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Investigation of the shifting-parameter as a function of density in the fluidization of a packed bed.

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Abstract

Accurate predictions of pressure drops in fluidized beds are of great importance in the industry. Up to date no satisfactory correlation exists to predict the pressure drop in a fluidized bed as the bed is traversing from one regime to another.
In the present study experiments have been performed in an experimental fluidized bed reactor. The experimental tower has been equipped with a set of nine pressure sensors located at different positions along the height of the tower. The tower has a diameter of 7.2 *cm* and a height of 1.5 *m*.
A procedure providing a correlation for data in the transition region between asymptotic solutions or limiting correlations have been described by Churchill & Usagi (1947). This correlation can generally be expressed as $y^s\{x\} = y_o^s\{x\} + y_\infty^s\{x\}$, where $y_o\{x\}$ and $y_\infty\{x\}$ represents the asymptotic solutions for large and small values of the independent variable x and s is the shifting parameter. Changing the value of s shifts the correlation given by $y\{x\}$ closer to or away from the asymptotic solutions. This procedure has been proven to give good correlations in a wide range of applications.
A series of different powders have been used to investigate the influence of a particular parameters on the shifting parameter, s. Up to date no expression has been stated for this shifting parameter to govern the transition from fixed to fluidized bed. Two powders have been used in the present study and they are Zirconium Oxide (*ZrO*) and spherical glass particles. The powders have the same size distributions but very different densities. The effect of different densities on the shifting parameter was investigated. Several different drag models were used to serve as a control for investigating the shifting parameter. The results are given in the form of pressure drop data versus superficial velocity data. Experimental data are presented with the drag model correlations and the investigated values of the shifting parameter, s. Some of the drag models that were used were the Syamlal O' Brien drag model (Syamlal, Rogers & O'Brien (1993)) and the extended Hill-Koch-Ladd drag correlation (Benyahia, Syamlal & O'Brien (2006)). The results are evaluated and discussed.

Nomenclature

Roman symbols

$d_{p\ mean}$	surface-volume mean diameter (m)
g	gravitational constant ($\mathrm{ms^{-2}}$)
L	bed height (m)
p	pressure ($\mathrm{Nm^{-2}}$)
q	superficial velocity ($\mathrm{ms^{-1}}$)
Re	Reynolds number (-)
y	canonical dependent variable ($\mathrm{Nm^{-3}}$)

Greek symbols

ϵ	porosity (-)
ρ	density ($\mathrm{kg\ m^{-3}}$)

Subscripts

c	critical point
f	fluid
mf	point of minimum fluidization
p	particle
o	limiting condition for small values of the independent variable
∞	limiting condition for large values of the independent variable

Superscipts

s	shifting parameter

Introduction

Fluidized bed reactors are widely employed in the chemical, petrochemical, metallurgical and pharmaceutical industries (Stein, Ding, Seville & Parker (2000)). Better understanding of the complex multiphase fluid and solid movement are essential for optimal reactor design. The powered addition technique serves as a method to correlated data in the transition region between two limiting conditions. This technique has got the possibility of a wide range of applications as described in the work by Churchill & Usagi (1947). Applying the technique to a fluidized bed traversing from a fixed to fluidize bed proved very useful. In general the shifting parameter, *s*, can be defined as follows

$$y^s\{q\} = y_o^s\{q\} + y_\infty^s\{q\}, \tag{1}$$

where $y_o\{q\}$ and $y_\infty\{q\}$ represents the asymptotic solutions for large and small values of the superficial velocity, q, and *s* is the shifting parameter. In fluidized beds the lower bound, $y_o\{q\}$, is described by drag models. Over the years numerous drag models have been proposed. In general two types of experimental data can be used to create a fluid-solid drag model (Syamlal, Rogers & O'Brien (1993)). The first type is with packed-bed pressure drop data expressed in the form of a correlation and the second is provided in the form of correlations for the terminal velocity in a fluidized or settling bed, expressed as a function of porosity and Reynolds number (Syamlal, Rogers & O'Brien (1993)). A well known example of a drag model based on the packed bed pressure drop data is the Ergun equation (Kunii & Levenspiel (1991)) and an example of a drag model using the terminal velocity correlation is the Syamlal O'Brien drag model (Syamlal, Rogers & O'Brien (1993)). It will be discussed later in the present study how these to basic formulations of drag models may influence the pressure drop predicted by these models in the fixed bed regime of a fluidized-bed reactor.

