ultimately the cumulative %mass that reported to the overflow for a specific flow rate. Ash analyses were not done on the coal A runs, since this was a scoping test, but a full set of data is provided through coal B.

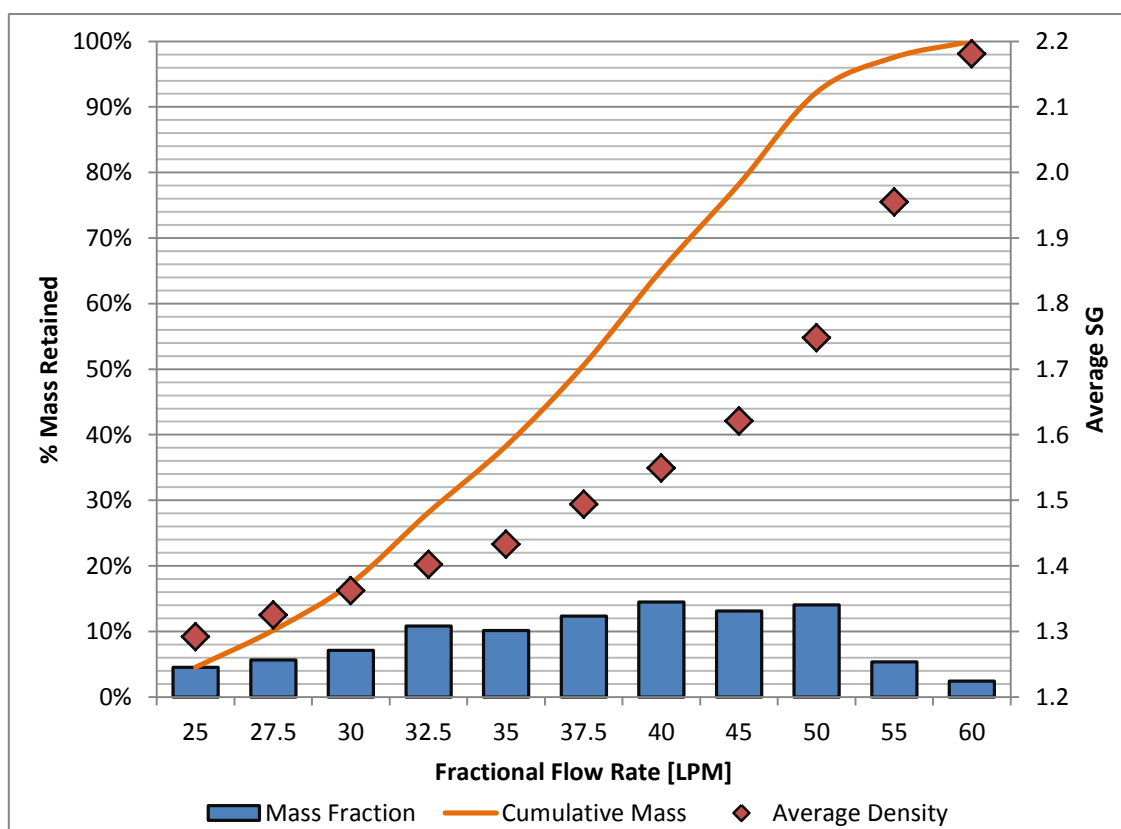


Figure 4-5: The amount of coal, average density and cumulative %mass in each flow fraction (Fractionation: Coal A)

The data in Figure 4-5 summarises each overflow sample's mass and average density. A scattered plot (Figure 4-6) correlates *average density* and fluidisation rate. To compare coal and tracer data, some mathematical manipulation is necessary to convert cut-point to average density of fractionated samples.

Sections 4.1.1 and 4.1.2 (tracer tests) report the correlation data as the actual cut-point of each flow rate obtained from cumulative data. Figure 4-5 reports the density obtained as the average density of each collected sample. Since no float-sink analyses were done to determine the actual cut-point of each fractionation instance for coal, density averages must be calculated for the tracer data. There are two ways to report this data: fractional and cumulative.

Cumulative density averages were obtained, when the data series that produced the tracer cut-point data was used. This is due to the manner in which the cut-point data was obtained. Thus, the coal fractional average densities must be converted to cumulative data as well. This is done by dividing the cumulative mass by the cumulative volume of each fraction. The solids volume was determined by dividing the fractional mass by the average density. Figure 4-6 shows the cumulative average density of each fractionation instance for tracers in a water only diluted and hindered settling system, and for coal A.

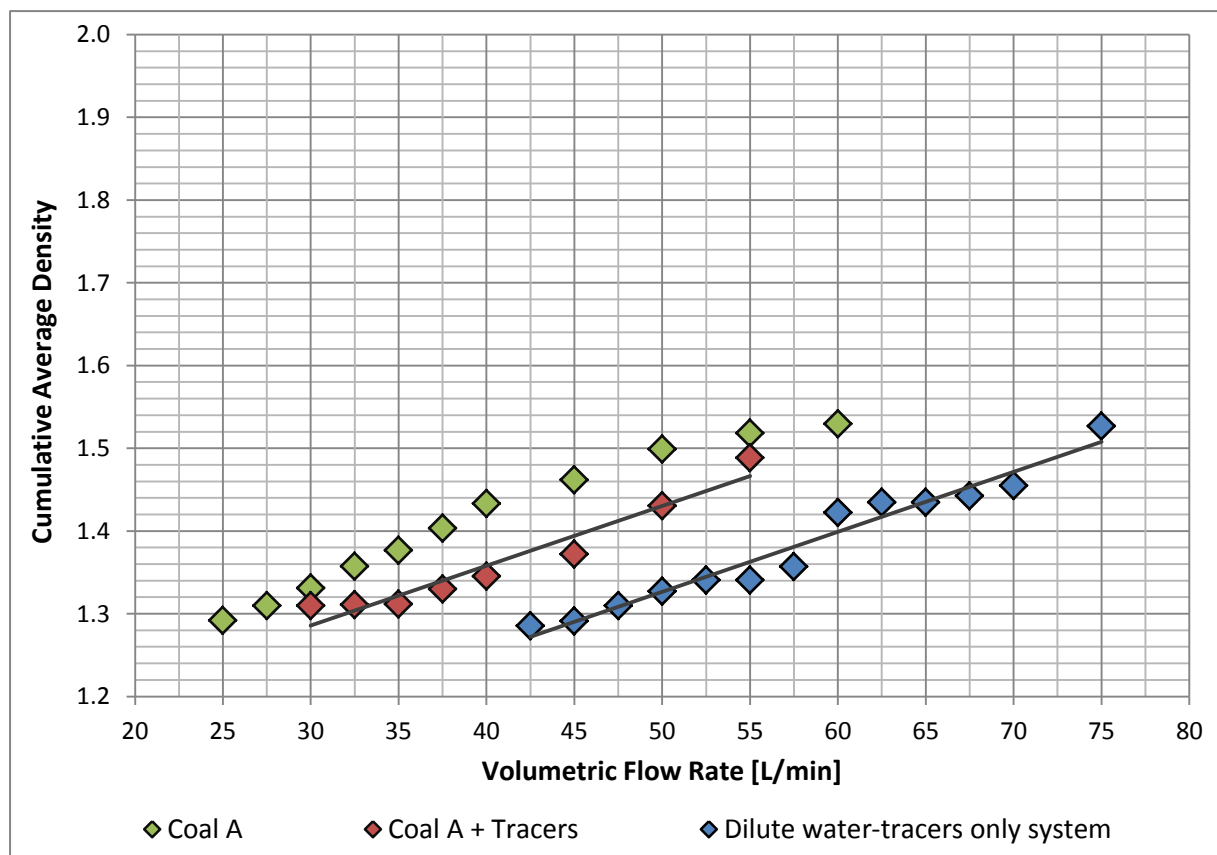


Figure 4-6: Correlation: Average density for cumulative tracer and coal A data

Figure 4-6 is not very informative from a fractionation point of view, because it gives the average density of the product sample if a separation run was done at a specific flow rate. The flattening of the cumulative average density curve is normal, because each increasing point is drawn down by the amount of lower density coal preceding the point. However, this exercise gives the averaged density of the feed sample as 1.58 when the cumulative data includes the heaviest sample fraction that remained in the system after the fractionation test was completed (SG of 2.55).

The data can also be presented in a fractional manner, where the average density of each sample collected (in the case of the tracers) is calculated by dividing the mass by the volume of the tracers that reported to the overflow for each fraction separately. The fractional averaged density for each fractionation instance for tracers in a water only diluted and hindered settling system, and for coal A are given in Figure 4-7.

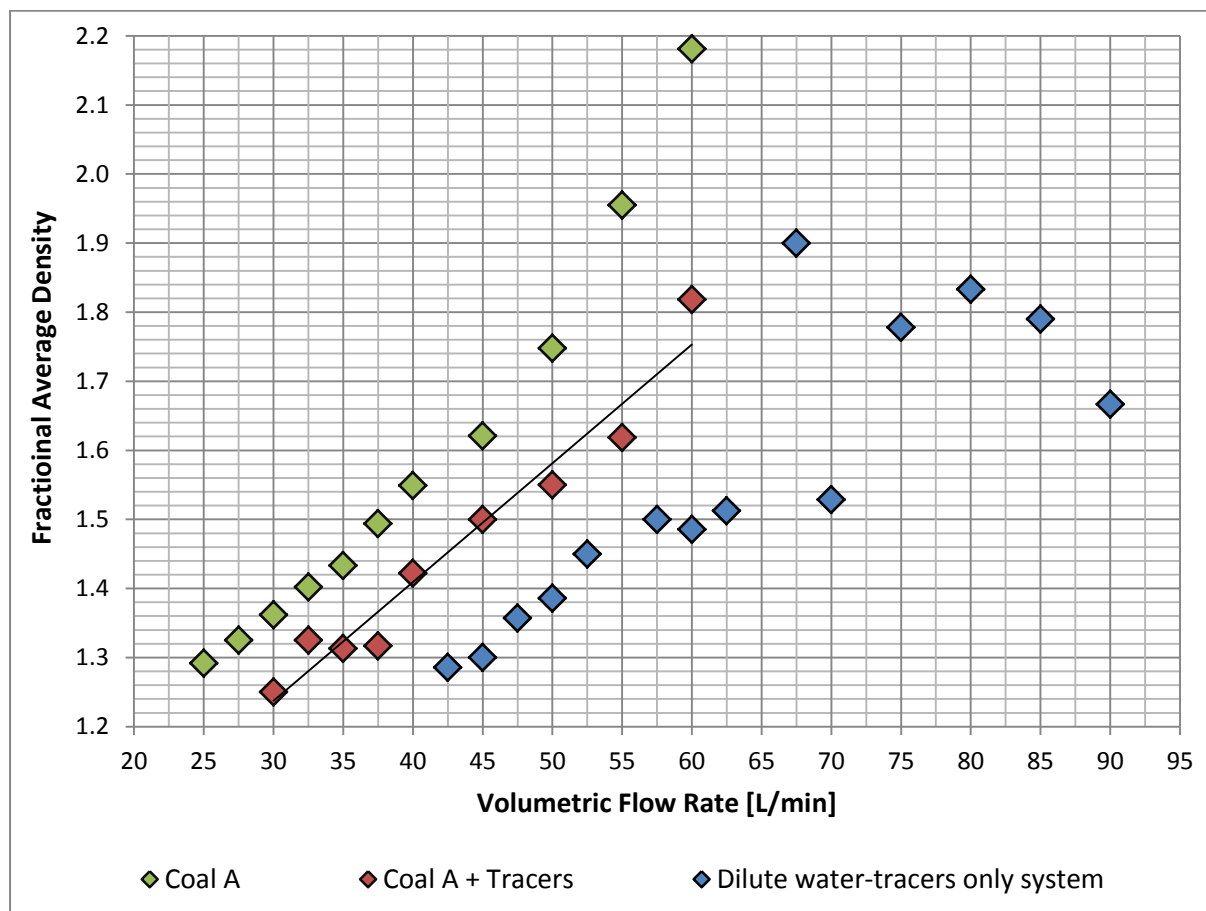


Figure 4-7: Correlation: fractional density averages for tracers in water only and hindered settling system, and for coal A

The relationship between coal and tracers is very similar when comparing Figure 4-6 and Figure 4-7. Figure 4-7 shows a linear relationship for the coal data set, parallel to the tracer data set up until 45 LPM. The fractional average density is a bit higher for the case of coal, when compared to the uniform $\pm 1000 \mu\text{m}$ cubic tracers. This is always expected in gravity

separation processes; where the combined cut-point of large light particles and small heavy particles may cut at a lower flow rate.

The tracer only data sets are scattered and skewed when presented as the fractional average density, especially the water-only system. This is due to the small amount of particles present (compared to Coal A and Coal A + tracers), and the low separation efficiency reported for this system (section 4.1.1). One misplaced particle can greatly distort the average density trend of solids captured per fraction. However, the fractional average density for the tracers in the coal system follows a smooth pattern as opposed to the scattered pattern for the tracers.

An important question arises: what is the extent to which the average density and the actual density cut-point differ? Figure 4-8 compares the average density and the density cut-point at each of the flow rates for the tracers in the hindered settling regime. This data set was chosen due to least scatter being present.

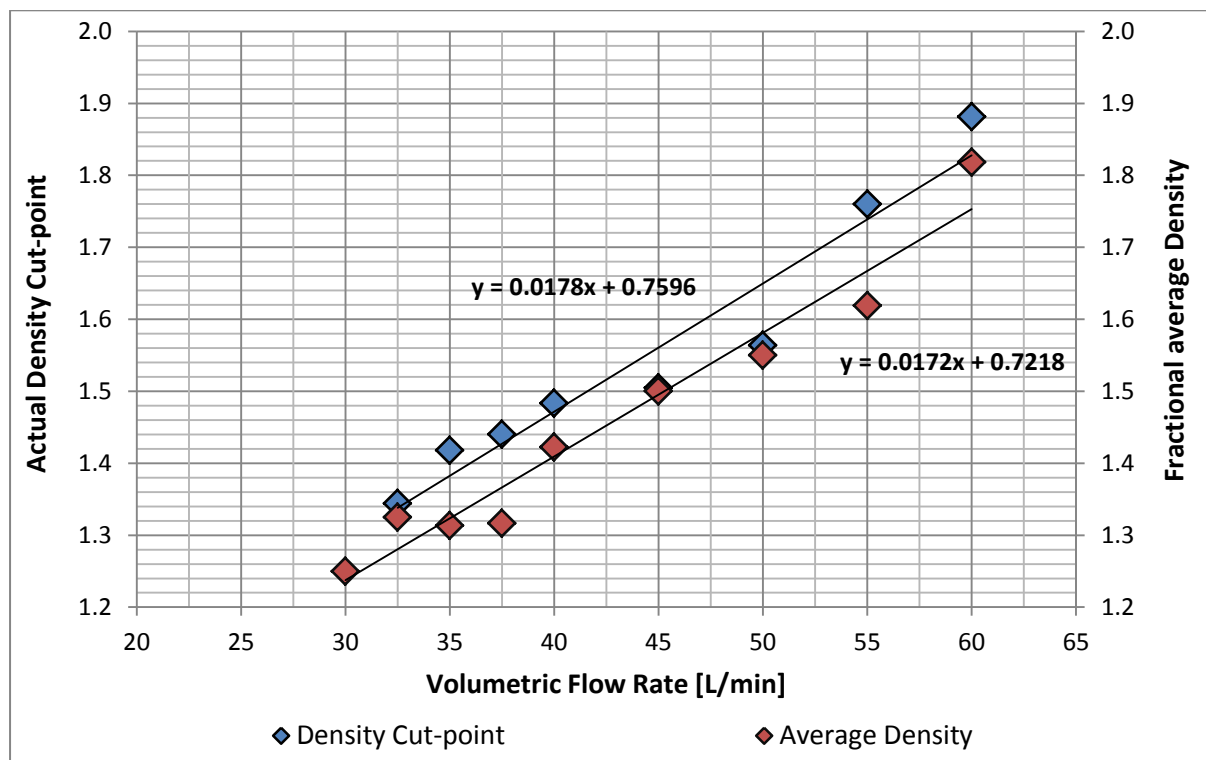


Figure 4-8: Comparing the actual density cut-point and the average density (Fractionation: Tracers – hindered settling)

It is expected that the average density will be lower than the density cut-point, because the cut-point is defined as the particles that had a 50/50 percent chance to be elutriated. Thus, particles that could not be elutriated by the preceding flow rate, will be elutriated in the current flow rate, and range from the lightest (100% chance to be elutriated) to the heaviest (particles near and slightly heavier than the cut-density). In the hindered settling regime, the difference in average density and density cut-point ranges from 0.058 to 0.070, according to

the tracer test. It is also expected that this difference will never be constant, and is dependent on the washability of the material and the separation efficiency.

4.1.4 Coal B

Cumulative %ash, average density and cumulative %mass data was generated for the fractionation test work performed on coal B and is displayed in Figure 4-9. In Figure 4-10, the fractional average density is correlated with the volumetric flow rate, and compared to the results obtained from the fractionation run performed on coal A.

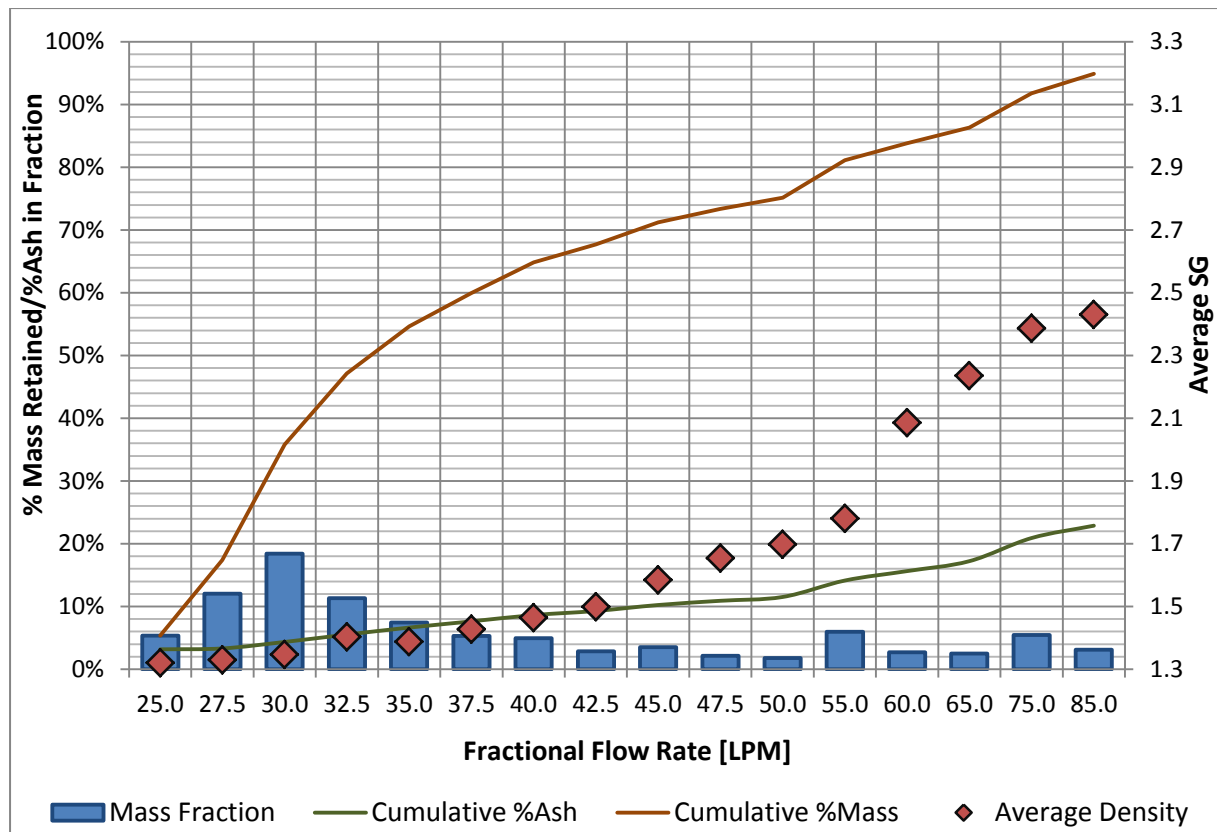


Figure 4-9: The amount of coal, average density and cumulative %mass and %ash in each flow fraction (Fractionation: Coal B)

The fractional average densities obtained yield a similar correlation to fluidisation flow rate for both coal sources. The initial trend is linear and then it breaks away from linearity. The particle size range used is also large, which shows that the density cut-point shift is relatively small.

4.1.5 Fractionation procedure – Summary

Break-away

Table 4-3 gives a summary of the conditions and particle size distributions of the material tested during the various fractionation test runs. $\rho_{B,i}$ signifies the initial bed density when particle entry to the lamellae is imminent, and $y_{s,i}$ and $y_{s,B}$ the solids concentration (by

volume) within the fluidised section initially and when the break-away (as observed in Figure 4-10) from linearity occurred respectively.

The average SG of each of the coal sources (1.58 for coal A and 1.54 for coal B) was calculated by mass balance. The known volume of the fluidised section is 2.39 L, and 749 grams of coal A and 588 grams of coal B were present within this volume at start-up. From this information it was a triviality to calculate the initial bed density and solids concentration.

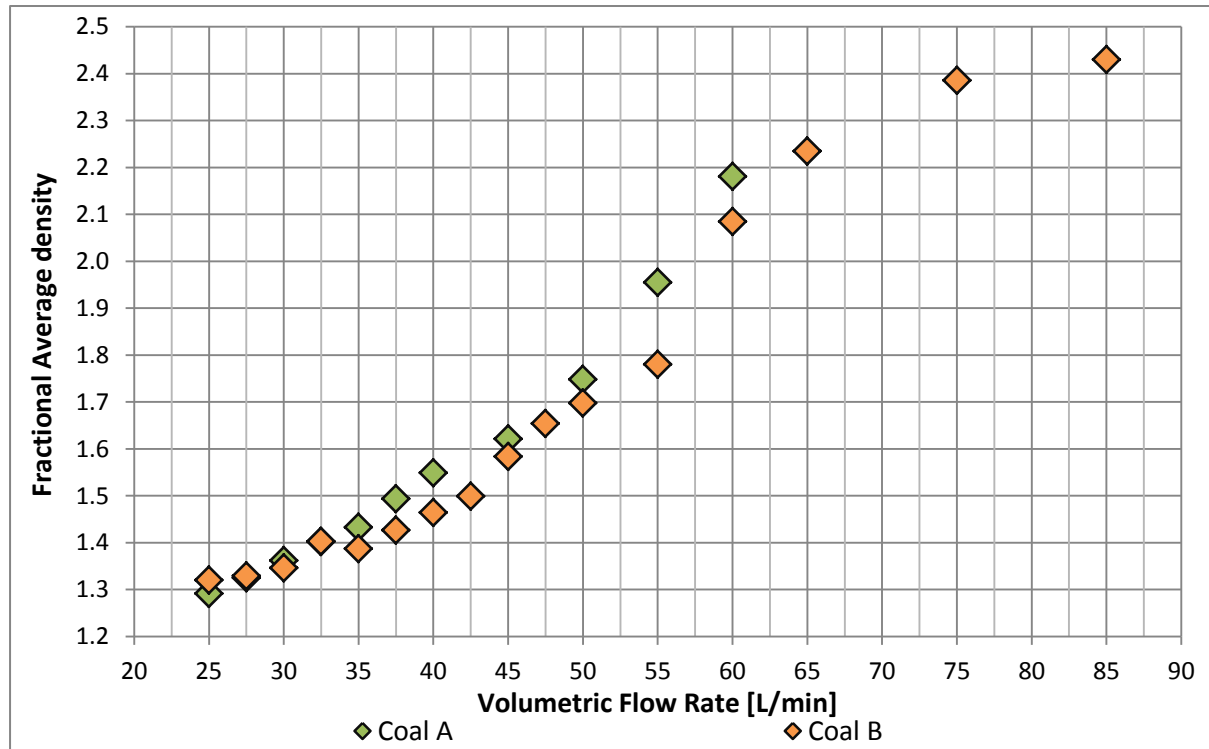


Figure 4-10: Correlation: Fractional average density and volumetric flow rate (Fractionation: Coal)

The bed densities were determined in a similar manner when the break-away from the linearity was imminent (55 L/min and 60 L/min for coal A and coal B respectively). The SGs used were 2.27 and 2.38 for coal A and coal B respectively, and the masses 111.92 grams and 110.97 grams for coal A and coal B respectively.

Table 4-3: Summary of the material and conditions during the fractionation test work

Run	Particle size distribution	Conditions	$\gamma_{s,i}$ (v/v)	$\gamma_{s,B}$ (v/v)
Tracer only	± 1 mm	Tracer in infinite diluted system	Negligible	Negligible
*Tracers + Coal A	± 1 mm	Coal ($\rho_{B,i} \approx 1.11$) + Tracers	0.198	0.021
*Coal A + Tracers	+75 -1180 μ m	Coal ($\rho_{B,i} \approx 1.11$) + Tracers	0.198	0.021
Coal B	+150 -1180 μ m	Coal ($\rho_{B,i} \approx 1.09$) only	0.160	0.019

**The same one test under the same conditions – differentiating between the material*

As originally proposed, it seems a minimum solids concentration (by volume) has to be present to maintain linearity. From the calculations and Table 4-3 it seems that this minimum solids concentration by volume is 0.20. It is possible that in this relatively low solids concentration a different flow regime was entered – giving this break-away from linearity. The tracer systems differ from the coal systems in the sense that all of the tracers are of an equal size. This may give a clue as to which properties the break-away affect adheres to – probably particle size. This is only an idea, and needs to be investigated further.

Another aspect that needs to be taken into account is that the data is represented in a fractional average density manner, and may also give non-linear data. This means that the actual cut-point may be linear, and the fractional average density not.

Overall fractionation summary

Figure 4-11 summarises the fractionation results, in fractional average densities. Similar results were obtained for both the tracer and coal tests. It is interesting to note that as the feed becomes finer, higher cut-densities are obtained – which is expected due to lower settling velocities of finer material. The tracers are in the order of 1000 μm , coal B in the range 300 μm to 1180 μm and coal A 75 μm to 1180 μm . Finer material may be present in the coal samples.

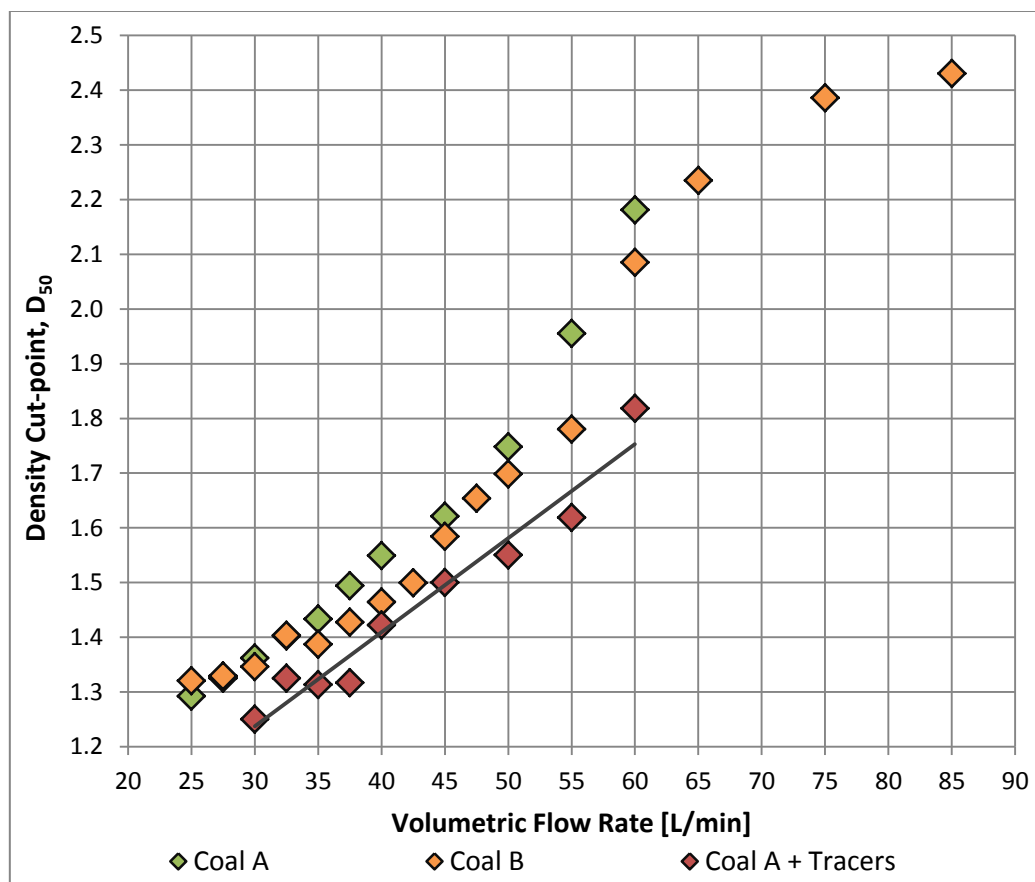


Figure 4-11: Summary of the correlations obtained during fractionation tests (fractional average density)

A clear difference in cut-density is observed regarding the hindered settling regime compared to the water-only diluted system. Higher cut-densities were observed in the coal system, due to the hindered settling effect. The tracers test data also showed a linear dependence of cut-density on fluid velocity in the plates. This corresponds to pilot reflux classifier trials done by Iveson *et al.* (2014).

4.2 Coal Washability

4.2.1 Specification and washability: Coal A

Coal A originated from a feed to an industrial reflux classifier pilot unit and had a particle size of less than 1.4 mm. The -1180 μm material was screened out by using the methods explained in section 3.2.2, while the +1180 μm material was reduced to -1180 μm , also by using methods described in section 3.2.2. The material was then de-slimed at 75 μm using wet sieve procedures. Thus, the particle size distribution of coal A was +75 -1180 μm during the tests. A float and sink analysis was done on the sample and the washability data for coal A is shown in Figure 4-12. The cumulative %ash for floats and sinks is given in Figure 4-12, and the calorific values are given in Figure 4-13.

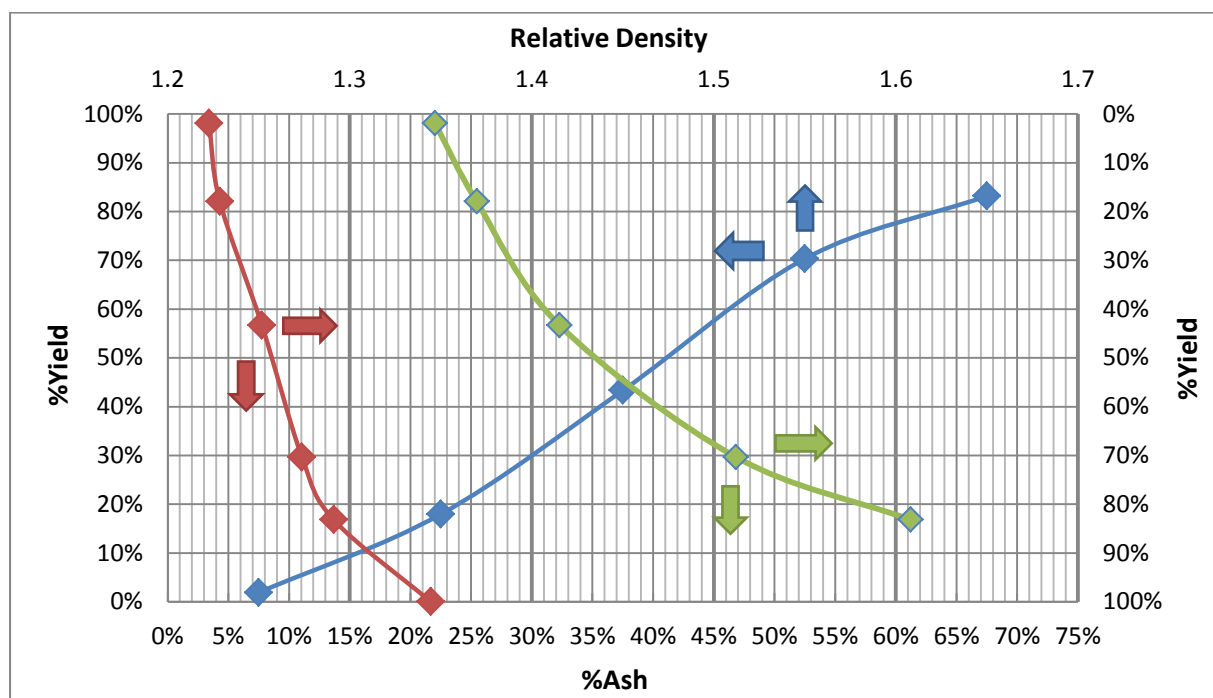


Figure 4-12: Densimetric and cumulative ash float and ash sink curves (Floats-Sinks: Coal A)

4.2.2 Specification and washability: Coal B

Coal B was also a feed to an industrial pilot unit, but from a different coal field as coal A. The particle size range was +500 μm -1400 μm . Coal B was also deslimed at 75 μm to ensure that all particles were bigger than 75 μm . The densimetric curves for coal B are given in Figure 4-14, while the calorific values are given in Figure 4-15.

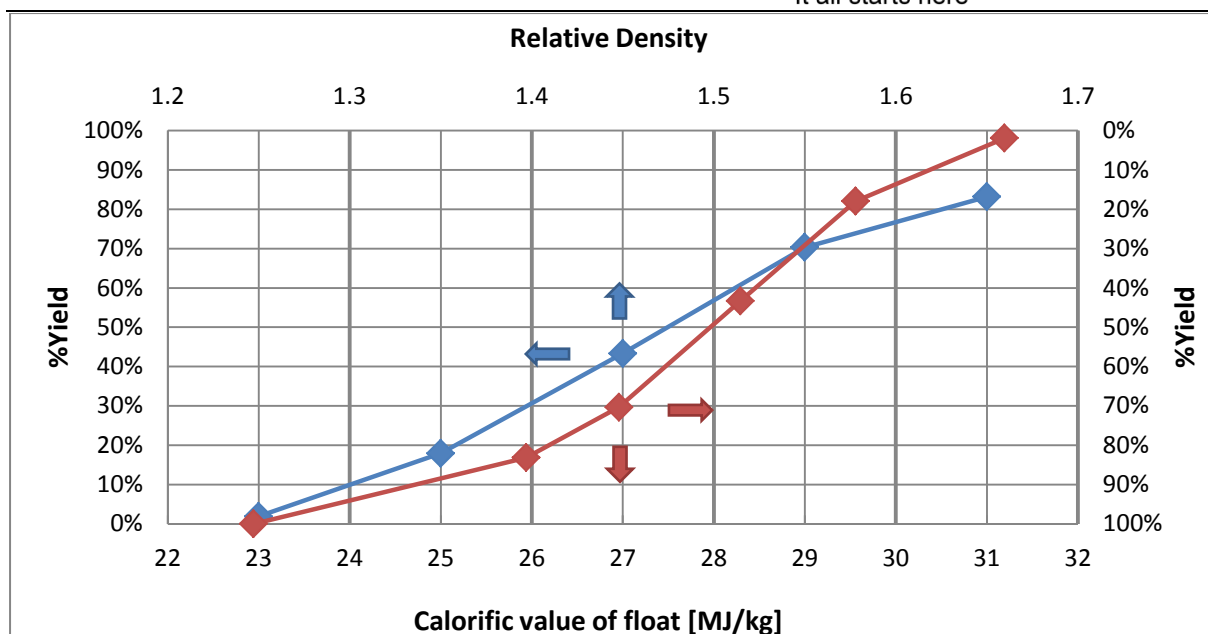


Figure 4-13: Densimetric curve and calorific value (Floats-Sinks: Coal A)

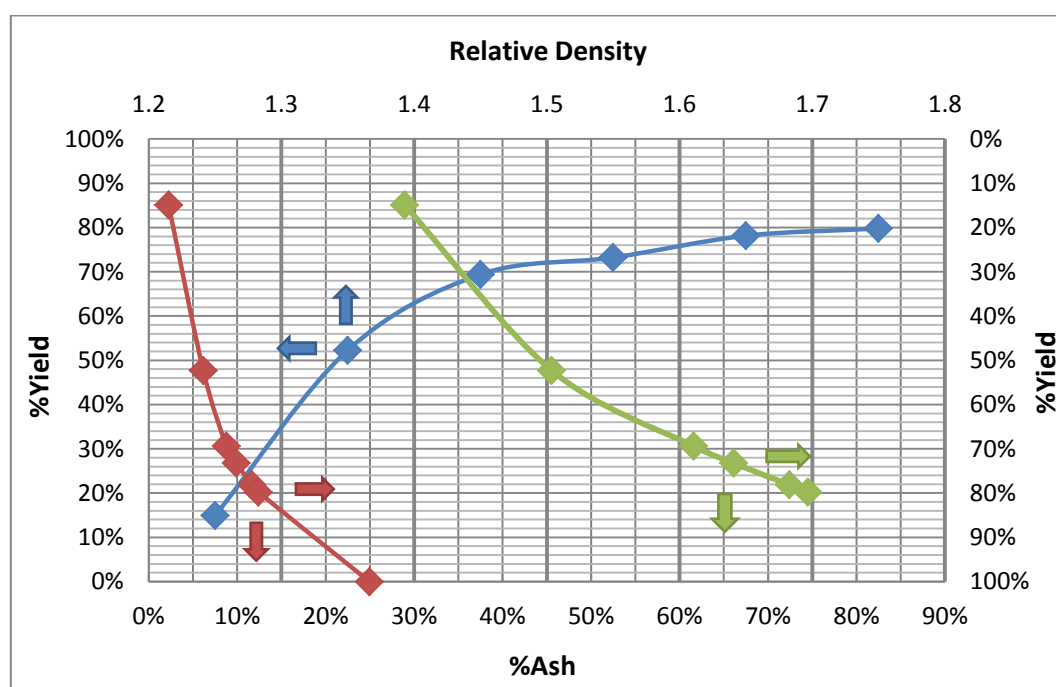


Figure 4-14: Densimetric and cumulative ash float and ash sink curves (Floats-Sinks: Coal B)

4.2.3 Washability analysis with RC105

In a previous study by Callen *et al.* (2008), it was shown that the reflux classifier can be used to determine the washability curves for coal resources. The data obtained during the fractionation test work (Figure 4-9) was also used to generate the densimetric and cumulative ash curves of coal B, which are shown in Figure 4-16. The densimetric and cumulative ash curves are compared to the densimetric and cumulative ash curves generated from the float and sink analysis in Figure 4-17 and Figure 4-18 respectively.

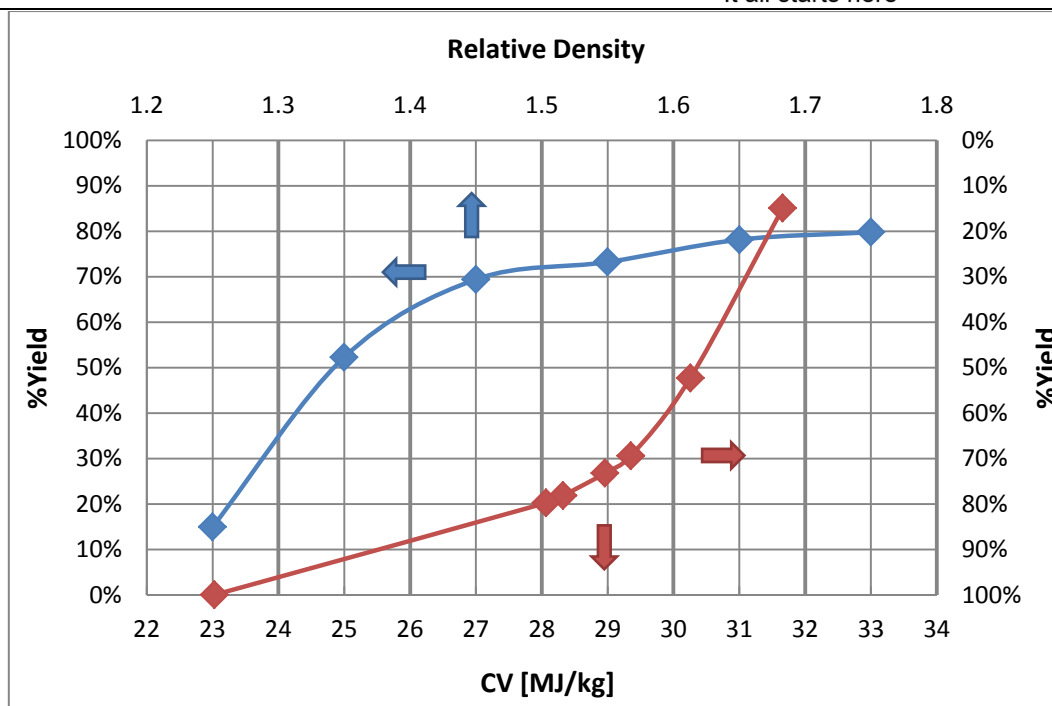


Figure 4-15 Densimetric curve and calorific value (Floats-Sinks: Coal B)

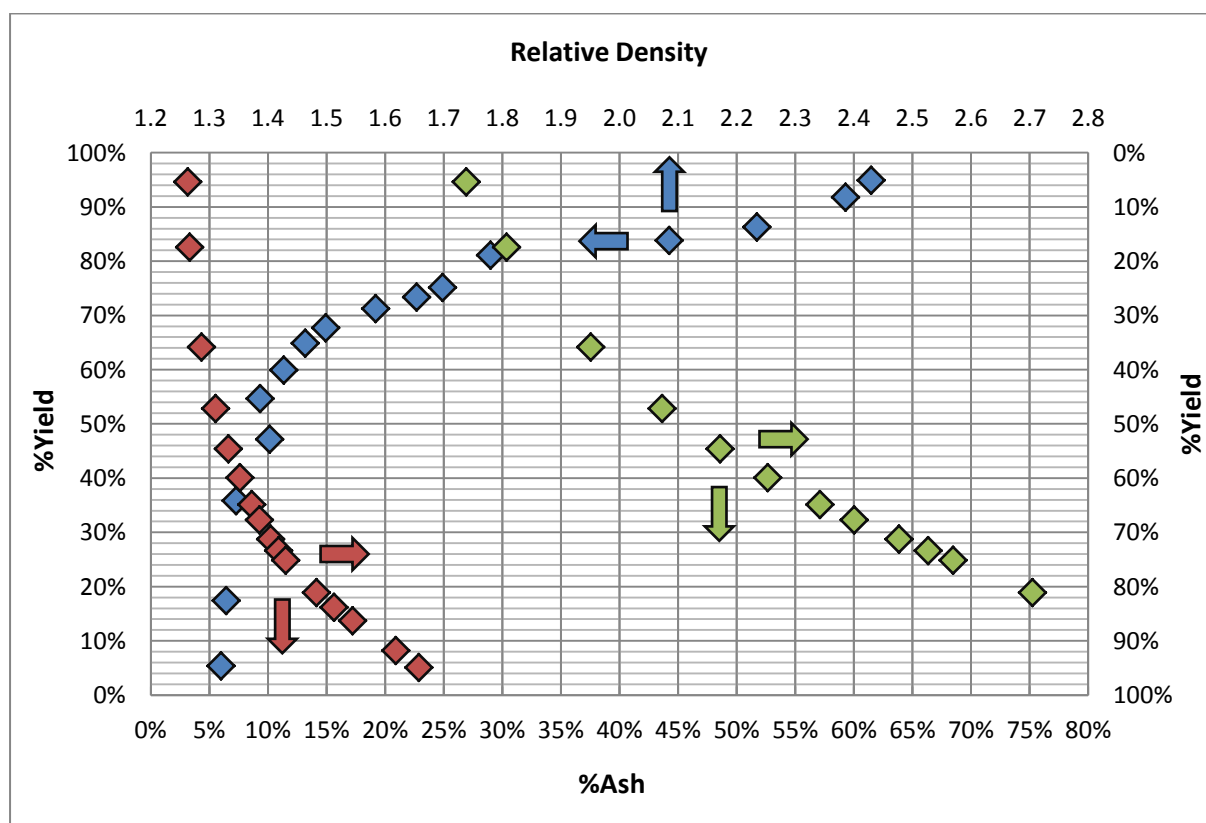


Figure 4-16: Densimetric and cumulative ash float and ash sink curves (RC105 fractionation: Coal B)

The laboratory reflux classifier setup can be used to determine the washability curves for coal. Both the densimetric and cumulative ash curves obtained through reflux classification fractionation are almost similar to the float and sink generated washability curves.

The deviation from the float-sink data at low densities (Figure 4-17) is a known phenomenon, and was first reported by Callen *et al.* (2008). They developed a mathematical model to predict the real washability curve from generated laboratory scale reflux classification data that matches the float-sink analysis data.

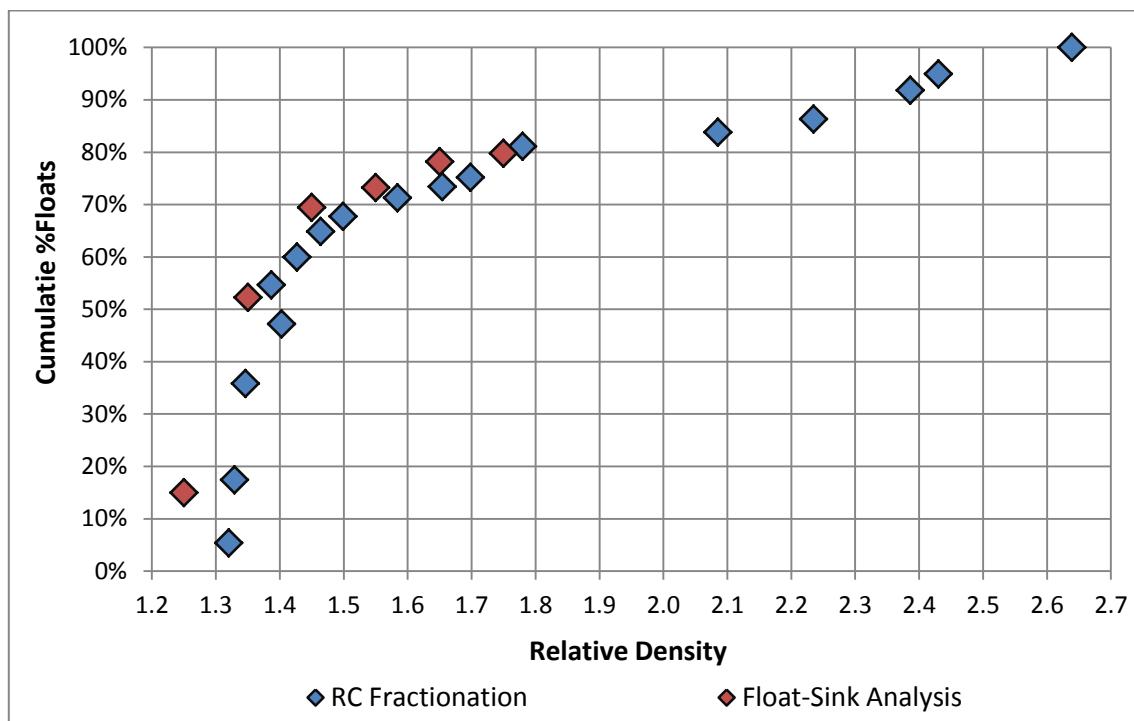


Figure 4-17: Comparing the densimetric curve obtained through float-sink analysis and RC105 fractionation: Coal B

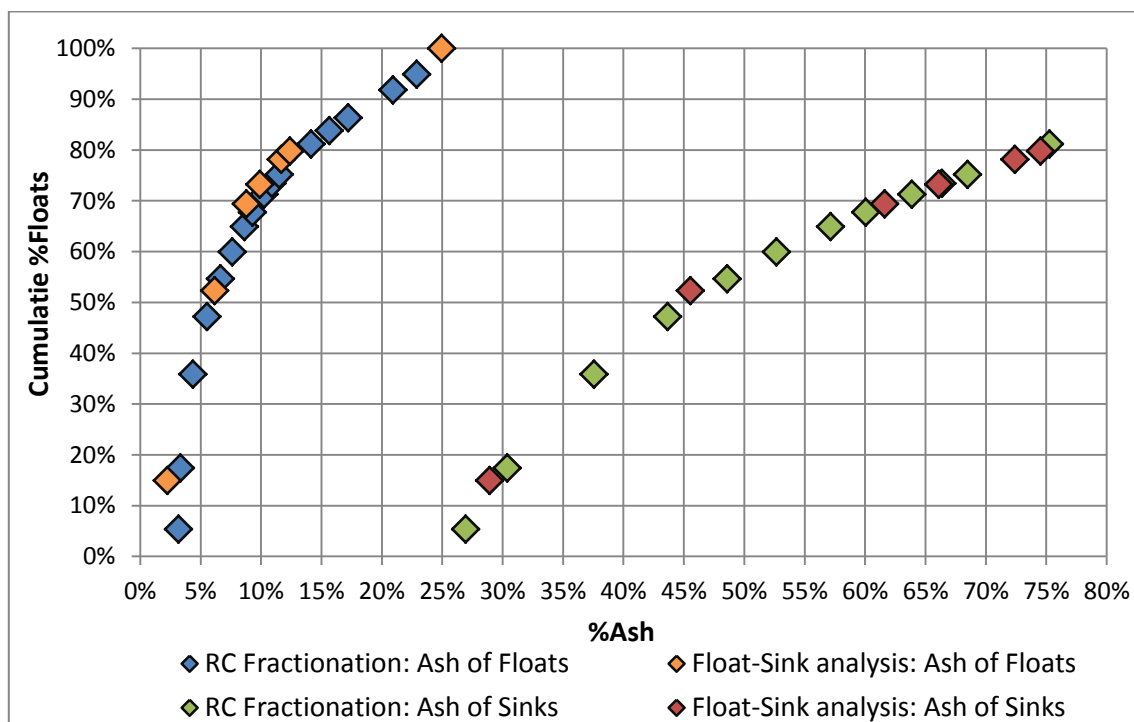


Figure 4-18: Comparing ash curves obtained through float-sink analysis and RC105 fractionation: Coal B

Advantages

The first advantage is illustrated in Figure 4-17: points for the relative density achieved are much higher than is possible with float sink analyses using a zinc chloride solution as heavy liquid. The densimetric curve generated through a float-sink analysis stops at a relative density of 1.75. This particular float-sink analysis was outsourced, but the highest density that can be obtained with a zinc chloride solution for instance is 1.75. Other advantages include:

- Reflux classification fractionation is a water-based operation, and no heavy organic or inorganic liquids are necessary that are hazardous to human health or the environment.
- More density intervals are obtained, with less effort than with zinc chloride solutions.
- The density intervals achieved are made narrower, by simple adjusting the water flow rate. This means higher resolution data, especially if near density or difficult to separate material is considered.

4.3 Separation Procedures

One outcome of the project is that it illustrates how the correlations developed in section 4.1 may be used to predict the density cut-point of separation runs performed at a fixed flow rate.

4.3.1 Tracers

The aim was to compare the data from the tracer fractionation runs (water-only system) to the performance of the reflux classifier where the unit was operated at a fixed flow rate for the entire run. Three flow rates were chosen as 45.0 LPM, 57.5 LPM and 67.5 LPM. Table 4-4 gives a summary of the actual and predicted cut-points at the chosen flow rates, where the predicted cut-points were obtained from Figure 4-3. This data is shown in Figure 4-19. It is evident that the cut-points were predicted well. It seems that in the case of a dilute system, the cut-points can be accurately predicted.