At minimum fluidization the total weight of the packed bed is supported by the upward force created by the gas moving upward through the porous structure. As the superficial velocity is increased above minimum fluidization velocity, the pressure drop remains practically the same (Kunii & Levenspiel (1991)). In some cases the pressure drop does not remain constant in the fully fluidized regime but actually increases. The explanation for the slight increase of pressure drop with an increase of superficial velocity may be attributed to wall effects which occurred due to the physical dimensions of the experimental tower used. More specifically, some slugging can occur and due to the formation of slugs additional potential energy is required to move the slug vertically. The result is an approximate linear increase in pressure drop across the fluidized bed (Chen, Gibilaro & Foscolo (1997)). In the present study the pressure drop in the fluidized regime will be assumed constant. This constant pressure drop will be assumed as the upper bound, $y_\infty\{q\}$, in our powered-addition correlation given in equation (1). At minimum fluidization velocity the pressure-drop is given by

$$\Delta p = (1-\epsilon)(\rho_p - \rho_f)Lg, \tag{2}$$

with ρ_p the particle's density, ρ_f the fluid density and L the bed height.

In previous studies it has been found that the shifting parameter *s* was relatively independent of the particle sizes that were investigated (Rautenbach, Melaaen & Halvorsen (2010)). The three particle size distributions that were used were 100-200 μm, 400-600 μm and 750-1000 μm. It was found that a shifting parameter value greater than about 12 but smaller than about 20 produced an acceptable correlation in the transition region between the fixed to fluidized regime.

Experimental set-up

The experiments that were carried out in the present study were performed at the TUC (Telemark University College) in Porssgrun Norway. A 1.5 *m* long experimental fluidized bed reactor were used. The pressure drop data were acquired using a set of nine pressure probes located at different height along the bed. This set-up is presented in Figure 1. A porous plate distributor was used in this study to produce an uniform entry profile to the bed.

In the present study the influence of the particle density on this shifting parameter, *s*, was investigated. The two powders used were Zirconium Oxide (*ZrO*) and glass particles. Both were spherical particles with a size distribution of 400-600 μm. The particle size distributions are given in Figure 2. The void fraction of the beds as well as the minimum fluidization velocities and the particle densities are presented in Table 1. The void fractions refer to the void fractions after the bed has been fluidized. This void fraction will be used in calculating the predictions of the drag models in the fixed bed regime. Two mixtures of the *ZrO*-powder and glass powder were used to create a powder mixture with a different effective density than the two original powders. One mixtures consisted of one third *ZrO*-powder and two thirds glass powder. The other mixture was half-half *ZrO*-powder and glass-powder. The mean particle diameters presented in Table 1 are the *surface-volume mean* diameter and can be expressed as

$$d_{p\ mean} = \frac{1}{\sum_i x_i/d_i}, \tag{3}$$

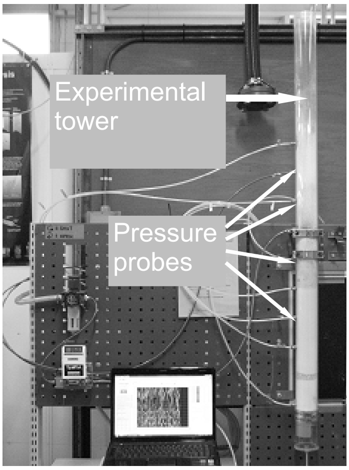


Figure 1: Experimental fluidization reactor equipped with nine pressure probes and fed through a porous plate distributer.

where x_i is the mass fraction of the particles with a diameter d_i. An estimation of the minimum fluidization velocity, U_{mf}, was determined by equating equation (2) with the pressure drop prediction of the Ergun equation (Kunii & Levenspiel (1991)). The intersection between these two predictions gives a fairly good correlation for U_{mf}. For the *ZrO*-powder the surface-volume mean diameter were used and for the glass particles the larges particle size (600 μm) were used as this gives the best correlation to experimental data using this particular intersection method. For both the *ZrO*-powder and the glass powder only the first term of the Ergun equation were used as this produced a good estimation of the minimum fluidization velocity. This approximation with the Ergun equation is usually only used with low superficial velocities or very small particles ($Re_{p,mf} < 20$) (Kunii & Levenspiel (1991)). This one term approximation of the Ergun equation did not give good enough correlations for the two mixture powders' minimum fluidization velocities and thus the total Ergun equation were used. The results are given in Table 1. For the mixture consisting of one third *ZrO*-powder a porosity of 0.38 were used and not the measured porosity of 0.42. This value is just the average value of the porosities of the original powders. This value of 0.38 was assumed to compensate for the suspected incapability of the drag models to compensate for the drag effect associated with segregation. This topic will be discussed later in the present work. In both of the mixtures' calculations of the minimum fluidization velocities an effective particle