Table 4-4: Summary of predicted and actual cut-densities (tracer fractionation)

Fixed Flow Rate [L/min]	Predicted Cut-density	Cut-density Obtained
45.0	1.32	1.40
57.5	1.56	1.58
67.5	1.75	1.78

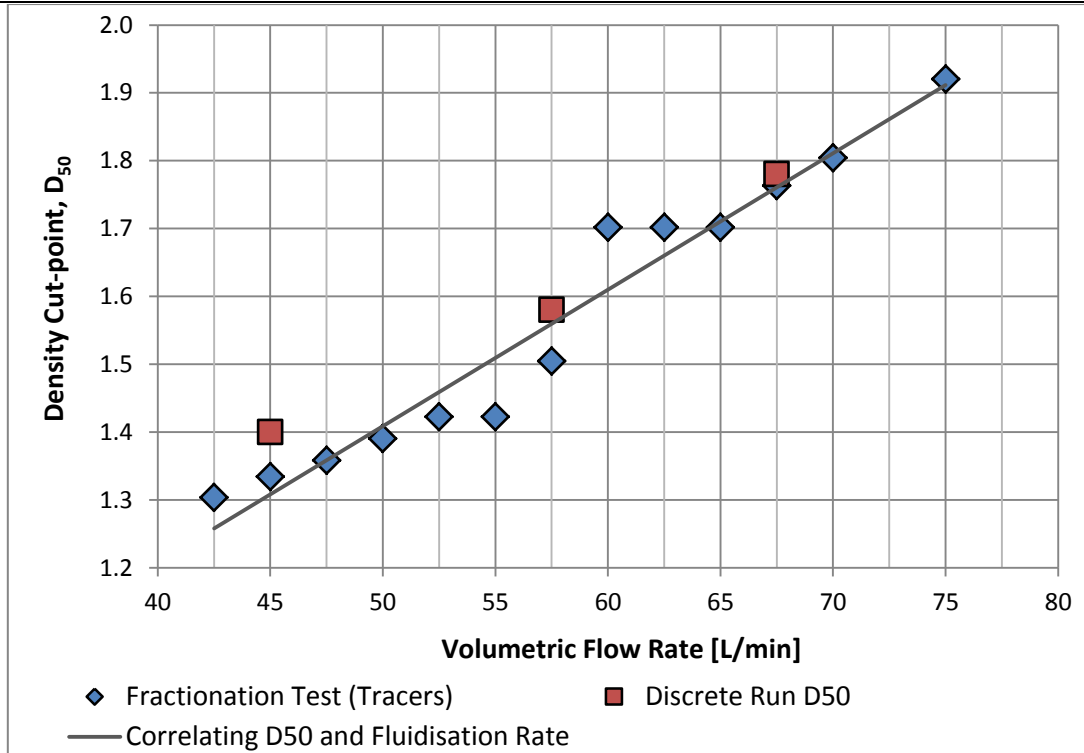


Figure 4-19: Obtained density cut-points compared to the fluidisation rate and D_{50} correlation (Tracers)

4.3.2 Coal A

A batch run was done using coal A at a water flow rate of 40 L/min. The overflow product had an ash content and calorific value of 12.3% and 26.7 MJ/kg respectively, while the discard had an ash content and calorific value of 45.8% and 14.5 MJ/kg respectively. The mass yield to the clean coal (overflow) was 73%. The partition curve (determined with float and sink analysis) for the separation run is given in Figure 4-20. The cut-point and e.p.m. calculated by means of the model are 1.69 and 0.02 respectively. An e.p.m. value of 0.02 is more competitive than for example fine dense-medium cyclones (De Korte, 2014).

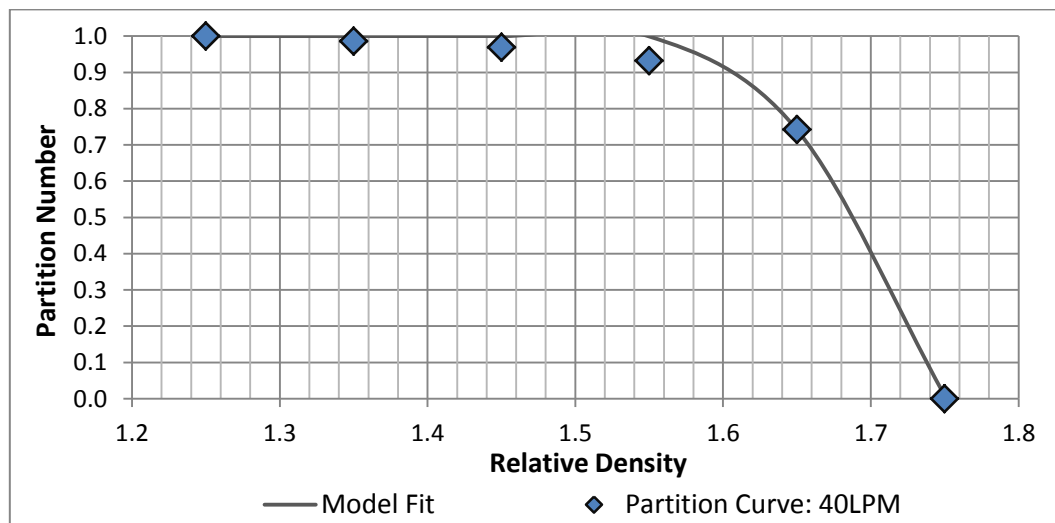


Figure 4-20: Partition Curve of the fixed flow rate run at 40 L/min: Coal A

4.3.3 Coal B

Separation runs on coal B were done at three fluidisation flow rates: low (32.5 L/min); normal (42.5 L/min); and high (60.0 L/min). The effect of initial bed density was also tested and one extra run at double the charging mass was performed on the 32.5 L/min flow rate. One run was also repeated to check repeatability, and one extra run at 32.5 L/min was done.

The float and sink analyses for these runs were outsourced to Witlab Pty Ltd in Witbank. Tromp curves of the low, normal and high flow rate separating runs were constructed and are shown in Figure 4-21. Cut-points and e.p.m values were determined, and are summarised in Table 4-6. Less efficient separation occurred for the coal B separation runs when compared to the separation run for coal A, especially when looking at the run done at 42.5 L/min.

Table 4-5: Cut-point and efficiency data: (Separation Runs: Coal B)

Flow rate [LPM]	e.p.m.	Cut-point Actual
32.5	0.039	1.41
42.5	0.083	1.62
60.0	0.038	1.85

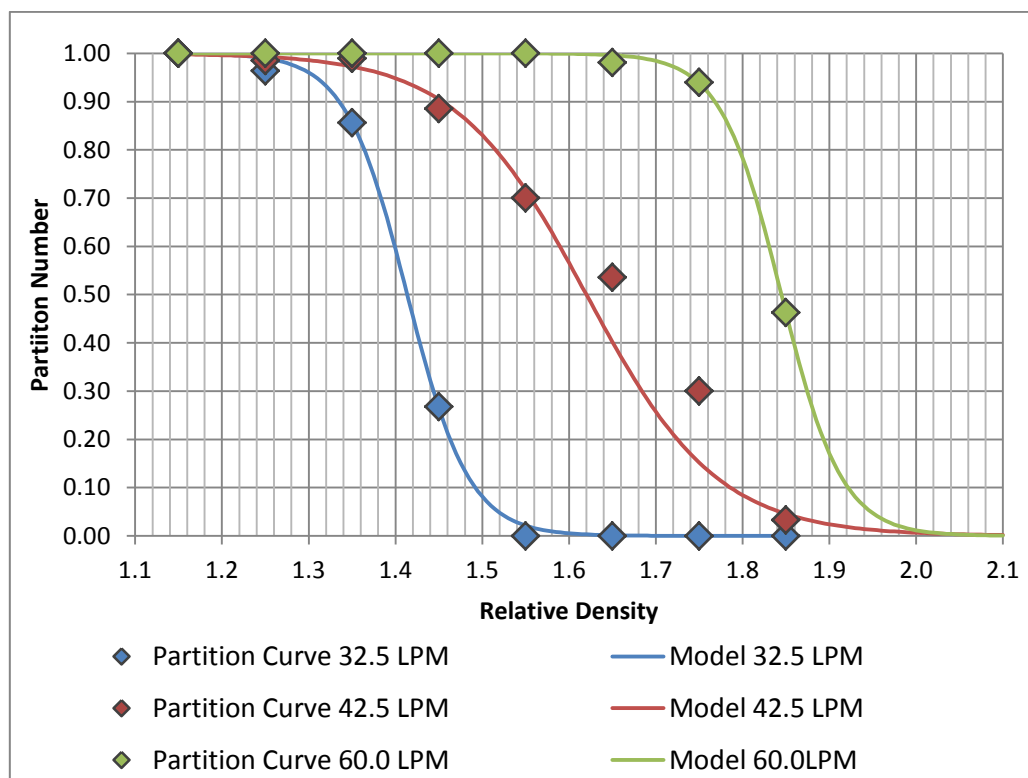


Figure 4-21: Partition curves for the three fixed flow rate runs: Separation runs (Coal B)

The cut-points for separation runs on coal could not be predicted, since Figure 4-9 provided average density predictions. However Figure 4-22 (cumulative data) was used to determine the average overflow density that was achievable for coal B. Table 4-6 summarises the predicted overflow average density and yield (where the yield was determined from Figure 4-9), and then also the average overflow density and yield obtained. These overflow average densities were determined by mass balance. From Table 4-6 it is evident that accurate predictions for the average density of the separation test products were made, with a maximum difference of 2.3%. The predicted values are the data and performance that were obtained during the separation runs performed.

Table 4-6: Comparing actual product average density and yield to predicted data (Separation Runs: Coal B)

Flow rate [LPM]	Predicted Yield	Actual Yield	Error Margin	Predicted Average Density	Actual Average Density	Error Margin
32.5	47%	54.7%	16.4%	1.35	1.32	2.2%
42.5	68%	73.8%	8.5%	1.38	1.37	0.7%
60.0	84%	85.6%	1.9%	1.43	1.40	2.0%

Figure 4-22 summarises the cumulative average densities of the various fractionation runs performed. The coal runs are not comparable, because the cumulative data is heavily dependent on the washability of the coal.

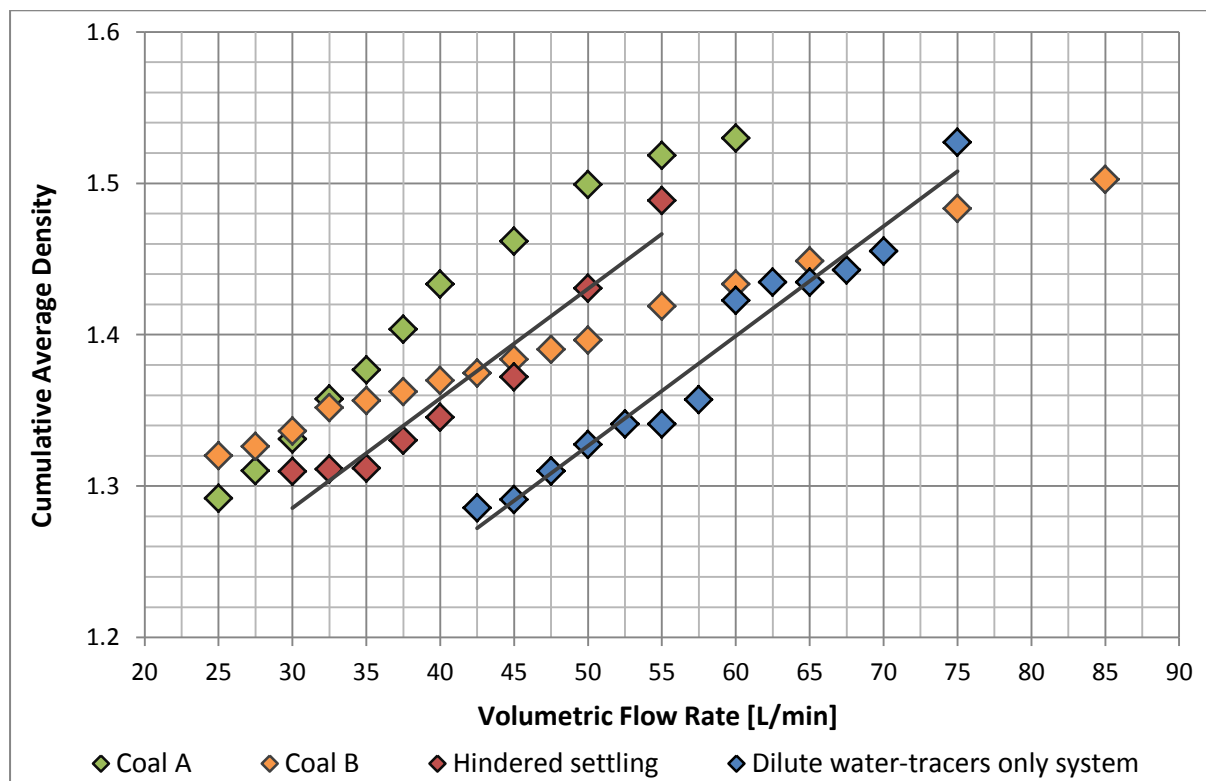


Figure 4-22: Cumulative average density for coal A and B, and tracer data

The predicted overflow average density and yield are also visually presented in Figure 4-23 and Figure 4-24 respectively. The average overflow densities were very accurately predicted. On the other hand, yields were not as accurately predicted. It seems that the more the flow rate is increased the more accurate the yield can be predicted. Interestingly, the overall overflow average density is less, while the yield is greater than was predicted.

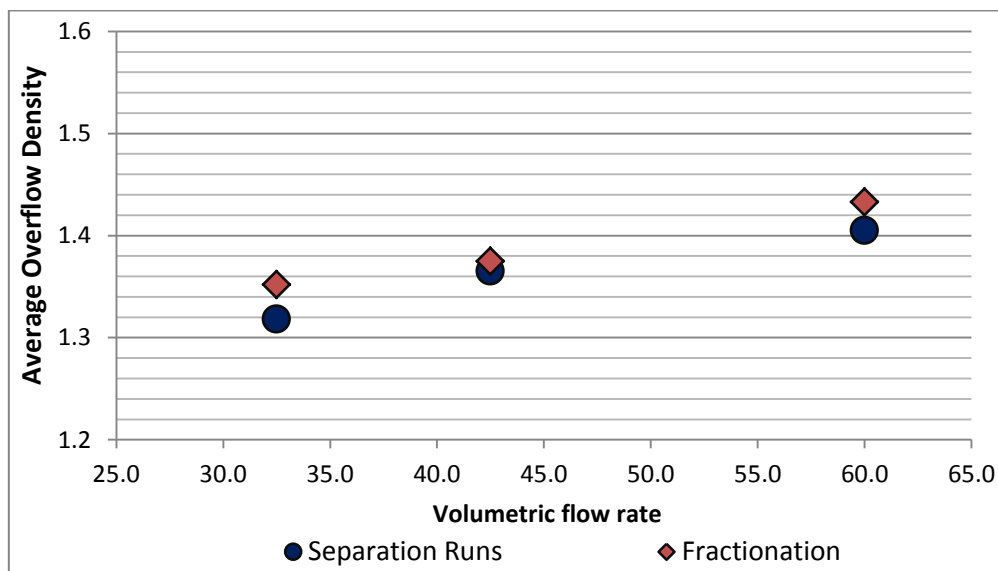


Figure 4-23: Comparing the predicted and obtained overflow average density: Separation runs (Coal B)

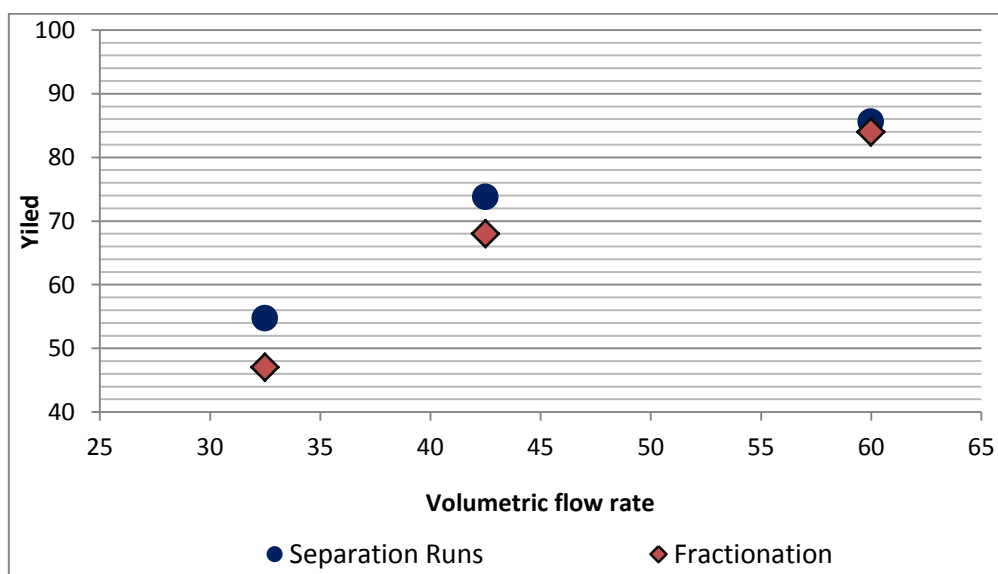


Figure 4-24: Comparing the predicted and obtained yield: Separation runs (Coal B)

Extra runs at 32.5 LPM

Two extra separation runs were done; one repeat at 32.5 L/min and one with double the material load to investigate a high initial bed density also at 32.5 L/min. Figure 4-25 shows the Tromp curves for these two runs and compares them to the Tromp curve of the first run done at 32.5 L/min.

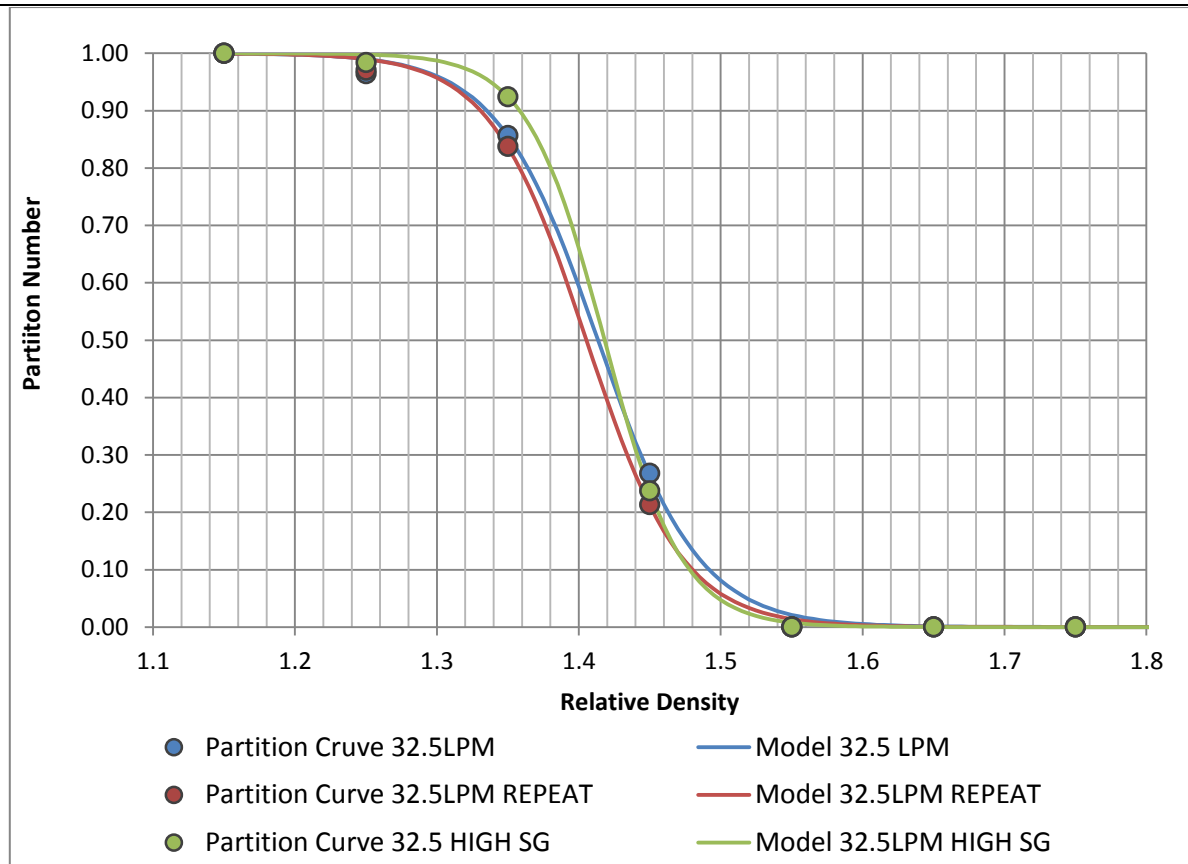


Figure 4-25: Comparing Tromp curves of the repeat and high $p_{B,i}$ runs with the original separation run at 32.5LPM: Coal B

The repeat yielded results very similar to the original results. It was expected that the repeatability would be good. The run investigating the high initial bed density had no effect on the density cut-point, as can be seen in Figure 4-26. It does however show a sharper separation – Figure 4-27 shows a decrease in the e.p.m. value from 0.038 to 0.030. Although it seems small, this kind of increase allows a much sharper separation to occur.

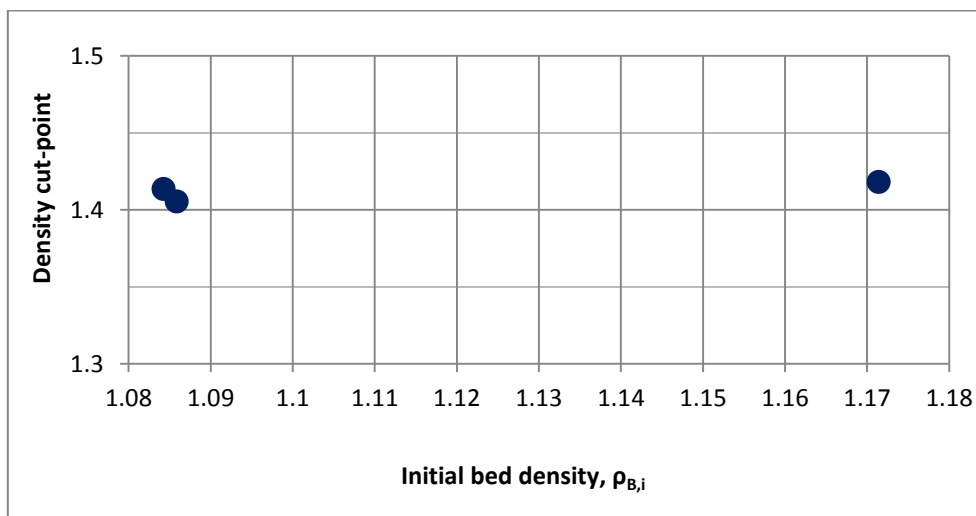


Figure 4-26: Correlating density cut-point to initial bed density in batch mode: Coal B

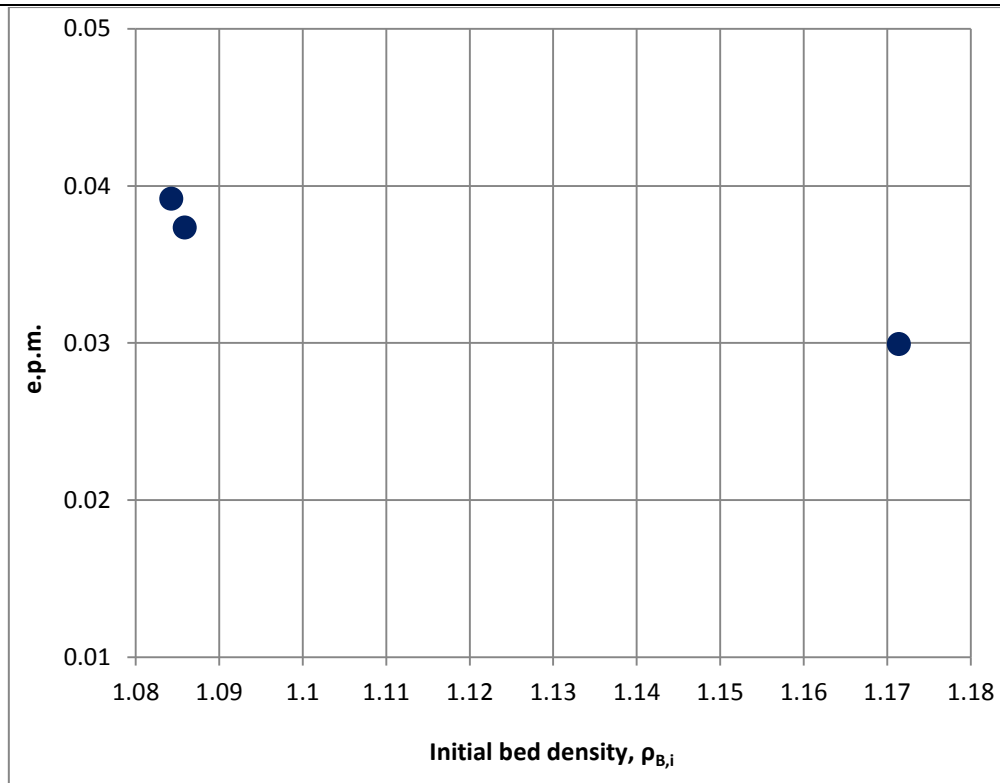


Figure 4-27: Correlating e.p.m. to initial bed density in batch mode: Coal B

High solids concentration advantage

When particles in the reflux classifier fluidised bed experience a high solids concentration they form an autogenous dense medium which ejects lighter particles into the lamellae, and when the solids concentration can be maintained, e.g. at 0.2, the reflux classifier will run at perfect conditions.

4.3.4 Separation test procedure – Summary

The density cut-point was accurately predicted from the density cut-point to volumetric flow rate correlation (Figure 4-3) for the case of infinite dilution and a narrow particle size range – tracers in water-only system.

Without float-sink analyses for the fractionation procedures, no estimated density cut-point prediction for the coal separation runs could have been made. However, the average density of the overflow product could accurately be predicted from cumulative average density data from the fractionation runs.

Other valuable insights into the laboratory unit operation were gained during the separation test procedures:

- The initial bed density does not influence the cut-point for batch operation within a certain range.
- An increase in initial bed density improved the separation efficiency.
- The repeatability of the batch mode discrete procedures is good.

4.4 The Influence of Particle Size

The supplier specified the particle size distribution of coal B as $-1400 +500 \mu\text{m}$. The material smaller than $1180 \mu\text{m}$ material was sieved out, and material $+1180 \mu\text{m}$ was crushed to $-1180 \mu\text{m}$. Figure 4-28 shows the particle size distribution of coal B, which was done in-house (as described in section 4.4). It also shows that the particle size distribution of the material used during the experiments is $-1180 +150 \mu\text{m}$, and that 28% passes $500 \mu\text{m}$. It also shows that a uniform size distribution exists, which is advantageous when looking into how particle size influences the separation efficiency of the reflux classifier.

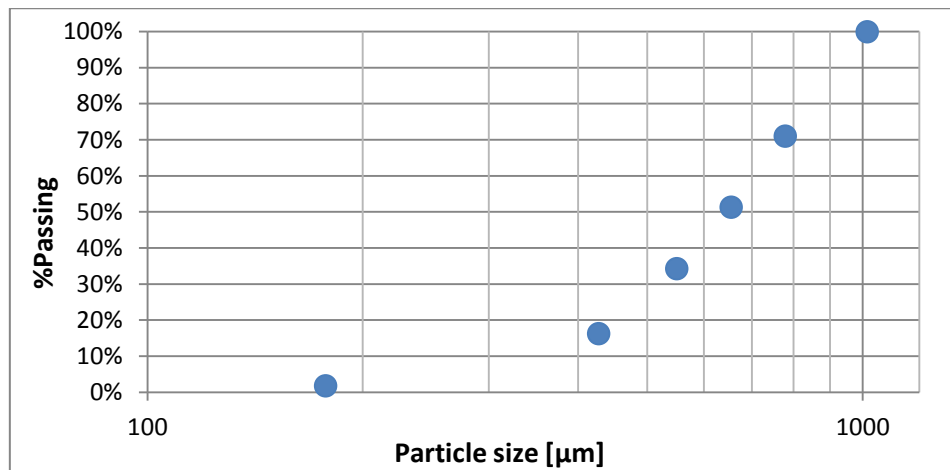


Figure 4-28: Particle size distribution: Coal B (Feed)

The relative density of particular particle size ranges was investigated. This will be used to compare where light big particles reported to, as opposed to heavy small particles. The float-sink data of four size ranges ($-355 +150 \mu\text{m}$; $-600 +355 \mu\text{m}$; $-850 +600 \mu\text{m}$; $-1180 +850 \mu\text{m}$) is shown in Figure 4-29, to illustrate how the density of the particles varied with particle size in the feed samples for coal B. From Figure 4-29 it is deduced that the mineral content was uniformly distributed amongst the particles, except for the $-355 \mu\text{m}$ particle size class. The particles smaller than $-355 \mu\text{m}$ were significantly lighter, and consequently had a lower mineral content. Normally the fines would be higher in mineral content, which produces problems in the separation efficiency of gravity separation units. These problems are generally overcome with de-sliming of the material at $150 \mu\text{m}$ prior to and following the gravity separation stage.

Runs varying the particle size ranges (i.e. comparing wide and narrow particle size ranges) were not executed. The main focus of the study was to correlate laboratory generated data with pilot unit reflux classification data. The investigation of the effect of particle size on the separation efficiency was limited to the determination of the density of different particle size ranges, and where the small heavy and large light particles reported to.

The coal samples obtained from the float-sink analyses of the underflow and overflow of the 32.5 L/min run were subjected to PSD analyses, and Tromp curves were drawn for each of

the particle size ranges (Figure 4-30). The density cut-point and e.p.m values of each of the curves are summarised in Table 4-7.

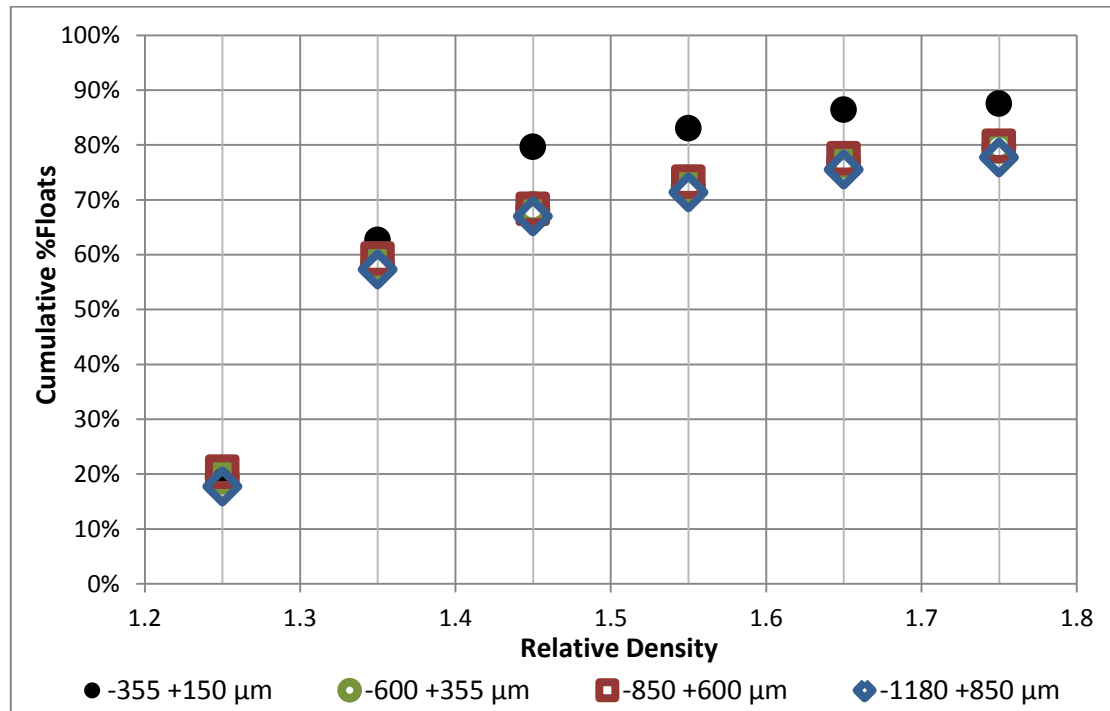


Figure 4-29: Float-Sink data for each of the four PSD ranges: Coal B (Feed)

Table 4-7: Summarising the density cut-point and e.p.m values of each particle size range's Tromp curve: Original run at 32.5 LPM (Coal B)

Tromp Curve	Density cut-point	e.p.m.
-1180 +850 µm	1.37	0.027
-850 +600 µm	1.40	0.030
-600 +355 µm	1.45	0.026
-355 +150 µm	1.47	0.012
Original	1.41	0.039

There is a clear increase in cut-point as the particle size decreases, and it is noted that higher separation efficiencies are experienced for the individual particle size ranges. This will always be the case in gravity separation units where a large particle size range is treated, and the combined separation efficiency will be lower than the individual particle size ranges.

The most narrow size range (-355 +150 µm) shows the best separation efficiency, while the relatively small density cut-point increase in this size range is due to a significant higher amount of lights. The density cut-point increase for the rest of the Tromp curves is about the same due to the uniform density distribution over the size range -1180 +355 µm.

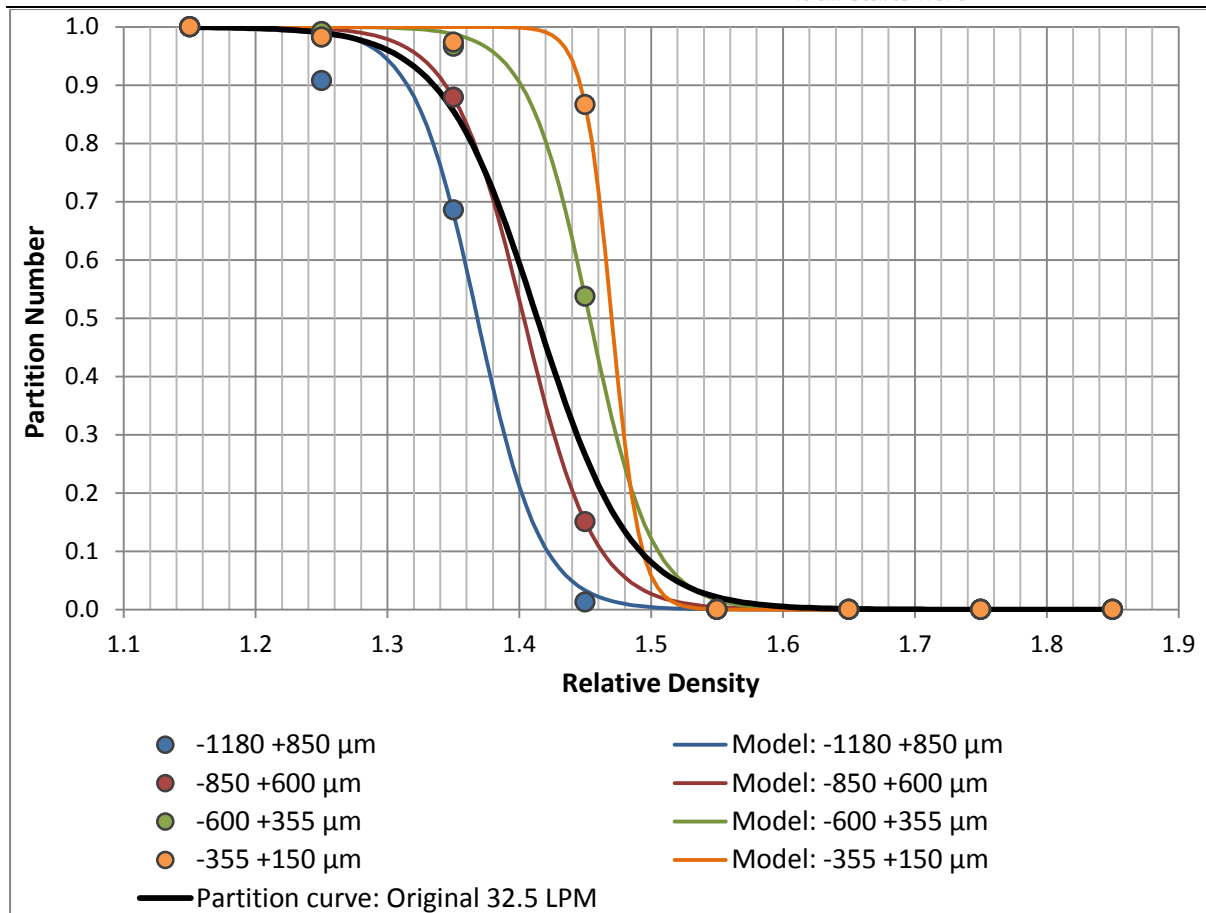


Figure 4-30: Comparing the Tromp curves for the three particles size distributions: Original 32.5 LPM run (Coal B)

4.4.1 Mineral content and particle weight

The weight of a coal particle is dependent on the amount of mineral content within that particular particle. Figure 4-31 shows the %ash within each of the particle size ranges in each of the float-sink fractions of the underflow and overflow samples.

The mineral content in each of these fractions increases as the particle size decreases, meaning that more small heavy particles preferentially report to the overflow, while more large light particles preferentially report to the underflow. This is not to a great extent though (where the extent can be quantified by the slight slope of each individual trend), as illustrated in Figure 4-31. The average slope is very small, and it can be deduced that the reflux classifier preferentially separates on the mineral content rather than particle size.

Also noteworthy, is that the cut-point for this specific run was 1.41. When looking at the mineral content present within the material that floated in a 1.4 SG solution for the overflow and underflow samples, the overflow sample had on average 5% less ash than the underflow, which also confirms that the reflux classifier separates preferentially on mineral content rather than particle size. This may also be taken as proof that smaller liquid density

intervals are needed to produce proper washability curves, and it is easier to do this with a laboratory scale reflux classifier.

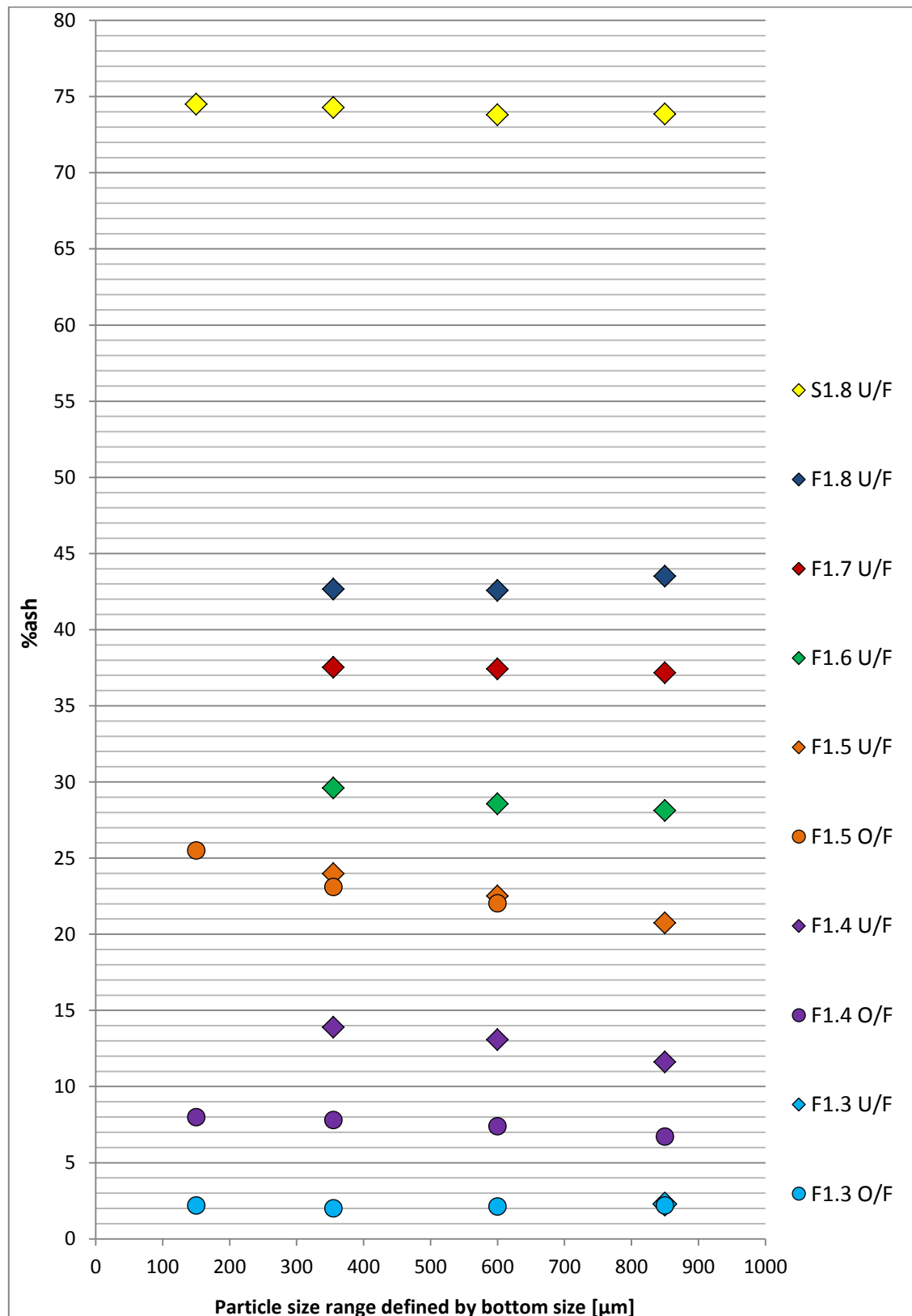


Figure 4-31: %ash of the particular size range in each float-sink fraction – the PSD is defined by bottom size:
Original run at 32.5 LPM (Coal B)

4.4.2 Comments on sample splitting technique

The size distribution of coal B was determined during the initial stages of the experimental endeavour, and is given in Figure 4-28. It is trivial to mention that the sample for this analysis had to be taken from the bulk sample. After a batch run at 32.5 L/min during a separation procedure, the float-sink samples of the overflow and underflow were split into four size fractions. The feed sample for this particular run was taken from the bulk sample at a later stage as the initial feed sample above.

These PSDs were compared to see how accurate the samples were split. The masses of each size interval from each float and sink analyses of the overflow and underflow were combined to obtain the reconstituted PSD fed during the batch run. The initial feed sample's PSD data had to be reconstructed to form the same four particle size range intervals, in order to compare the two. These particle size distributions are given in Figure 4-32.

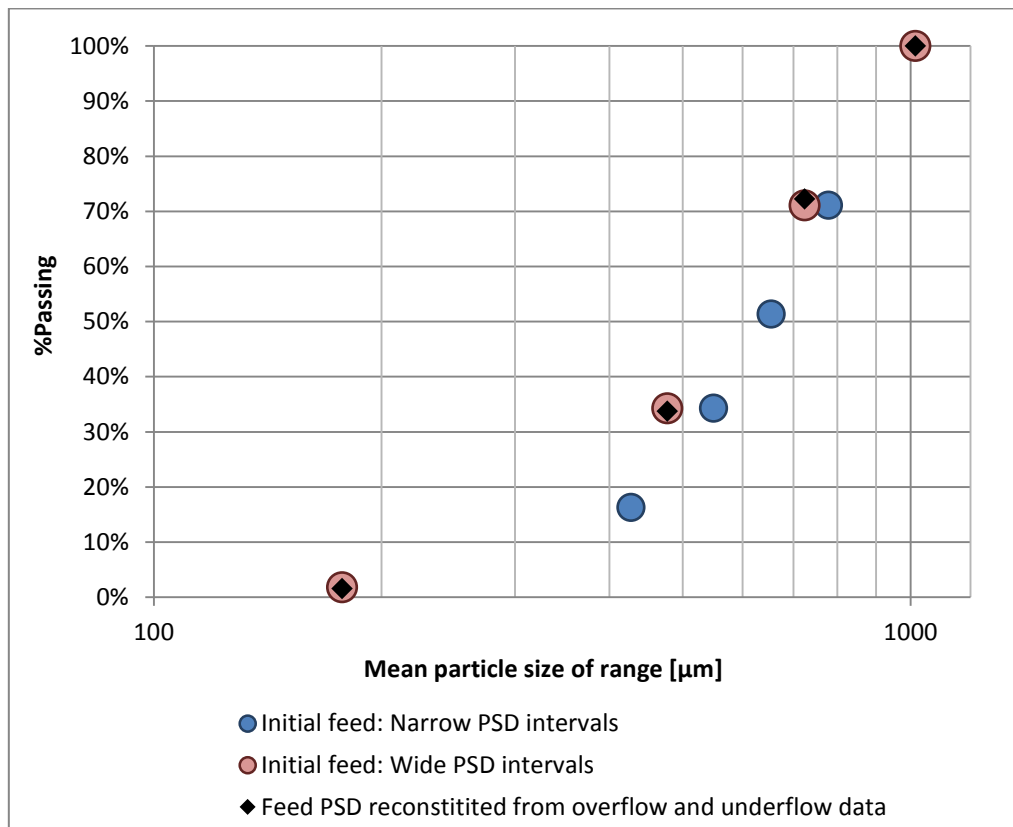


Figure 4-32: Comparing PSD's of sample split products from different sample split instances

The brown and blue dots signify the initial feed sample taken, where the blue dots give the six interval PSD and the red dots the four interval PSD. The black diamond shapes represent the four particle size ranges of the reconstituted feed from the overflow and underflow float and sink analyses products.

The reconstituted data was derived from a sample that was split from the bulk at another instance than the sample used to derive the data represented by the blue and red dots.

Thus, it is deduced that a sound sample splitting technique was used – a rotating riffing technique.

Another important observation is that exactly the same data presented in wider size ranges seems finer than when they are presented in narrower particle size ranges. The same observation is made when looking at the washability curves (Figure 4-17 and Figure 4-18) obtained through the use of the RC105, where much more density intervals are obtained at smaller intervals used during float and sink analysis procedures.

4.5 Comparing correlations with literature

Walton (2011:68) derived three equations that can be used to determine the minimum fluidisation water flow rate needed to elutriate a particle with a certain size and density for each of the three settling regimes in normal reflux classifier operation when the plates are narrowly spaced. The equation of interest (equation 4.2) relates the superficial fluid velocity required to the plates spacing z , particle size d_p , fluid density ρ_F and fluid viscosity μ the intermediate settling regime.

$$U = 0.0479z \frac{d_p^{1/7} \Delta\rho^{5/7} g^{5/7}}{\rho_F^{2/7} \mu^{3/7}} ; 2 < Re_p < 10^3 \quad 4.2$$

$\Delta\rho$ is the density difference of the conveying fluid and the particle that experiences just enough superficial fluid velocity to be elutriated. It can then be taken that this density is the cut-density, because the particle just mentioned has an equal opportunity to be either elutriated or land on the plate edge and slide down.

It has been well established in this dissertation and other reference works (Iveson *et al.*, 2014; Galvin & Lui, 2011) that the density cut-point and volumetric flow rate relationship is linear to slightly curved. To compare the correlations obtained in previous sections with equation 4.2, the correlation data must be modified. Firstly, the correlation obtained in section 4.1.5 (Figure 4-11) is average density data. Secondly, the volumetric flow rate on the x-axis needs to be converted to superficial fluid velocity. The latter is easily done when the area and volumetric flow rate is known.

After the two outer channels of the plates section were closed off, only 22 channels remained spaced 3.5 mm apart with a plate thickness of 0.9 mm. The volumetric flow rate within 1 channel will then be $Q_{ch} = Q_W \div 22$ and the channel area is $A_{ch} = 0.00372 \times 0.105 = 3.911 \times 10^{-4}$. The width was taken as $3.5 \div \sin 70$, because this distance will be the length normal to the horizontal plane; thus, calculating the area which was seen by the fluid. Equation 4.3 can then be used to calculate the superficial velocity of each of the flow fractions for the RC105 system.

$$U_{Ch,s}[m.s^{-1}] = \frac{Q_w [m^3.s^{-1}]}{8.604 \times 10^{-3}} \quad 4.3$$

To address the issue of the density cut-point and average density, an overall correlation for the coal data was obtained. The fractionation data up to 55 L/min for both coal A and coal B was considered in order to develop a modified correlation between the fractional overflow density and the flow rate. The new correlation is shown in Figure 4-33, and has a slope of 0.0164 and intercept of 0.864.

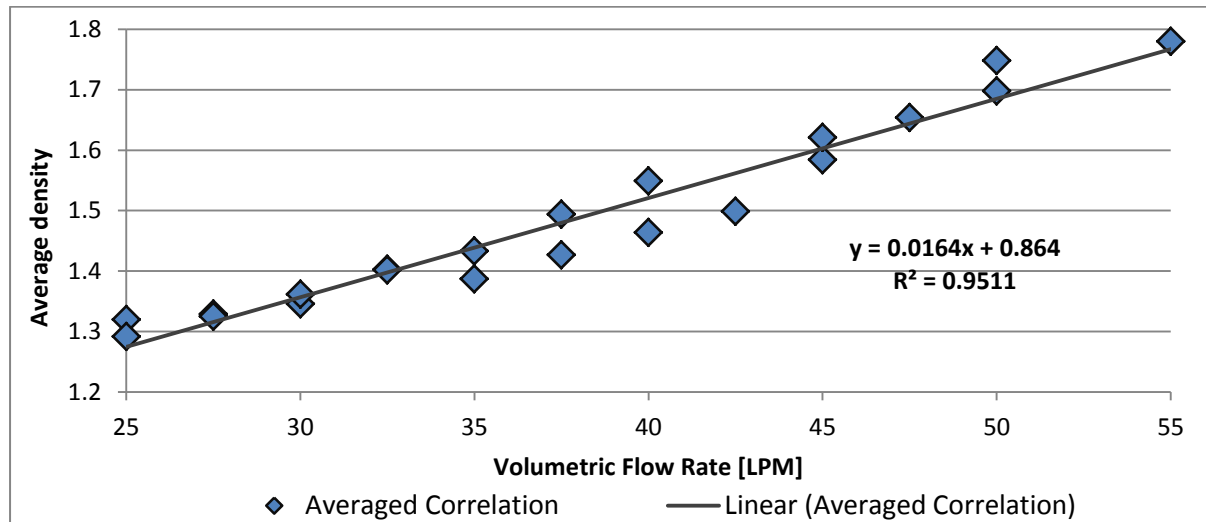


Figure 4-33: Averaged correlation: Density cut-point and volumetric flow rate

The actual density cut-points of the separation runs done on coal B were then plotted on this new correlation (Figure 4-34) with slope of 0.0164 and intercept of 0.864, and it was found that the correlation models the actual cut-points accurately. After the correlation shown in Figure 4-34 was converted to superficial fluid velocity by means of equation 4.3, it was compared to the correlation developed by Walton (2011:68) – equation 4.2.

Different particle sizes are modelled in each of the graphs in Figure 4-35, Figure 4-36 and Figure 4-37, to show how a data matrix can be obtained. The other variable is plate spacing and the sensitivity to plate spacing is compared through the three graphs, where 300µm was added to (Figure 4-36) and subtracted from (Figure 4-37) the actual plate spacing (Figure 4-35).

Particle size does seem to have an influence on the superficial fluid velocity required to elutriate the particles. The black line which represents the correlation obtained in Figure 4-34 is in excellent agreement with the yellow line which presents a plot of equation 4.2 for a particle size of 1180 µm and an RC system with plates spaced 3.5 mm apart. The top sizes of all of the runs were 1180 µm, and it is seen that the relationship is not necessarily linear.

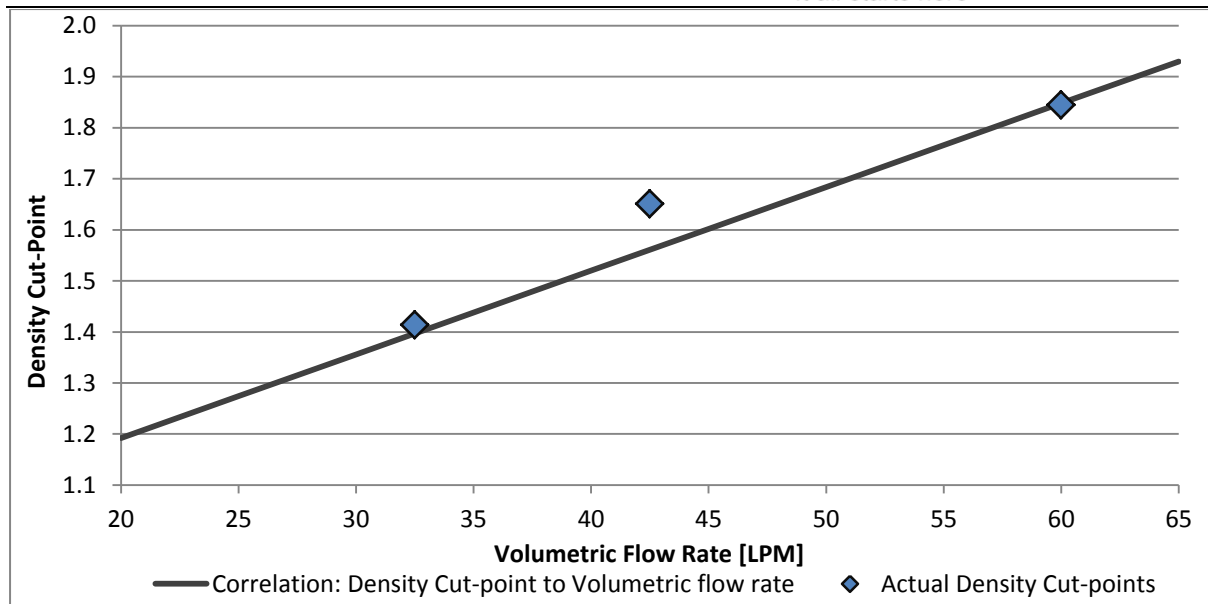


Figure 4-34: Comparing the actual density cut-points obtained to the new correlation

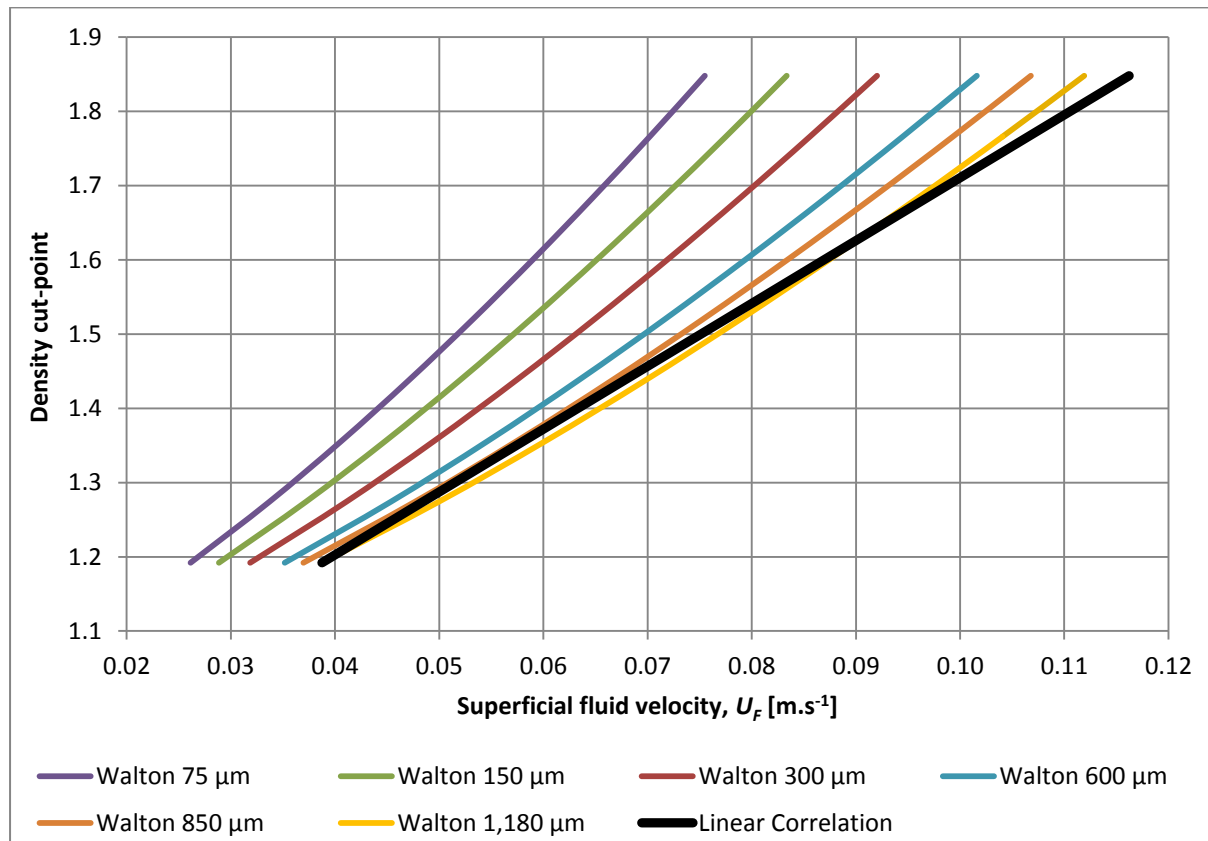


Figure 4-35: Comparing correlations to Walton (2011): $z = 0.0035$ m

Figure 4-36 and Figure 4-37 shows the effect when the plate spacing is changed incrementally by 0.3 mm in both directions from 3.5 mm. It is seen from Figure 4-36 and Figure 4-37 (and equation 4.2) that when a wider plate spacing is used, a higher fluidisation flow rate is required.

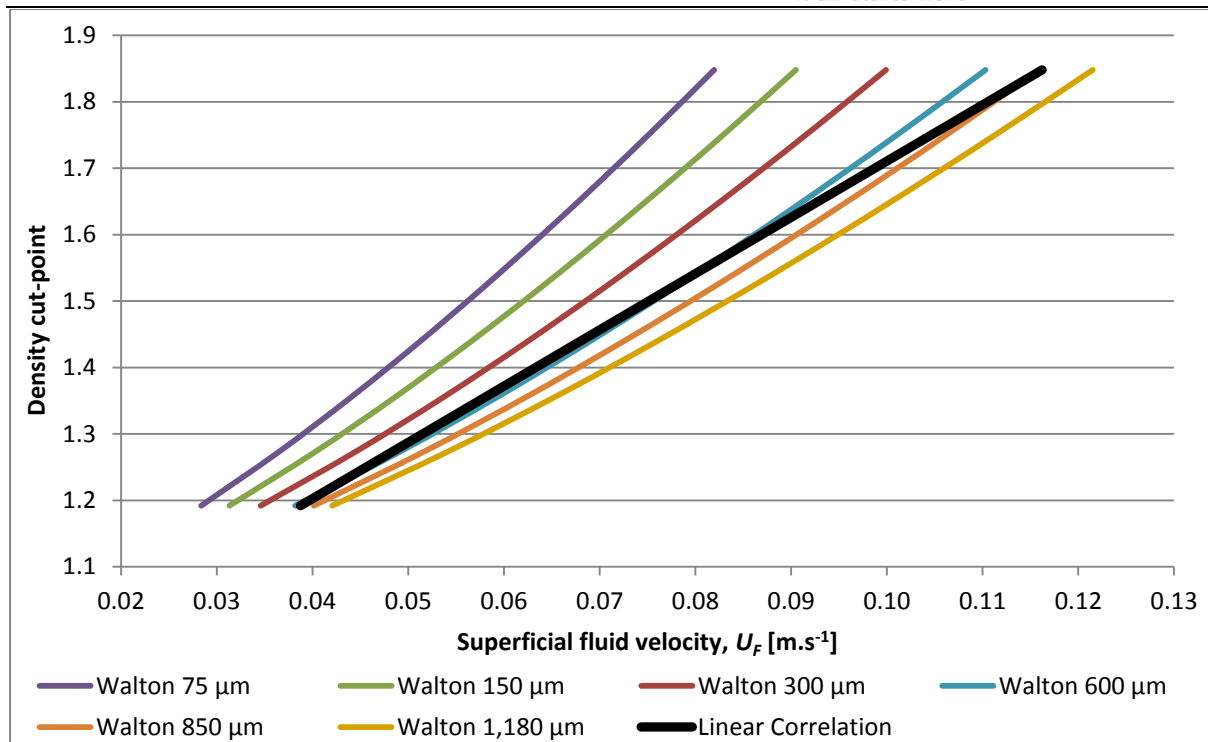


Figure 4-36: Comparing correlations to Walton (2011): $z = 0.0038$ m

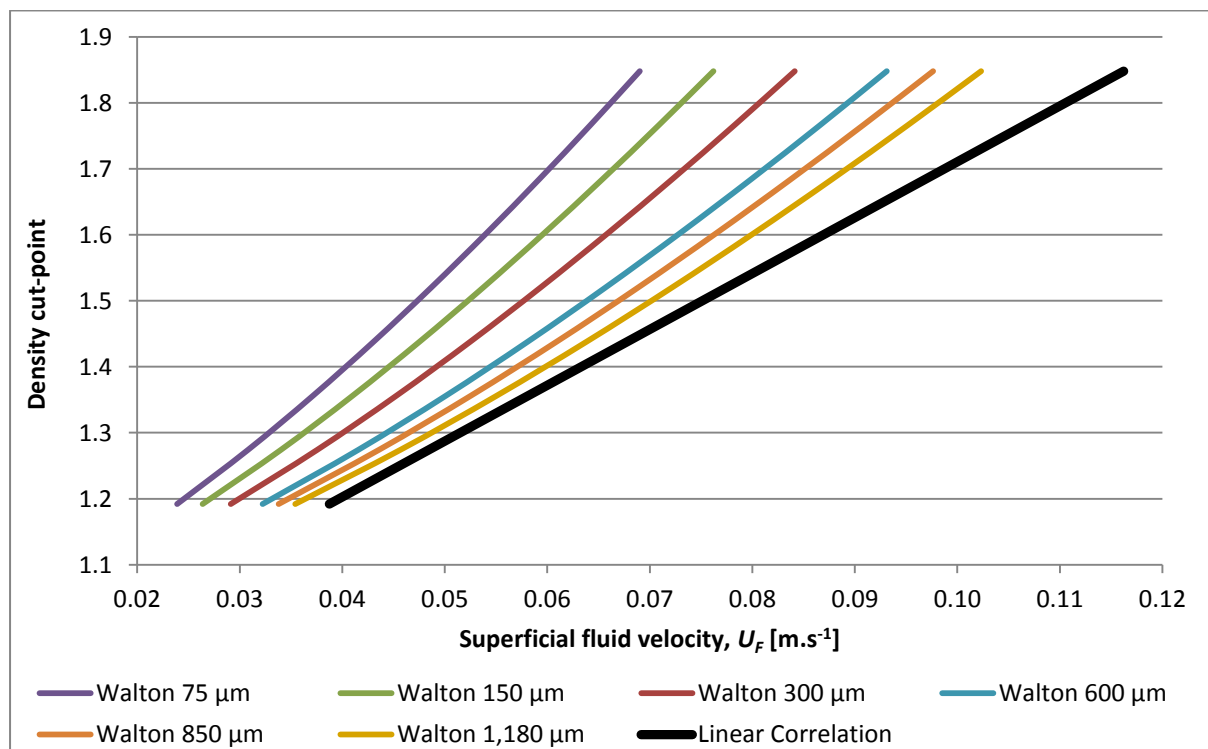


Figure 4-37: Comparing correlations to Walton (2011): $z = 0.0032$ m

Since the correlation is not entirely linear, it is advisable to plot the new correlation to the fractionation data. The correlation from Walton (2011:68) fits the data just as well as the linear correlation.

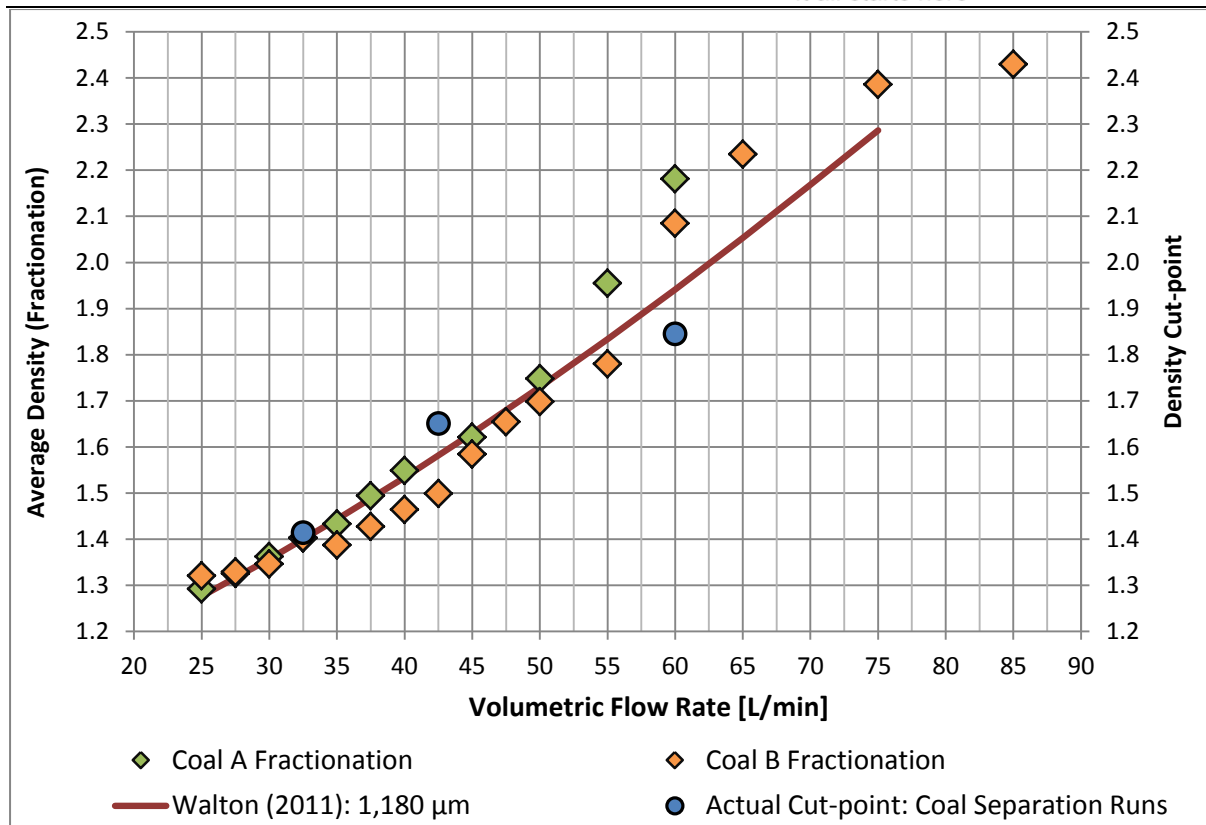


Figure 4-38 Fitting Walton's (2011:68) equation to fractionation data

The break-away from the correlation at the developed stage of the fractionation procedures can with more certainty now be confirmed as real. The reason this occurs still needs to be investigated, because it cannot be attributed to the fact that only heavy particles are left at these stages with the assumption that most of the heavy high ash particles are fine particles. Most of the fine particles are lighter as seen in Figure 4-29. Although both coal sources show this phenomenon when the solids concentration is roughly 20%, a low solids concentration should contribute to a negative break-away.

It can also be confirmed that the density cut-point to fluidisation flow rate is almost linear.

CHAPTER 5 - Closing Remarks

5.1 Conclusions

Cut-density and volumetric flow rate correlation

The first part of the experimental work was aimed at determining the relationship between cut-point and fluidisation flow rate. Through a few fractionation test runs it was shown that the relationship was linear until a certain point was reached. It is postulated that the break-away from linearity was due to a too low solids concentration within the fluidisation section. This was confirmed by a discrete run at 60 L/min. This point was within the inconclusive zone of non-linearity, according to Figure 4-11, but when the line was extended the cut-point of this discrete run fell precisely on this linear relationship.

It was then deduced that during fractionation test work the density to volumetric flow rate correlation can be used to determine the density of each fraction, as long as the solids concentration is 2% and higher by volume.

It was also shown that particle size did not have much of an influence on the relationship, but infinite dilution did indeed. The effect of particle size was not explicitly tested, but observed by plotting the cut-points obtained by tracers and coal of the same run against the volumetric flow rate. The shift in the relationship was small, and can be seen in Figure 4-11.

Two scenarios were also observed during the tracer tests, and are designated as either an infinite dilution system or a hindered settling system. The relationships of volumetric flow rate and cut-density between these two scenarios were parallel, but a higher fluid superficial velocity was needed to convey the particles to the overflow as opposed to the case of hindered settling. This is due to the lower settling velocity of particles in a hindered settling regime. Tracers only signify a dilute system, while tracers charged along with coal signify a hindered settling system.

All of the discrete runs were also accurately predicted by this relationship. It was also noted that one has to compare either cumulative or fractional average densities when correlating densities in overflow to fluidisation flow rate – where tracer tests give density cut-point. Tracer tests must then be converted to cumulative average density to overflow during fractionation runs.

When the linear relationship was compared to Walton (2011:68), it was found that the two were in agreement and described the relationship between cut-density and volumetric flow rate well. The correlation of Walton (2011:68) is not entirely linear and is slightly curved. It still fits the linear model. It was found that equation 4.2 fits the data when the top size of the material is used.

Coal washability

It was shown that the laboratory reflux classifier unit of the North-West University successfully generated washability curves. This is a major breakthrough, as it has many advantages. The main advantage of this operation is the avoidance of the safety and environmental issues with heavy liquids in dense-medium separation, and the fact that the operation is a water-only operation. There is also no need to determine every fraction's density, because the density is linked to the volumetric flow rate via the relationship. The samples only have to be dried and weighed, and then the densimetric curve can be drawn. Only when determining ash and calorific value do analyses need to be done.

RC105 to RC300 data up-scaling

The current configuration of the RC105 is not suitable for up-scaling data to the RC300, due to the current setup not handling continuous operations. The following discussion will explain the operating variables during pilot testing. It will address which issues made the configuration unsuitable. Some extensive semi-batch test runs were done, but were omitted due to their not adding any value to the dissertation.

The RC300 ran at $4.2 \text{ m}^3/\text{h}$ water fluidisation rate, which correlated to 18.9 L/min flow rate when scaled to the RC105 system, and by keeping the Reynolds number equal for both systems. Our unit can give a constant fluidisation water volumetric flow rate of 100 L/min , where 93.4 L/min is the minimum to remain within the laminar flow regime. Flow rate is not the constraint. (Refer to sections B.2 and B.4 of Appendix B for detailed calculations.)

The other variable during the pilot unit (RC300) test work was bed density, and varied at 1.2 and 1.3 specific gravity. The bed density was controlled at these specific gravities by varying the discard flow rate.

During the laboratory test work on the RC105 semi-batch tests relative densities obtained within the fluidised section varied between 1.02 and 1.29. Thus, obtaining similar relative densities was not an issue.

Maintaining a constant specific gravity within the fluidised section proved problematic – operationally and for controllability – due to the following:

- There was no way to determine the bed density without upsetting the system.
- Initially it was reckoned that the bed density could be controlled visually, but this proved unsuccessful.
- The manometer installed was also insufficient to measure the bed density.
- It is believed that a higher throughput could have been obtained, but a high throughput complicated the idea of controlling the system even further. It was then decided to keep the feed flow rate constant throughout the test work at close to 5 L/min .

- Another major factor contributing to the argument, is that it was later discovered that the fluidised bed section was not long enough, which made it difficult to reach stable bed conditions.
- Regardless, a proper control system must be considered in addition to lengthening the fluidised bed section.

The RC105 could then not be operated continuously, and was never under steady state conditions. Although the correlation between laboratory generated data and pilot test data was the main subject of the investigation, other useful data and findings were generated and discovered. These are explained in the above sections. Had more time been available, these findings could have been pursued in more detail.

Section 5.2 gives some recommendations for more effective semi-batch operations.

5.2 Recommendations

RC105 to RC300 up-scaling test work

To ensure continuous operation, the bed density within the fluidised bed section needs to be measured and controlled without interrupting the operation. The material should also be recirculated to ensure enough time is allowed for it to reach steady-state. A new process flow diagram should be designed, and possibly a new RC unit built. A high solids throughput is also desired.

Other concerns are: the short fluidisation section, and the method of measuring the bed density. The fluidised bed section length needs to be increased to allow for a steady bed to form, space for feeding the unit continuously and to space for refluxing. With the increased length it will also be easier to measure the bed density with the purpose of controlling the underflow.

Since there are two components fed into the system (water and coal slurry), the excess water needs to be removed before the coal slurry is recycled back to the slurry holding tank. This will also help to remove fines before the fluidisation water is recycled back to the fluidisation water tank. A lamella clarifier can be sized similar to what is shown in Figure 5-1.

In continuous operation the idea is to have a submersible pump that recycles back into the clarifier box (to keep the solids in suspension and controlling the flow rate of the stream going to the slurry holding tank). The main stream is then pumped back to the slurry holding tank. When the flow rate of the clear water (clarifier overflow) is the same as the fluidisation water flow rate read off on the rotameter feeding the RC, the system is in balance. For this process, both the overflow and underflow need to be discharged into the clarifier box.

In batch processes the clarifier can be used when very fine material is present in the sample treated, and no bottom size constraint will exist. The submersible pump may be removed

for the batch processes, and the solids collected (which now reside within the clarifier box) after a run is complete.

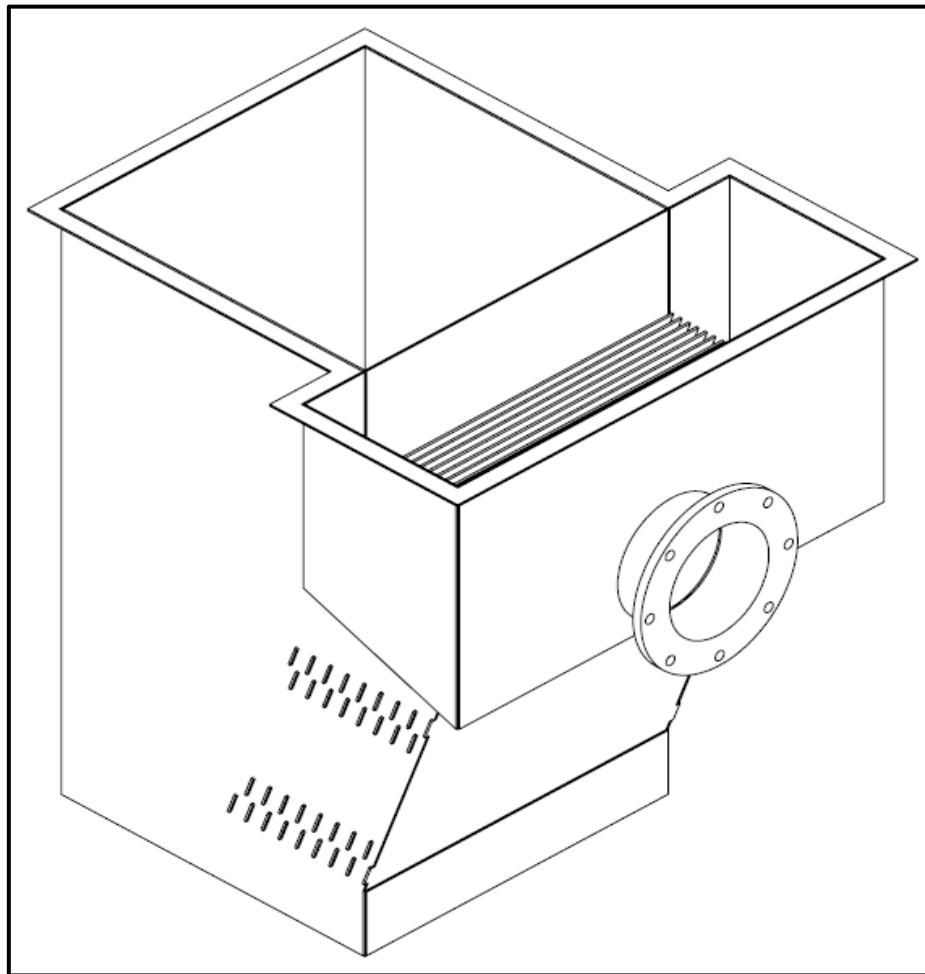


Figure 5-1: Water clarifier concept

When sizing the clarifier, equation 4.2 can be used. A 6 mm plate spacing is still considered narrow, and is more comfortable to work with. The lightest and smallest particles should also be captured within the clarifier and not leave through the overflow. Using equation 4.2 the maximum water superficial velocity that can be reached within the plates is 0.03 m/s for a 5 μm particle that has a density of 1200 kg/m^3 . If the volumetric flow rate is 90 L/min, the horizontally projected area should be 0.048 m^2 . This corresponds to an open area of 0.350 m X 0.115 m if an oversize factor of 10% is used. This gives 25 channels if the channel spacing is 6 mm, and with a plate thickness of 1 mm, the horizontally projected area of the plates section will be 350 mm X 177 mm. One can even do it the other way around and make the 350 mm wide side the slotted/slit slide, and have narrower plates, which are more effectively supported within the slits.

Regarding the control of the unit, the bed density should be measured by means of a differential pressure gauge or gamma diffraction (which is not the best option). The discard should then be controlled by an actuated pinch or rotating door valve. The idea of a small

peristaltic pump is also not completely absurd. The measurement of the density within the fluidised bed is the most important though, and the pinch valve may be controlled by hand. The gate valve should be removed on both the discard stream and the slurry holding tank.

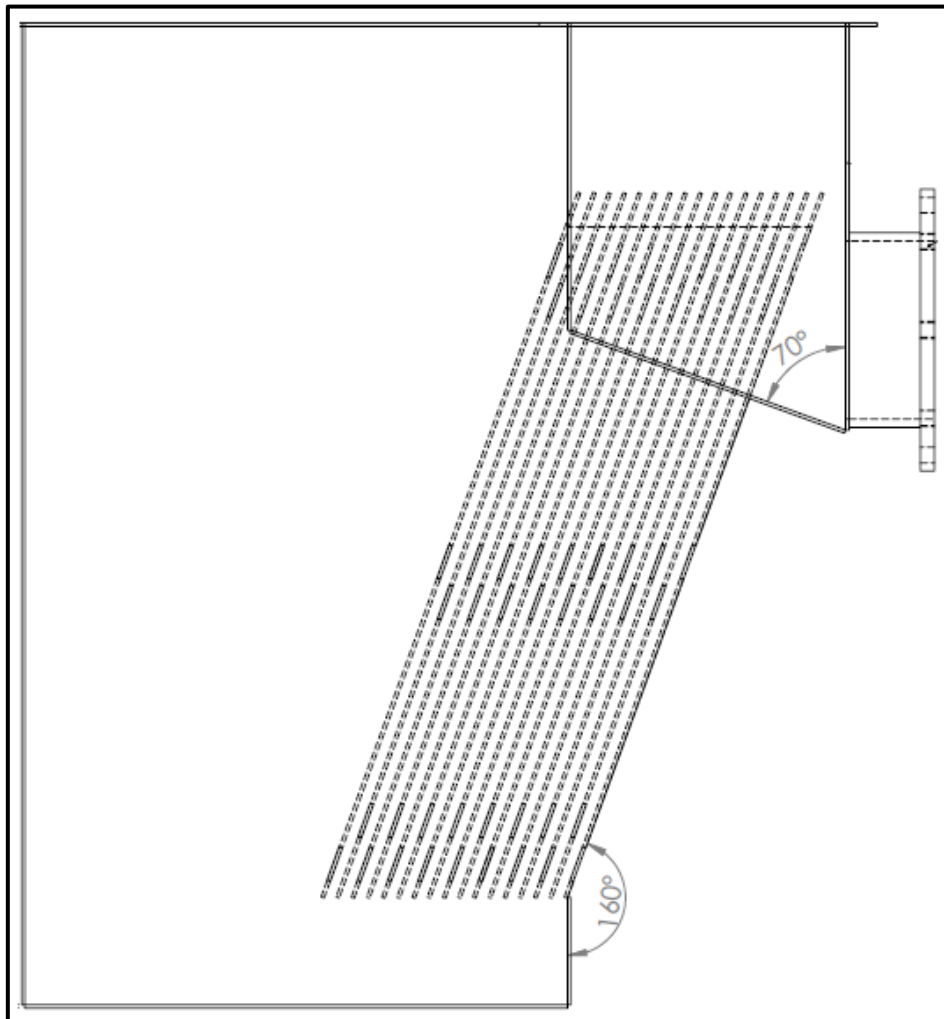


Figure 5-2: Water clarifier concept: side view

The slurry holding tank should be baffled to ensure better mixing.

The investigation into a correlation between laboratory and pilot generated data can still be done if the experimental setup is optimised to run continuously.

Cut-density and volumetric flow rate correlation test work

A new idea has presented itself: that, when doing batch test work (low bed density), equation 4.2 can be used as the cut-density and volumetric flow rate correlation when the particle size that is plugged in to equation 4.2 is the top-size of the material tested. If for batch test work the particle top-size as the particle size can be used with equation 4.2, this correlation will work in determining the densities during washability analysis. It is also recommended to fill the unit with 1 kg of coal sample to ensure that the 2% solids concentration by the volume of the fluidised bed section occurs at the heavy solids particles.

Closing Remarks

Since at high bed densities (1.2 to 1.3) lower volumetric flow rates are needed to convey the particles to the overflow, it should be investigated whether the d_{50} of the particle size range of the material needs to be used. Equation 4.2's sensitivity to a constant bed density should also be investigated.

References

- ALLEN, T. & KHAN, A.A. 1970. Critical evaluation of power sampling procedures. *The chemical engineer*. 238:108-112
- ALLEN, T. 2003. Powder sampling and particle size determination. 1st ed. Amsterdam: Elsevier B.V.
- BOYCOTT, A.E. 1920. Sedimentation of blood corpuscles. *Nature*, 104:532.
- CALLEN, A., MOGHTADERI, B. & GALVIN, K.P. 2007. Use of parallel inclined plates to control elutriation from a gas fluidized bed. *Chemical engineering science*, 62:356-370.
- CALLEN, A.M., PATEL, B., ZHOU, J. & GALVIN, P. 2008. Development of water-based methods for determining coal washability data. *International journal of coal preparation and utilisation*, 28:33-50.
- CARRIS, D.M. 2007. A historic perspective. (In Arnold, B.J., Klima, M.S. & Bethell, P.J., ed. Designing the coal preparation plant of the future. 1st ed. Littleton, CO: Society for mining, metallurgy, and exploration Inc. p. 3-8)
- DARBY, R. 2001. Chemical engineering fluid mechanics. 2nd ed. New York, NY: Marcel Dekker Inc.
- DE KORTE, G.J. 2014. The current focus for R&D in the South African environment. 15th Australian coal preparation society Conference: Queensland, Australia.
- DE KORTE, G.J. 2002. Dense-medium beneficiation of fine coal revisited. *The journal of the Southern African institute of mining and metallurgy*, October: 393-396.
- DE KORTE, J. & BOSMAN, J. 2007. Optimal coal processing route for the 3 x 0.5 mm size fraction? *The journal of the Southern African institute of mining and metallurgy*, 107:411-414.
- DOMAN, L.E. & ARORA, V. 2013. World energy demand and economic outlook. (In Conti, J., Holtberg, P., Beamon, J.A., Napolitano, S., Schaal, A.M., Turnure, J.T., Westfall, L & Doman, L.E., ed. International energy outlook 2013. July. Washington, DC: U.S. energy information administration. p. 9-21)
- DOROODCHI, E., FLETCHER, D.F. & GALVIN, K.P. 2004. Influence of inclined plates on the expansion behaviour of particulate suspension in a liquid fluidised bed. *Chemical engineering science*, 59:3559-3567.

References

- DOROODCHI, E., ZHOU, J., FLETCHER, D.F & GALVIN, K.P. 2006. Particle size classification in a fluidized bed containing parallel inclined plates. *Minerals engineering*, 19:162-171.
- EBERHARD, A. 2011. The future of South African coal: Market, investment, and policy challenges. *Freeman Spogli institute for international studies*. Last accessed: 8 September 2012. http://iis-db.stanford.edu/pubs/23082/WP_100_Eberhard_Future_of_South_African_Coal.pdf.
- ENGLAND, T. 2002. Coal preparation in South Africa. 4th ed. Pietermaritzburg: The South African coal processing society.
- FALCON, R. & HAM, A.J. 1988. The characteristics of Southern African coals. *Journal of the South African institute of mining and metallurgy*, 88:145-161.
- GALVIN, K.P., DOROODCHI, E., CALLEN, A.M., LAMBERT, N. & PRATTEN, S.J. 2002. Pilot plant trial of the reflux classifier. *Minerals engineering*, 15:19-25.
- GALVIN, K.P & NGUYENTRANLAM, G. 2002. Influence of parallel inclined plates in a liquid fluidised bed system. *Chemical engineering science*, 57:1231-1234.
- GALVIN, K.P., CALLEN, A., ZHOU, J. & DOROODCHI, E. 2005. Performance of the reflux classifier for gravity separation at full scale. *Minerals engineering*, 18:19-24.
- GALVIN, K.P., WALTON, K. & ZHOU, J. 2009. How to elutriate particles according to their density. *Chemical engineering science*, 64:2003-2010.
- GALVIN, K.P., ZHOU, J. & WALTON, K. 2010a. Application of closely spaced inclined channels in gravity separation of fine particles. *Minerals engineering*, 32:326-338
- GALVIN, K.P., CALLEN, A.M. & SPEAR, S. 2010b. Gravity separation of coarse particles using the reflux classifier. *Minerals engineering*, 23:339-349.
- GALVIN, K.P., CALLEN, A., SPEAR, S., WALTON, K. & ZHOU, J. 2010c. Gravity separation of coal in the reflux classifier – new mechanism for suppressing the effects of particle size. *International journal of coal preparation and utilization*, 30:130-144
- GALVIN, K.P. & LIU, H. 2011. Role of inertial lift in elutriating particles according to their density. *Chemical engineering science*, 66:3687-3691.
- IEA STATISTICS. 2010. Coal information. *International energy agency*.
- IVESON, S.M., MASON, M. & GALVIN, K.P. 2014. Gravity separation and desliming of fine coal: pilot plant study using reflux classifiers in series. *International journal of coal preparation and utilization*, 34(5):239-259.
- KELLY, E.G. & SPOTTISWOOD D.J. 1989. Introduction to mineral processing. 1st ed. Denver: John Wiley & Sons, Inc.

References

- KING, M.R. & LEIGHTON Jr., D.T. 1997. Measurement of the inertial lift on a moving sphere in contact with a plane wall in a shear flow. *American institute of physics – Fluids*, 9(5):1248-1255.
- KOHMUENCH, J.N., MANKOSA, M.J. & LUTTRELL, G.H. 2007. Fine coal cleaning: A review of column flotation options and design considerations. (In Arnold, B.J., Klima, M.S. & Bethell, P.J., ed. Designing the coal preparation plant of the future. 1st ed. Littleton, CO: Society for mining, metallurgy, and exploration Inc. p. 97-117)
- LASKOVSKI, D., DUNCAN, P., STEVENSON, P., ZHOU, J. & GALVIN, K.P. 2006. Segregation of hydraulically suspended particles in inclined channels. *Chemical engineering science*, 61:7269-7278.
- LUNDT, M. & DE KORTE, G.J. 2010. Leeuwpan fine coal dense medium plant. *The journal of the Southern African institute of mining and metallurgy*, 10:671-676
- LUTTRELL, G.H., HONAKER, R.Q., BETHELL, P.J. & STANLEY, F.L. 2007. Design of high-efficiency spiral circuits for preparation plants. (In Arnold, B.J., Klima, M.S. & Bethell, P.J., ed. Designing the coal preparation plant of the future. 1st ed. Littleton, CO: Society for mining, metallurgy, and exploration Inc. p. 73-88)
- MACPHERSON, S.A., IVESON, S.M. & GALVIN, K.P. 2010. Density based separations in the reflux classifier with an air-sand dense-medium and vibration. *Minerals engineering*, 23:74-82.
- MACPHERSON, S.A. 2011. Dry beneficiation in a reflux classifier. Newcastle, AU: University of Newcastle. (Thesis – PhD)
- MACPHERSON, S.A., IVESON, S.M. & GALVIN, K.P. 2011. Density-based separation in a vibrated reflux classifier with an air-sand dense-medium: Tracer studies with simultaneous underflow and overflow removal. *Minerals engineering*, 24:1046-1052.
- MUNROE, H.S. 1888. The English verses the Continental system of jigging – is close sizing advantageous? *Transactions of the American institute of mining, metallurgical and petroleum engineers*. 17:637-659.
- NGUYENTRANLAM, G. & GALVIN, K.P. 2001. Particle classification in the reflux classifier. *Minerals engineering*, 14:1081-1091.
- NGUYENTRANLAM, G. & GALVIN, K.P. 2004. Application of the reflux classifier in solid-liquid operations. *International journal of mineral processing*, 73:83-89.
- REDDICK, J.F., VON BLOTTNITZ, H. & KOTHUIS, B. 2007. A cleaner production assessment of the ultra-fine coal waste generated in South Africa. *The journal of the Southern African institute of mining and metallurgy*, 107:811-816

References

- RHODES, M. 2008. Introduction to particle technology. 2nd ed. Wiltshire: John Wiley & Sons, Inc.
- RICHARDSON, J.F. & ZAKI, W.N. 1954. The sedimentation of a suspension of uniform spheres under conditions of viscous flow. *Chemical engineering science*, 8:65-73.
- SALAH, M.A. 2012. Cooling tower blowdown treatment using an inclined plate clarifier. *PennWell Corporation*. <http://www.waterworld.com/articles/iww/print/volume-7/issue-1/features/cooling-tower-blowdown-treatment-using-an-inclined-plate-clarifier.html>. Last accessed: 8 September 2012.
- SVAROVSKY, L. 2000. Solid-liquid separation. 4th ed. Woburn, MA: Butterworth-Heinemann.
- TILTON, J. N. 1997. Section 6: Fluid and particle dynamics. (In Perry, R.H., Green, D.W. & Maloney, J.O., ed. Perry's chemical engineers' handbook. 7th Ed. New York, NY: McGraw-Hill Companies, Inc.)
- VAN DER WALT, P.J. 2003. Comment on G.J. de Korte's 'Dense-medium beneficiation of fine coal revisited'. *The journal of the Southern African institute for mining and metallurgy*, 102:73.
- VANCE, W.H. & MOULTON, R.W. 1965. A study of slip ratios for the flow of steam-water mixtures at high void fractions. *American institute of chemical engineering*, (November: 1114-1124).
- WALTON, K., ZHOU, J. & GALVIN, K.P. 2010. Processing of fine particles using closely spaced channels. *Advanced powder technology*, 21(4):386-391.
- WALTON, K.J. 2011. Influence of narrow inclined channels on fine particle separations. Newcastle, AU: The University of Newcastle. (Thesis – PhD)
- WELTY, J.R., WICKS, C.E., WILSON, R.E. & RORRER, G.L. 2008. Fundamentals of momentum, heat, and mass transfer. 5th ed. Danvers, MA: John Wiley & Sons, Inc.
- WILLS, B.A & NAPIER-MUNN, T.J. 2006. Wills' mineral processing technology. 6th ed. Brisbane: Elsevier Science & Technology Books.
- ZHANG, X. & DAVIS, R.H. 1990. Particle classification using inclined settlers in series and with underflow recycle. *Industrial engineering and chemical research*, 29:1894-1900.
- ZHOU, J., WALTON, K., LASKOVSKI, D., DUNCAN, P. & GALVIN, K.P. 2006. Enhanced separation of mineral sands using the reflux classifier. *Minerals engineering*, 19:1573-1579.
- ZIGRANG, D.J. & SYLVESTER, N.D. 1981. An explicit equation for particle settling velocities in solid-liquid systems. *American institute of chemical engineering*, 27(6):1043-1044.

Appendices

APPENDIX A - RC105 Design Drawings	102
A.1 The Plates Section	102
A.2 The Fluidisation and Feeding Chamber.....	104
A.3 The Overflow Bucket and Flow Distribution sections.....	105
APPENDIX B - Detailed Calculations	106
B.1 Fluidisation Water Required	106
B.2 Reynolds Number	107
B.3 Pressure Drop.....	108
B.4 Scaling calculations – Between RC300 and RC105.....	108
APPENDIX C - Photos of the RC105 Rig.....	110
APPENDIX D - Partition Curves.....	118
D.1 Tromp curves for diluted tracers only system	118
D.2 Tromp curves for coal A and tracer system	122

APPENDIX A - RC105 Design Drawings

A.1 The Plates Section

- The slits were cut 1 mm wide in order for the 0.9 mm thick aluminium plates to easily slide into the slits.
- To obtain a 3.50 mm plate spacing, the slit spacing is 3.40 mm and at the end sides (where PVC sections are one of the channel barriers), the slit spacing is 3.45 mm.
- The length of the part was kept at 1.250 m, because the assembly will be cut to 1.000 m and at an angle of 70° to the horizon.
- Figure A-1 shows the plate supporting PVC side with the 1 mm wide slits spaced 3.40 mm apart.
- Figure A-2 shows how the two plate supporting PVC sides and the remaining two sides were glued together to form the capsule assembly – still 1.250 m in length before it is cut to 70°. The 10 mm thick PVC parts were glued to a 10 mm step (depicted in Figure A-1 and Figure A-2) in the plate supporting sides.
- Figure A-3 shows the plates section capsule after it has been cut to 1.000 m and 70° to the horizon.

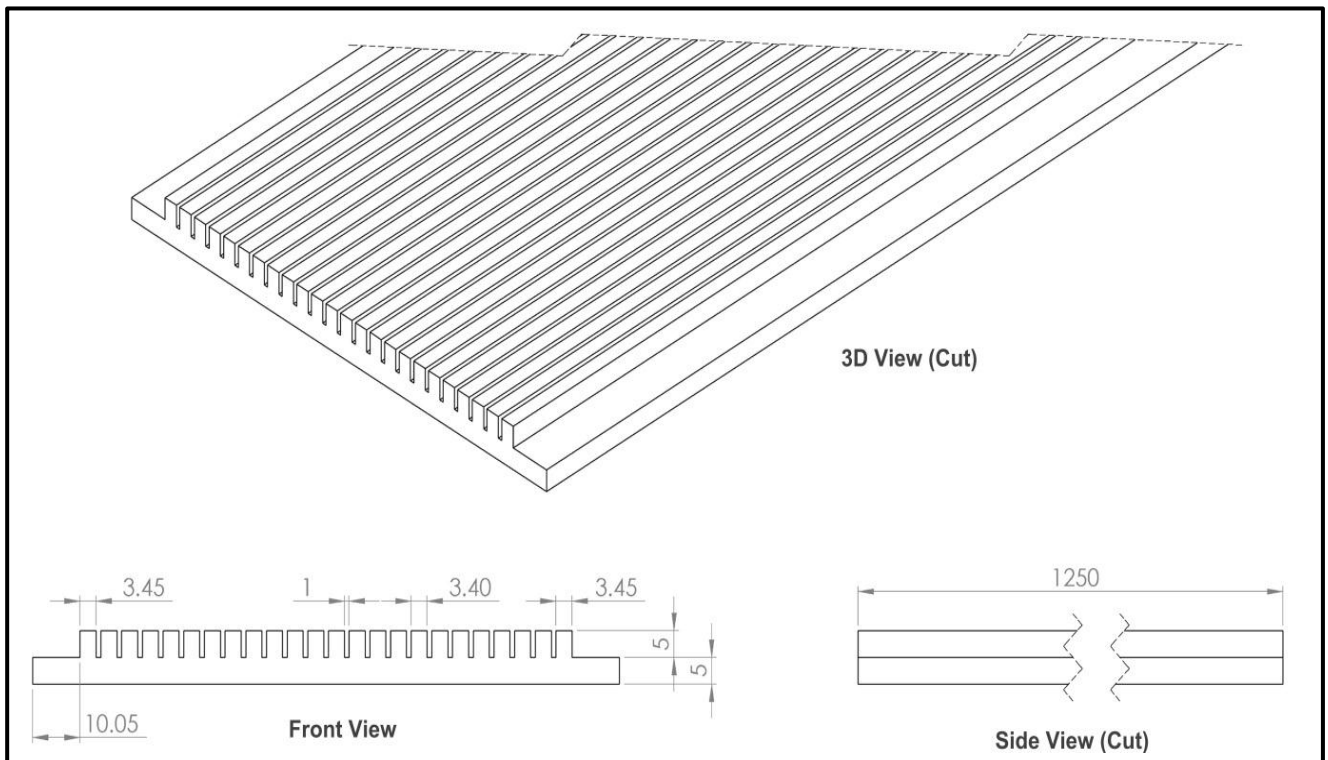


Figure A-1: Detailed drawing of the plates supporting PVC side of the plates section

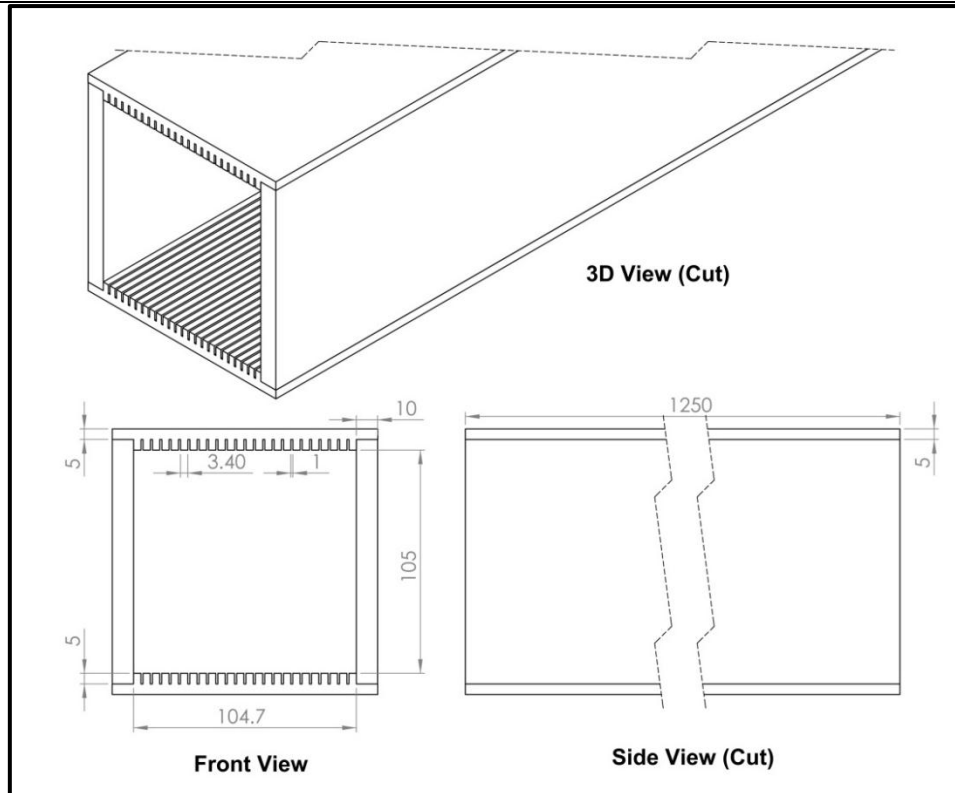


Figure A-2: Detailed drawing of the 3.5 mm channel capsule before cutting to 70°

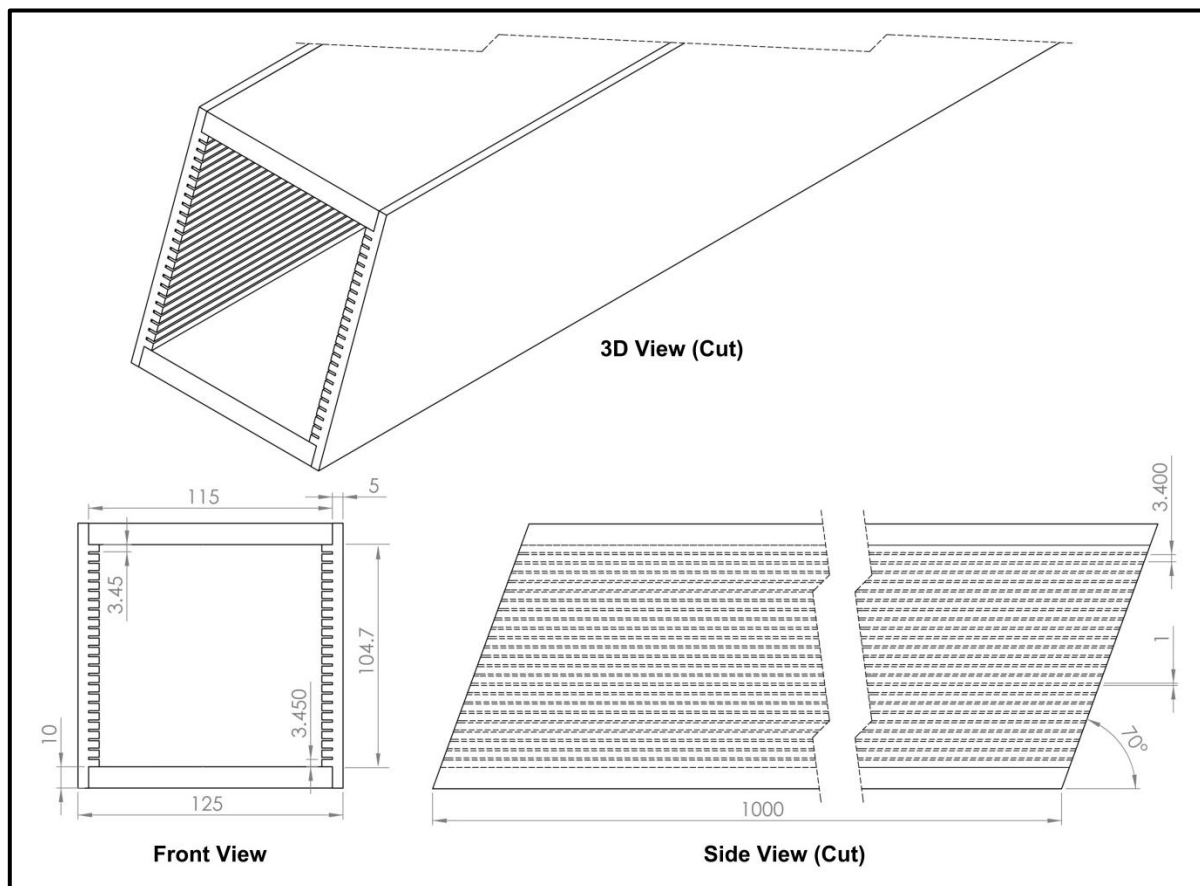


Figure A-3: Detailed drawing of the plate section cut to 1 m and 70° to the horizon

A.2 The Fluidisation and Feeding Chamber

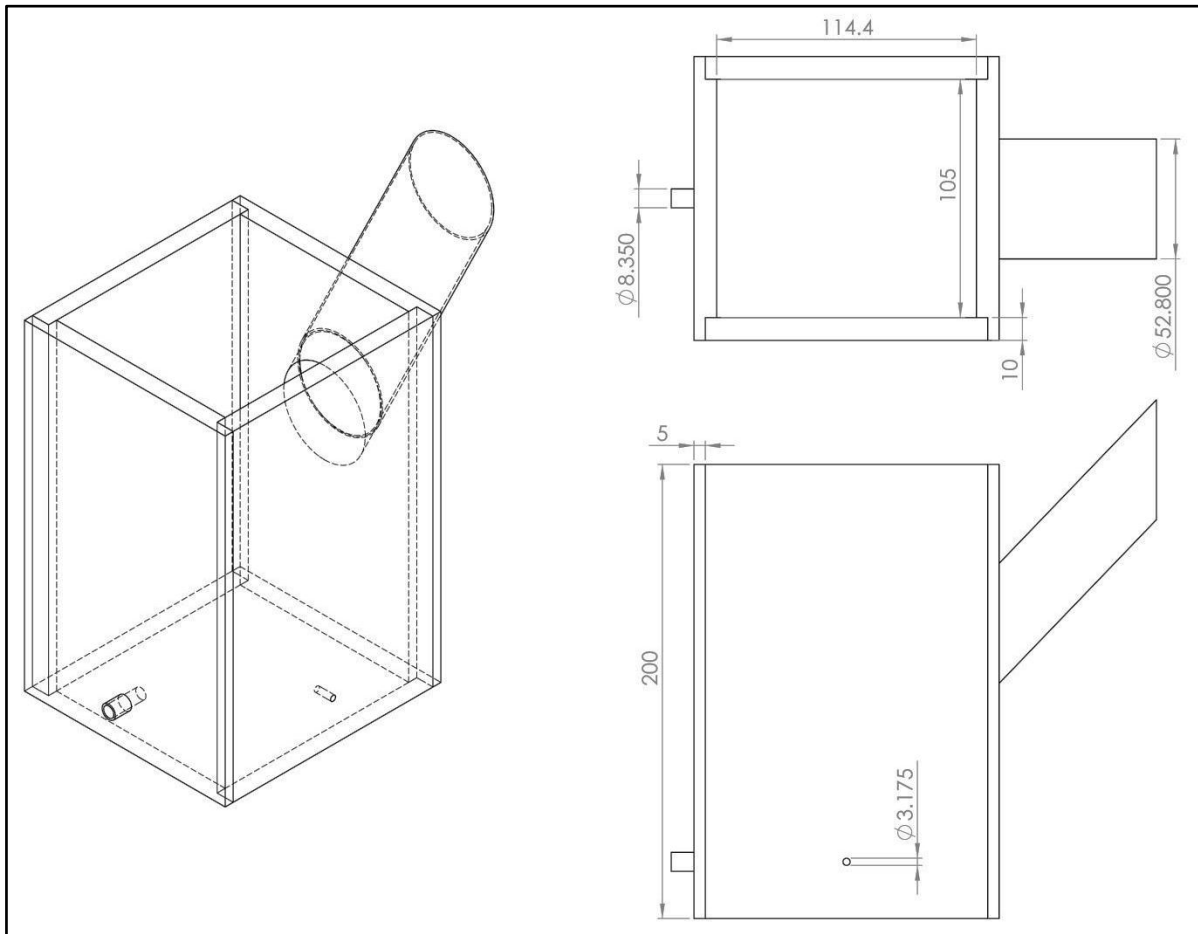


Figure A-4: Detailed drawing of the fluidisation and feeding port section

A.3 The Overflow Bucket and Flow Distribution sections

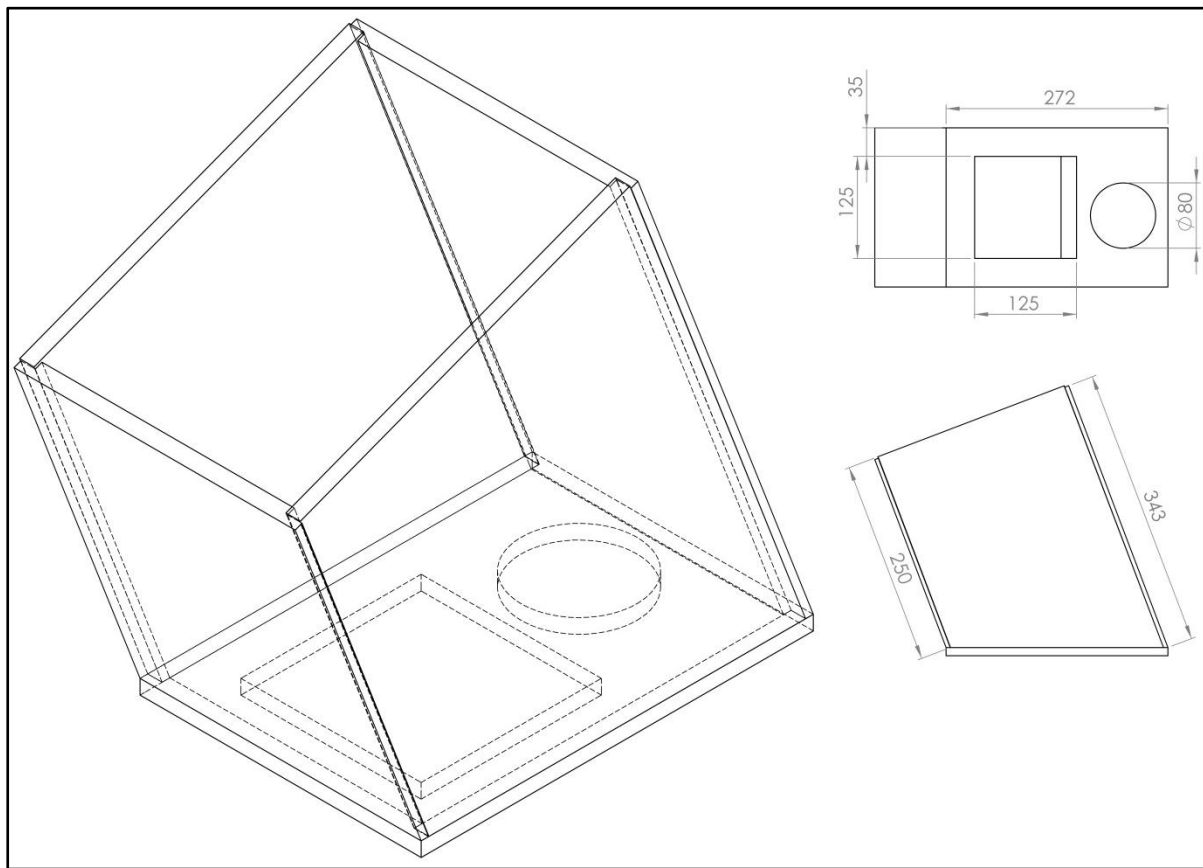


Figure A-5: Detailed drawing of the overflow bucket

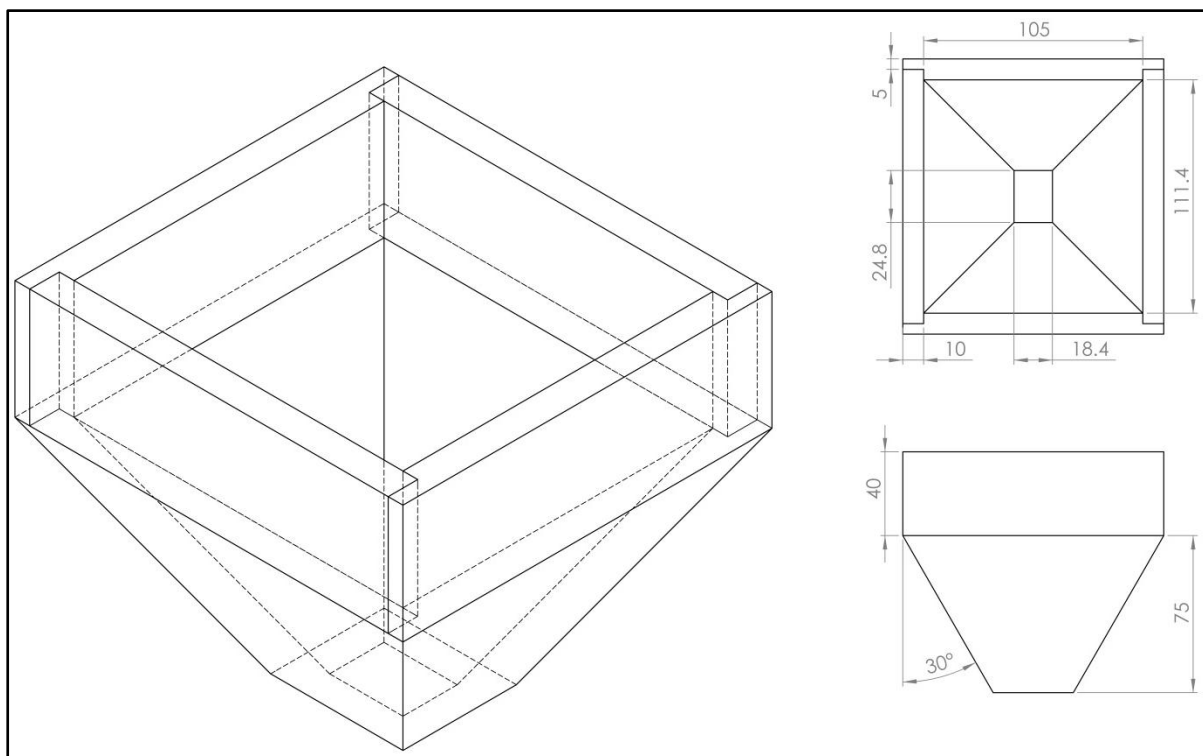


Figure A-6: Detailed drawing of the flow distribution section

APPENDIX B - Detailed Calculations

B.1 Fluidisation Water Required

- RC600 cross-sectional area: $0.60 \times 0.60 = 0.360 \text{ m}^2$
- RC600 open cross-sectional area: $0.36 \times 0.70 = 0.252 \text{ m}^2$

The open cross-sectional area was determined by multiplying the assumed open space fraction as given in section 3.1.2. $A_{O,RC600} = 0.252 \text{ m}^2$

- Solids throughput: $45 \text{ t} \cdot \text{m}^{-2} \cdot \text{h}^{-1}$
- Solids mass flow: $45 \times 0.252 = 11.34 \text{ t} \cdot \text{h}^{-1}$
- Solids volumetric flow rate: $11,340 \div 1,500 = 7.56 \text{ m}^3 \cdot \text{h}^{-1}$
- Water volumetric flow rate: $7.56 \times 0.8 \div 0.2 = 30.24 \text{ m}^3 \cdot \text{h}^{-1}$
- Superficial fluid velocity required: $30.24 \text{ m}^3 \cdot \text{h}^{-1} \div 0.252 \text{ m}^2 = 120 \text{ m} \cdot \text{h}^{-1}$
- Superficial fluid velocity required: $0.033 \text{ m} \cdot \text{s}^{-1}$

The superficial fluidisation water velocity required was determined by calculating the mass flow rate of the solids and then using the average density assumed for the coal to determine the volumetric flow rate of the solids. The water volumetric flow rate was determined by using the assumed solids concentration within the plates. The velocity is simply the volumetric flow rate divided by the open area of the plates section. $U_{RC600} = 0.033 \text{ m} \cdot \text{s}^{-1}$

- RC600 equivalent diameter (equation 2.13): $D_{E,RC600} = 0.026 \text{ m}$
 - $a = 0.600 \text{ m}$
 - $b = 0.006 \text{ m}$
 - $a/b = 100$
 - $K = 12$

$$D_E = \left(\frac{128ab^3}{\pi K} \right)^{1/4} \quad 2.13$$

- Reynolds number for operating conditions in RC600 (equation 2.3): $Re = 858$
 - $U_{RC600} = 0.033 \text{ m} \cdot \text{s}^{-1}$
 - $D_E = 0.026 \text{ m}$
 - $\rho_w = 1,000 \text{ kg} \cdot \text{m}^{-3}$
 - $\mu_w = 0.001 \text{ kg} \cdot \text{m}^{-1} \cdot \text{s}^{-1}$

$$Re = \frac{DU\rho_p}{\mu} \quad 2.3$$

APPENDIX B - Detailed Calculations

When scaling systems, both systems must operate at the same Reynolds number and the systems' geometries must be the same. Thus, the superficial fluidisation water velocity required in the RC105 (U_{RC105}) will be determined by using $Re = 858$ and the equivalent diameter of the RC105.

- RC105 equivalent diameter (equation 2.13): $D_{E,RC105} = 0.011 \text{ m}$
 - $a = 0.105 \text{ m}$
 - $b = 0.0035 \text{ m}$
 - $a/b = 30$
 - $K = 12$
- Using $Re_{RC600} = Re_{RC105} = 858$ (equation 2.3): $U_{RC105} = 0.078 \text{ m} \cdot \text{s}^{-1}$
 - $Re = 858$
 - $D_{E,RC105} = 0.011 \text{ m}$
 - $\rho_w = 1,000 \text{ kg} \cdot \text{m}^{-3}$
 - $\mu_w = 0.001 \text{ kg} \cdot \text{m}^{-1} \cdot \text{s}^{-1}$
- RC105 cross-sectional area: $0.105 \times 0.1047 = 0.011 \text{ m}^2$
- RC105 open cross-sectional area: $0.011 - (23 \times 0.0009 \times 0.105) = 0.009 \text{ m}^2$
- Fluidisation volumetric flow rate required: $0.009 \times 0.078 = 6.88 \times 10^{-4} \text{ m}^3 \cdot \text{s}^{-1}$
- Fluidisation water volumetric flow rate required: $41.3 \text{ L} \cdot \text{min}^{-1}$

B.2 Reynolds Number

The Reynolds number for a fluidisation rate of 50 L/min and the maximum flow rate before entering the transitional flow regime are calculated:

- Critical Reynolds number for rectangular conduits: $1900 < Re_c < 2800$
- Fluidisation volumetric flow rate: $50 \text{ L} \cdot \text{min}^{-1} \div 1000 \div 60 = 8.33 \times 10^{-4} \text{ m}^3 \cdot \text{s}^{-1}$
- U_{50LPM} in RC105: $8.33 \times 10^{-4} \text{ m}^3 \cdot \text{h}^{-1} \div 0.009 \text{ m}^2 = 0.094 \text{ m} \cdot \text{s}^{-1}$
- $Re_{50LPM} = 1034$
 - $U_{50LPM} = 0.094 \text{ m} \cdot \text{s}^{-1}$
 - $D_{E,RC105} = 0.011 \text{ m}$ (Determined as in section B.1)
 - $\rho_w = 1,000 \text{ kg} \cdot \text{m}^{-3}$
 - $\mu_w = 0.001 \text{ kg} \cdot \text{m}^{-1} \cdot \text{s}^{-1}$
- $U_{MAX} = 0.173 \text{ m} \cdot \text{s}^{-1}$
 - $Re_c = 1900$
 - $D_{E,RC105} = 0.011 \text{ m}$ (Determined as in section B.1)
 - $\rho_w = 1,000 \text{ kg} \cdot \text{m}^{-3}$
 - $\mu_w = 0.001 \text{ kg} \cdot \text{m}^{-1} \cdot \text{s}^{-1}$
- Maximum volumetric flow rate: $0.173 \text{ m} \cdot \text{s}^{-1} \times 0.009 \text{ m}^2 = 15.23 \times 10^{-4} \text{ m}^3 \cdot \text{s}^{-1}$
- Maximum volumetric flow rate: $15.23 \times 10^{-4} \text{ m}^3 \cdot \text{s}^{-1} \times 1000 \times 60 = 91.4 \text{ L} \cdot \text{min}^{-1}$

B.3 Pressure Drop

Frictional head

- Frictional pressure drop in one channel (equation 2.9): $\Delta P_{f,3.5} = 0.185 \text{ kPa}$
 - $Q = (99 [\text{l} \cdot \text{min}^{-1}] \div 24 \text{ channels}) \times \left(\frac{1}{1000 \cdot 60}\right) = 6.88 \times 10^{-5} \text{ m}^3 \cdot \text{s}^{-1}$
 - $\mu_w = 0.001 \text{ kg} \cdot \text{m}^{-1} \cdot \text{s}^{-1}$
 - $L_{ch} = 1 \text{ m}$
 - $D_{E,3.5\text{mm}} = 0.011 \text{ m}$
- Frictional pressure drop in plates section: $\Delta P_{f,1.5} = 0.185 \times 24 = 4.44 \text{ kPa}$

$$Q = \frac{\pi \Delta P D^4}{128 \mu L} ; Re \leq 2100 \quad 2.9$$

The equivalent diameter for the 3.5 mm channels was determined as 0.011 m, using equation 2.13. The pressure drop due to friction within one 3.5 mm channel was calculated as 0.170 kPa for a maximum volumetric flow rate of 91.4 L/min. The total pressure drop within the plates section with **24 × 3.5 mm channels is 4.4 kPa.**

Static head

- From reference point 2: $P_2 = 11.42 \text{ kPa}$
 - Gravitational acceleration: $g = 9.81 \text{ m} \cdot \text{s}^{-2}$
 - Plate section height: $h_{PS} = 1 \text{ m} \times \sin 70^\circ = 0.940 \text{ m}$
 - SG within plates section: $\rho_{PS} = 1100 \text{ kg} \cdot \text{m}^{-3}$
 - Static head within plates section: **$P_{PS} = 10.14 \text{ kPa}$**
 - Fluidisation section height: $h_{FS} = 0.100 \text{ m}$
 - SG within fluidisation section: $\rho_{FS} = 1300 \text{ kg} \cdot \text{m}^{-3}$
 - Static head within fluidisation section: **$P_{FS} = 1.28 \text{ kPa}$**

Slurry holding tank height

- Tank height from unit inlet: $H_T = 1.40 \text{ m}$
 - Slurry density (25% solids v/v): $SG_{sl} = [(1.6 \times 25 \text{ l}) + 75] \div 100 = 1.15$
 - $H_T = 15,820 \text{ Pa} \div (9.81 \text{ m} \cdot \text{s}^{-2} \times 1150 \text{ kg} \cdot \text{m}^{-3})$

B.4 Scaling calculations – Between RC300 and RC105

Again, the two systems must have similar geometries and the same operating conditions must be experienced within the systems. The two systems are designed to be geometrically similar. When the Reynolds number within the plates sections is the same, the operating conditions will be the same and the system may be scaled.

- RC300 equivalent diameter (equation 2.13): $D_{E,RC300} = 0.022 \text{ m}$
 - $a = 0.300 \text{ m}$
 - $b = 0.006 \text{ m}$
 - $a/b = 100$

APPENDIX B - Detailed Calculations

- $K = 12$
- The superficial velocity with the plates: $U_{RC300} = 0.018 \text{ m} \cdot \text{s}^{-1}$
 - $A_{O,RC300} = 0.064 \text{ m}^2$
 - Fluidisation water volumetric flow rate: $Q_{W,RC600} = 4.2 \text{ m}^3 \cdot \text{h}^{-1}$
- Re in RC300 (equation 2.3): $Re = 395$
 - $U_{RC300} = 0.018 \text{ m} \cdot \text{s}^{-1}$
 - $D_E = 0.022 \text{ m}$
 - $\rho_w = 1,000 \text{ kg} \cdot \text{m}^{-3}$
 - $\mu_w = 0.001 \text{ kg} \cdot \text{m}^{-1} \cdot \text{s}^{-1}$
- Using $Re_{RC300} = Re_{RC105} = 395$ (equation 2.3): $U_{RC105} = 0.036 \text{ m} \cdot \text{s}^{-1}$
 - $Re = 395$
 - $D_{E,RC105} = 0.011 \text{ m}$
 - $\rho_w = 1,000 \text{ kg} \cdot \text{m}^{-3}$
 - $\mu_w = 0.001 \text{ kg} \cdot \text{m}^{-1} \cdot \text{s}^{-1}$

A superficial fluidisation water velocity of $U_{RC105} = 0.036 \text{ m} \cdot \text{s}^{-1}$ gives a volumetric flow rate of $18.9 \text{ L} \cdot \text{min}^{-1}$. Thus, a fluidisation rate of $18.9 \text{ L} \cdot \text{min}^{-1}$ within the plates section of the RC105 will deliver the same Reynolds number as in the RC300.

APPENDIX C - Photos of the RC105 Rig



Figure C-1: Photo – Entire unit



Figure C-2: Photo – Rotameters and piping



Figure C-3: Photo – Piping

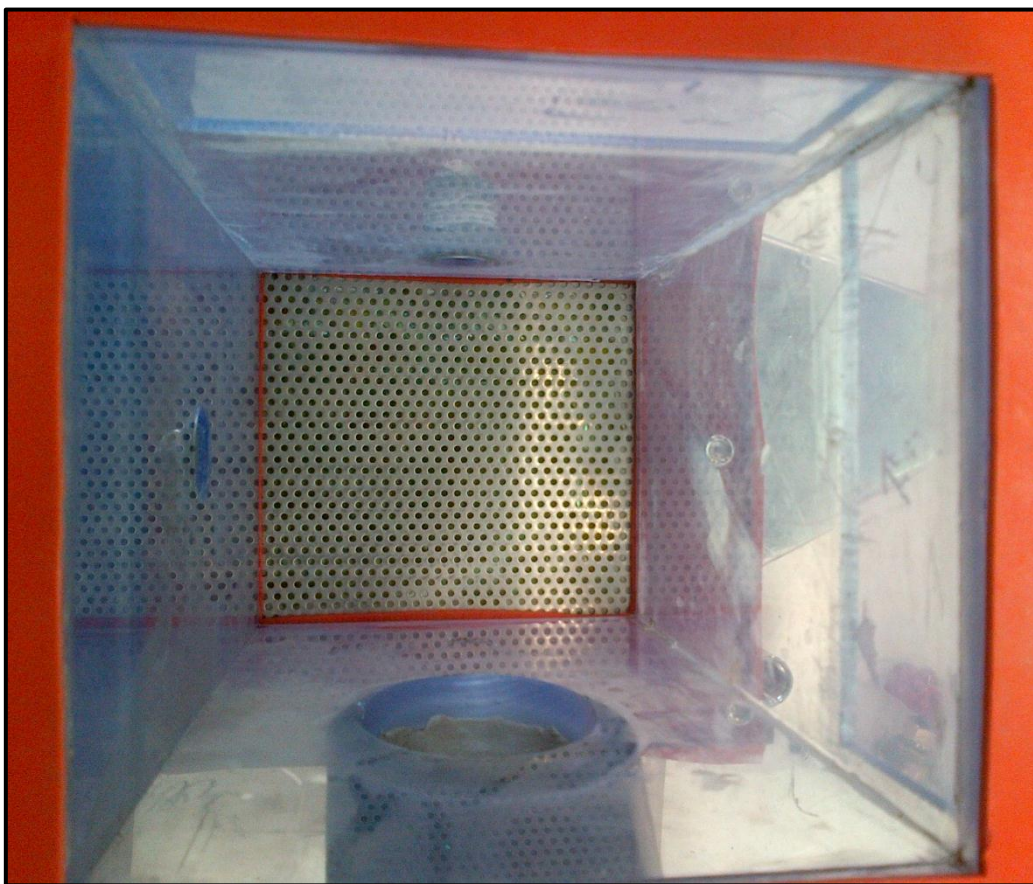


Figure C-4: Photo – Inside of fluidisation section



Figure C-5: Photo – Fluidisation section, discard and manometer



Figure C-6: Photo – Coal capture in semi-batch mode (High quality coal)

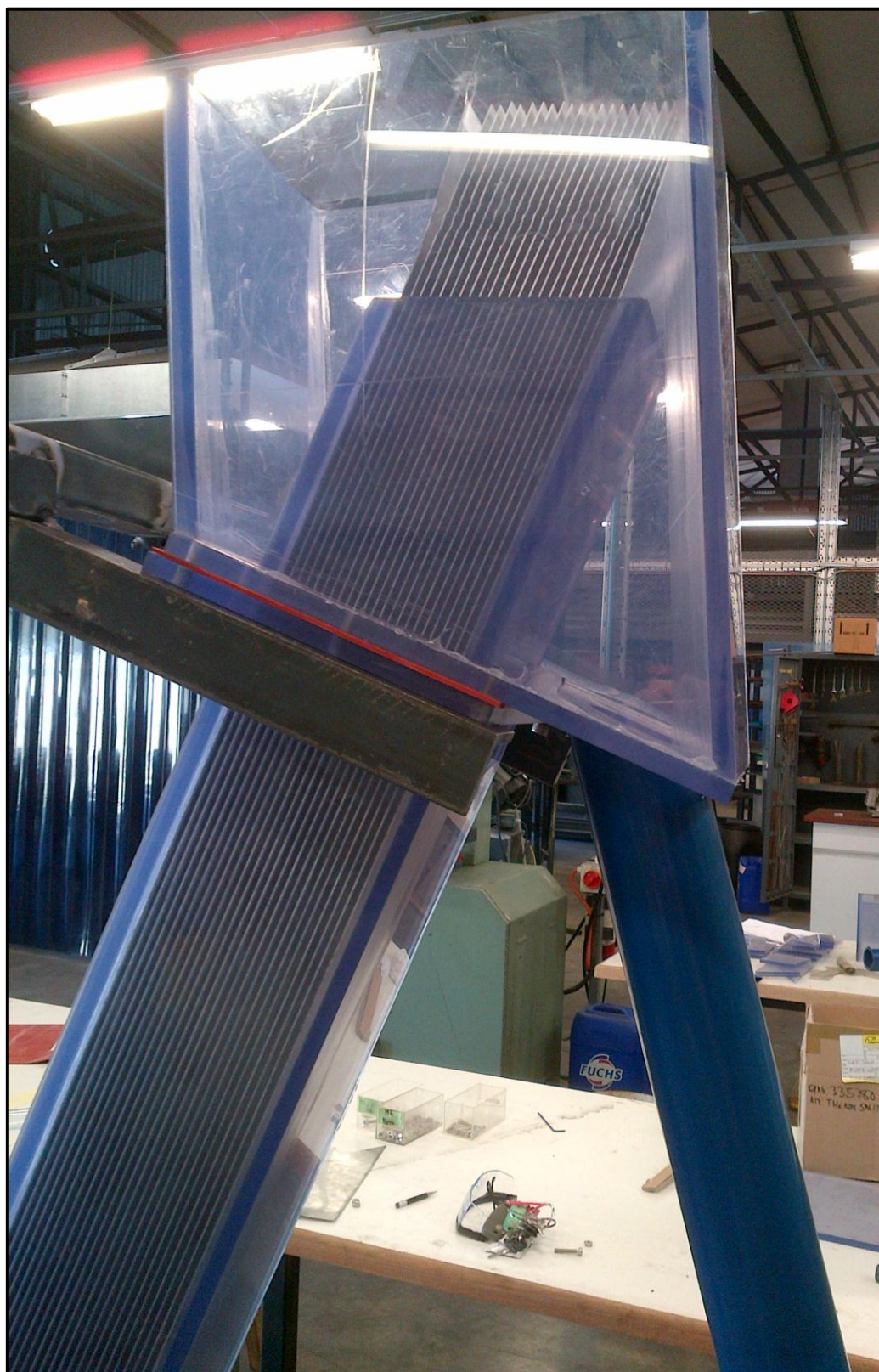


Figure C-7: Photo – Overflow compartment with plates protruding the plates section



Figure C-8: Photo – Gravity feeding of unit during semi-batch mode



Figure C-9: Photo – Initial stage of valve and pipe configuration of semi-batch feeding system



Figure C-10: Photo – Coal slurry tank with agitator motor and shaft

APPENDIX D - Partition Curves

D.1 Tromp curves for diluted tracers only system

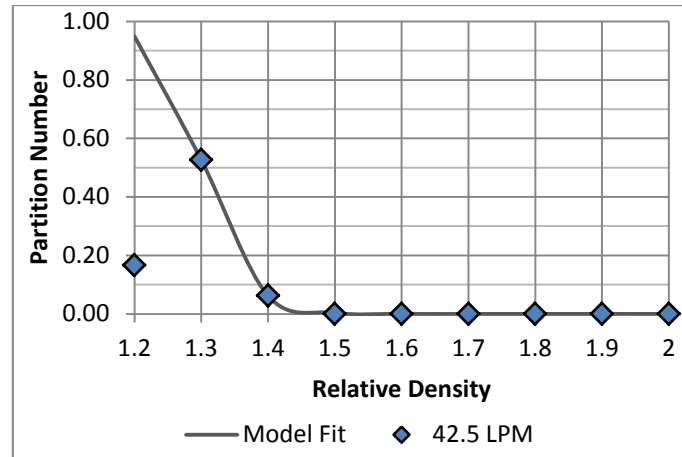


Figure D-1: Partition curve for water only tracer run at 42.5 LPM (Fractionation)

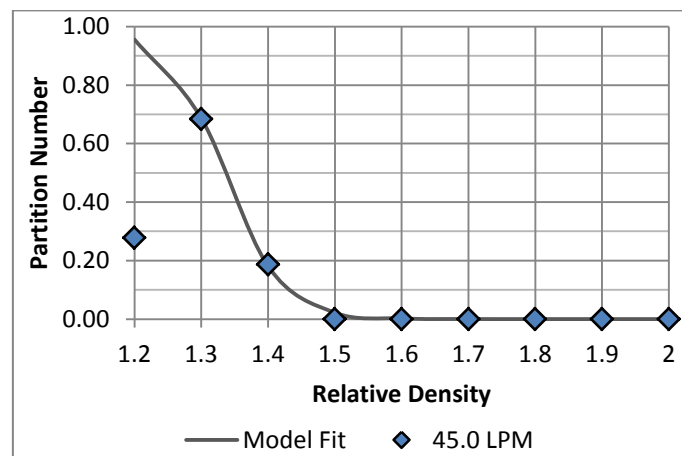


Figure D-2: Partition curve for water only tracer run at 45.0 LPM (Fractionation)

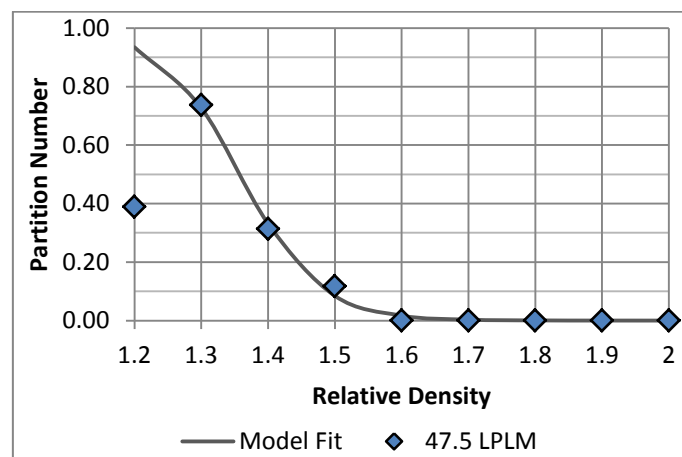


Figure D-3: Partition curve for water only tracer run at 47.5 LPM (Fractionation)

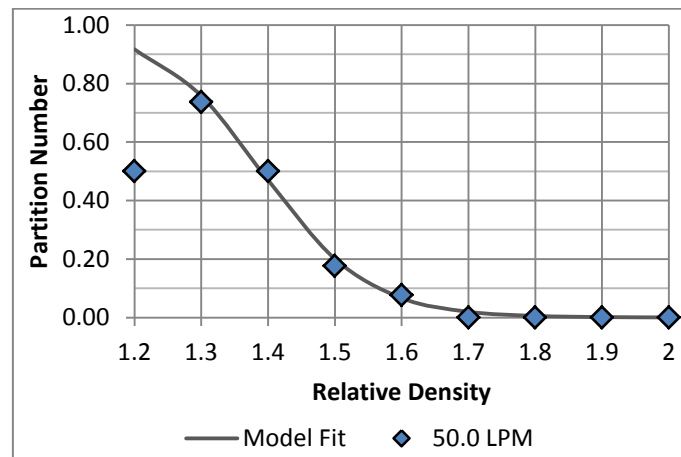


Figure D-4: Partition curve for water only tracer run at 50.0 LPM (Fractionation)

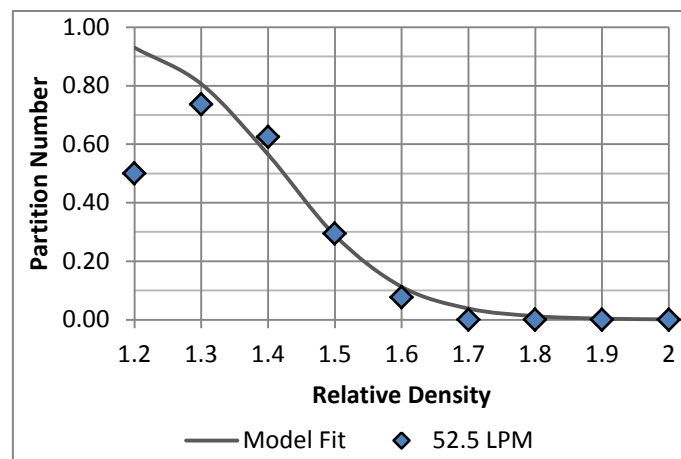


Figure D-5: Partition curve for water only tracer run at 52.5 LPM (Fractionation)

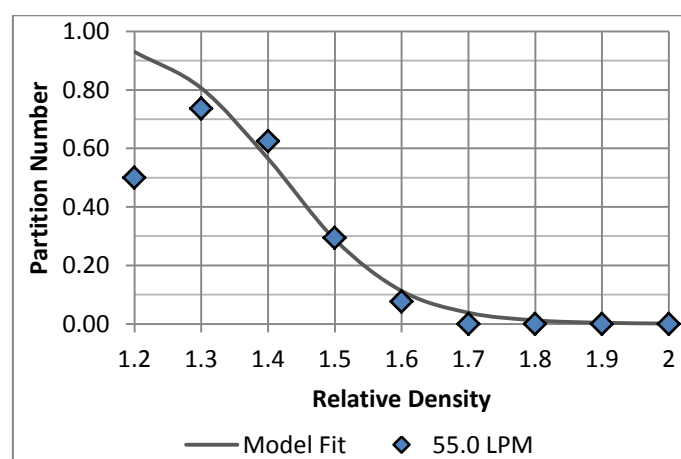


Figure D-6: Partition curve for water only tracer run at 55.0 LPM (Fractionation)

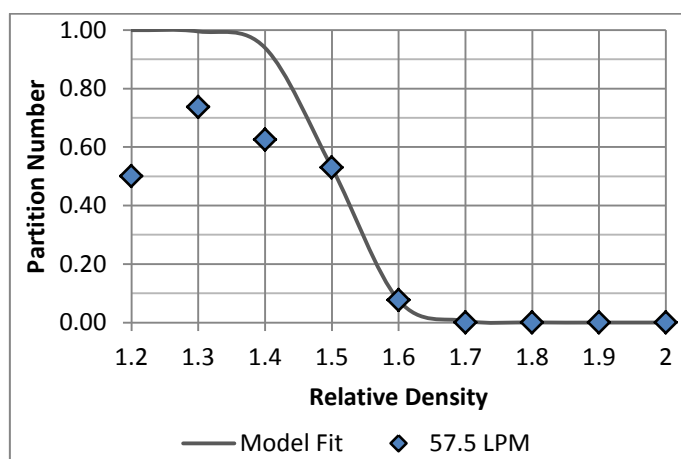


Figure D-7: Partition curve for water only tracer run at 57.5 LPM (Fractionation)

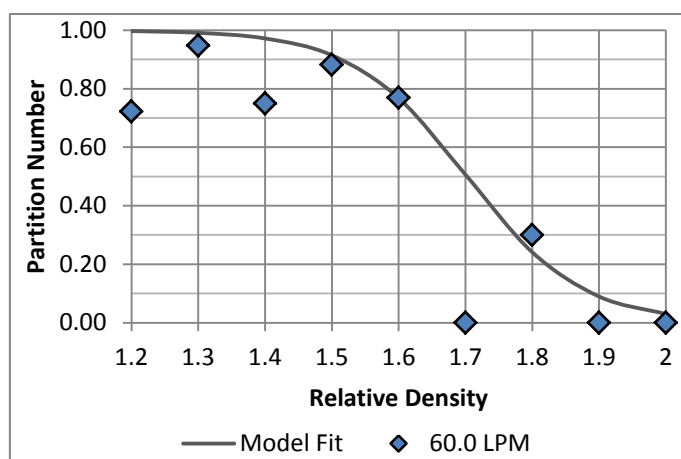


Figure D-8: Partition curve for water only tracer run at 60.0 LPM (Fractionation)

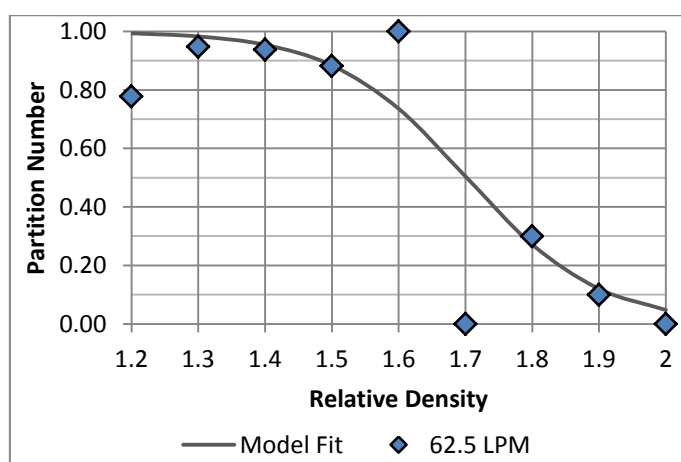


Figure D-9: Partition curve for water only tracer run at 62.5 LPM (Fractionation)

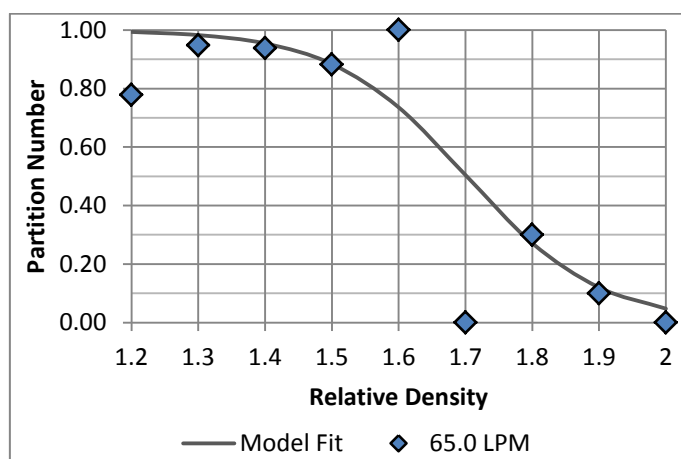


Figure D-10: Partition curve for water only tracer run at 65.0 LPM (Fractionation)

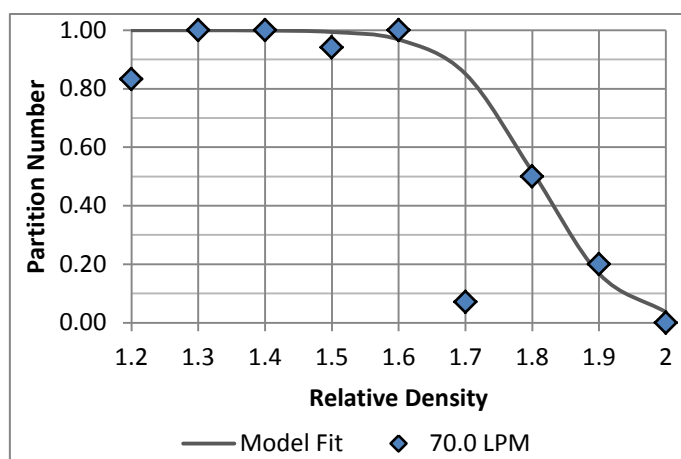


Figure D-11: Partition curve for water only tracer run at 70.0 LPM (Fractionation)

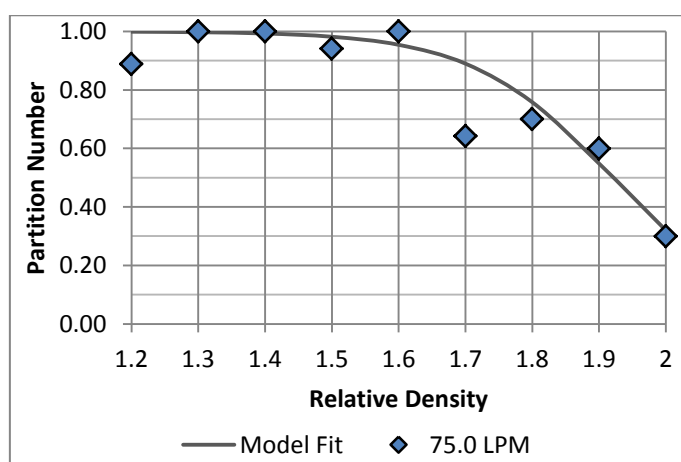


Figure D-12: Partition curve for water only tracer run at 75.0 LPM (Fractionation)

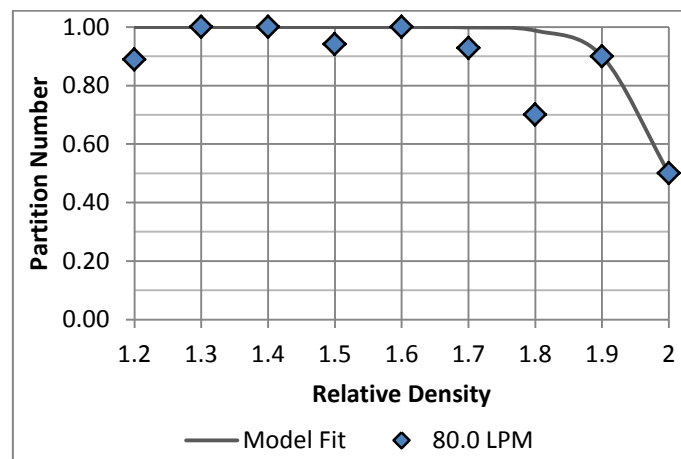


Figure D-13: Partition curve for water only tracer run at 80.0 LPM (Fractionation)

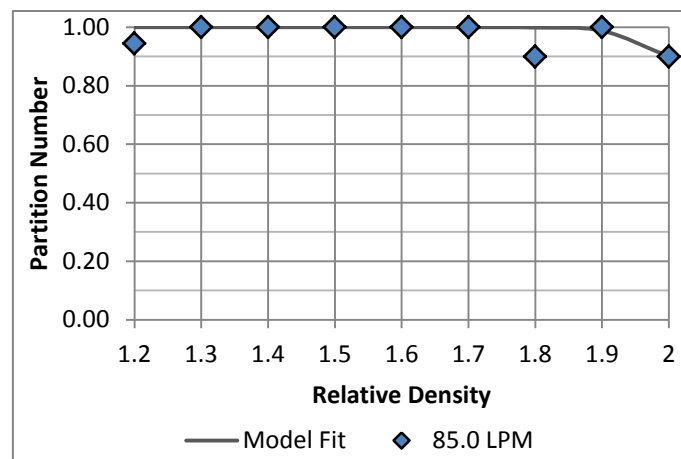


Figure D-14: Partition curve for water only tracer run at 85.0 LPM (Fractionation)

D.2 Tromp curves for coal A and tracer system

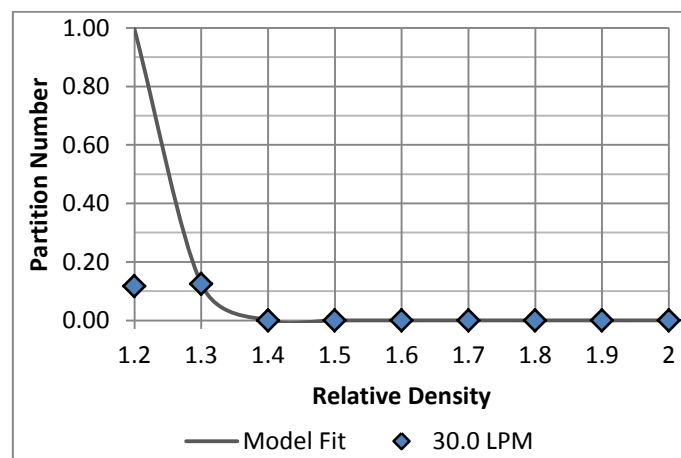


Figure D-15: Partition curve for coal and tracer run at 30.0 LPM (Fractionation)

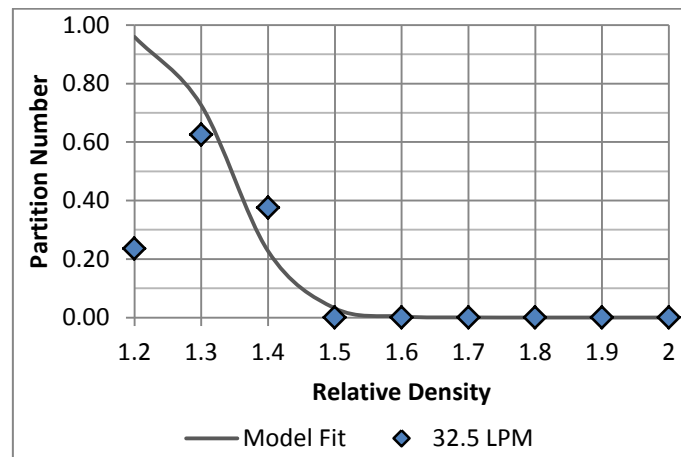


Figure D-16: Partition curve for coal and tracer run at 32.5 LPM (Fractionation)

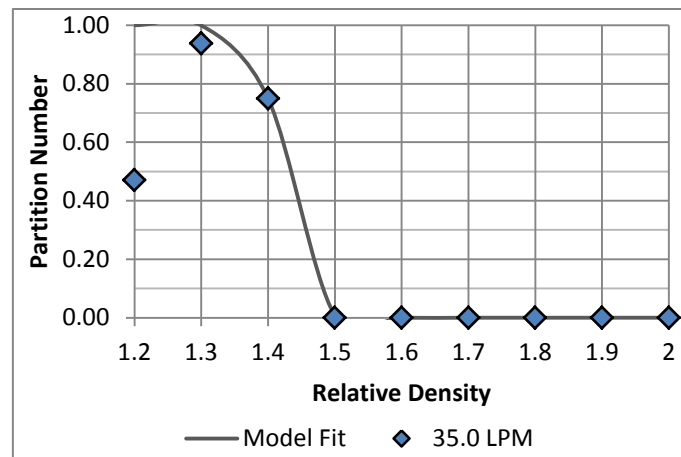


Figure D-17: Partition curve for coal and tracer run at 35.0 LPM (Fractionation)

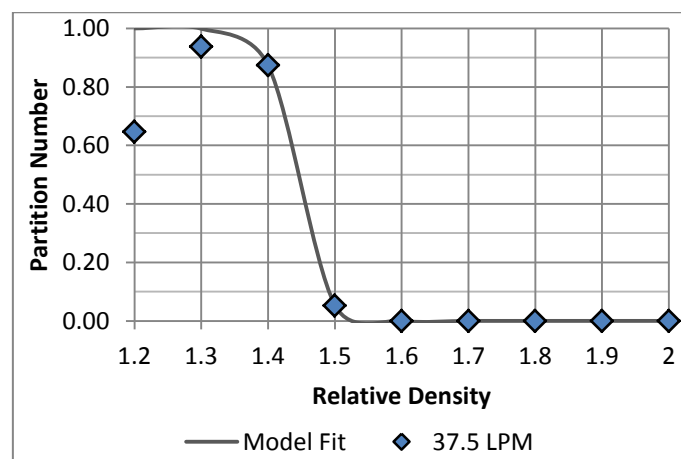


Figure D-18: Partition curve for coal and tracer run at 37.5 LPM (Fractionation)

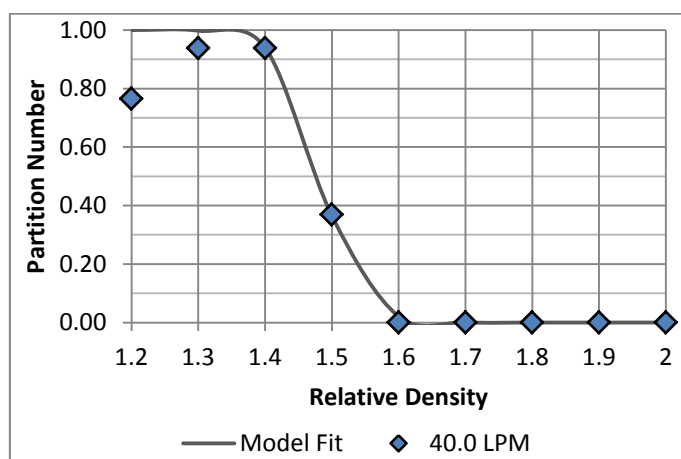


Figure D-19: Partition curve for coal and tracer run at 40.0 LPM (Fractionation)

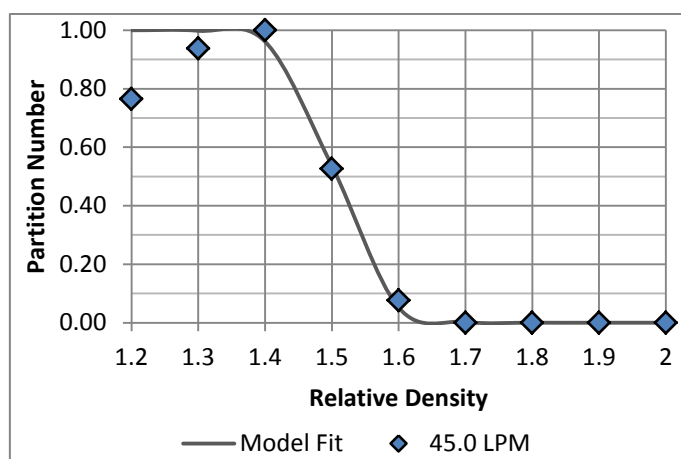


Figure D-20: Partition curve for coal and tracer run at 45.0 LPM (Fractionation)

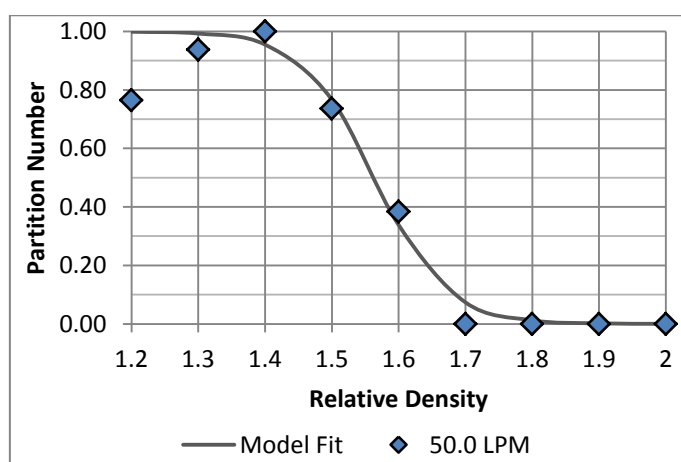


Figure D-21: Partition curve for coal and tracer run at 50.0 LPM (Fractionation)

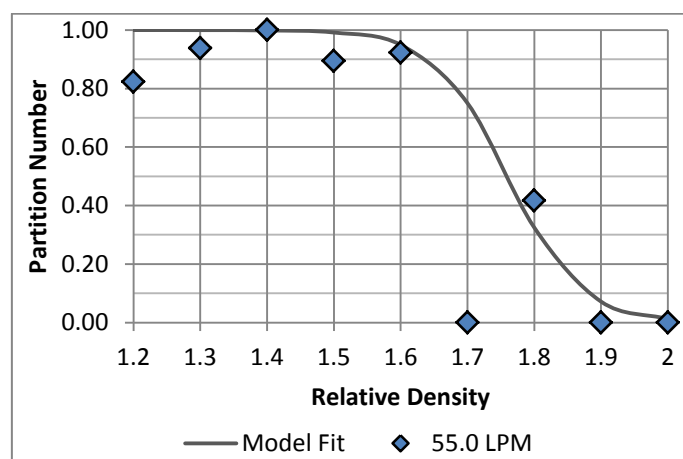


Figure D-22: Partition curve for coal and tracer run at 55.0 LPM (Fractionation)

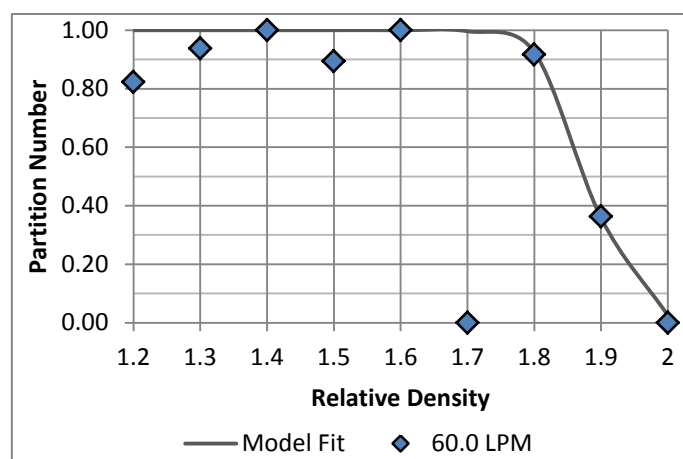


Figure D-23: Partition curve for coal and tracer run at 60.0 LPM (Fractionation)