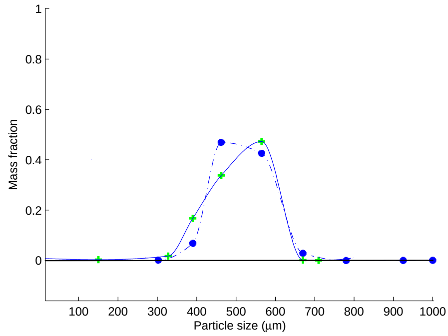


Figure 2: Particle size distributions for the *ZrO*-powder (-··-) and the glass powder (—).

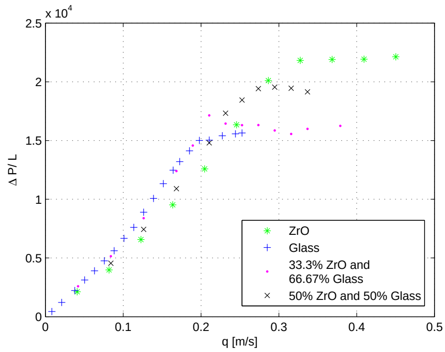


Figure 3: Pressure-gradient against superficial velocity data for all the powders investigated.

size of 600 μm were assumed as this produced an adequate result.

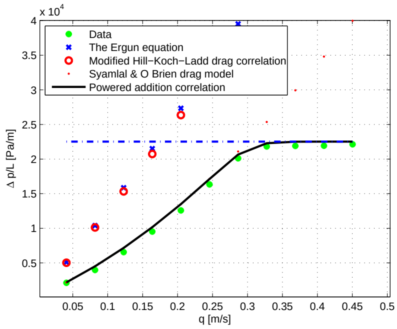
Results

The pressure-drop data retrieved for the powders investigated are given in Figure 3. It is very clear to see that the higher density Zirconium Oxide (*ZrO*) produces a much larger pressure drop across the bed and also fluidizes at a higher value for the superficial velocity, *q*.

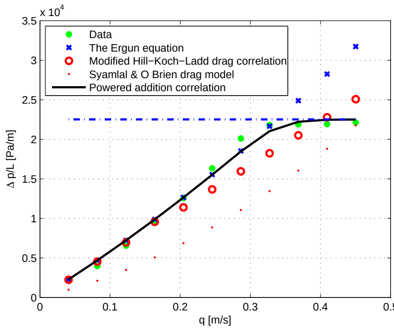
In the present study several drag model have been used to provide a pressure drop correlation in the fixed bed regime. Some of these models differ in the way that they have been derived but most of them make use of some sort of empirical basis. The models used in the present study was the well know Ergun equation (Kunii & Levenspiel (1991)), the Syamlal O' Brien drag model (Syamlal, Rogers & O'Brien (1993)) and the extended

Table 1: Summary of particle- and bed properties.

Variable	Value	Units
Zirconium Oxide		
ϵ	0.39	-
ρ_p	3800.0	kg/m^3
U_{mf}	0.3	m/s
$d_{p\ mean}$	503.29	μm
Glass		
ϵ	0.37	-
ρ_p	2500.0	kg/m^3
U_{mf}	0.21	m/s
$d_{p\ mean}$	482.93	μm
33.3% Zirconium Oxide and 66.7% glass		
ϵ	0.42	-
ρ_p	2933.33	kg/m^3
U_{mf}	0.24	m/s
$d_{p\ mean}$	489.53	μm
50% Zirconium Oxide and 50% glass		
ϵ	0.38	-
ρ_p	3150.0	kg/m^3
U_{mf}	0.25	m/s
$d_{p\ mean}$	492.9	μm



(a)



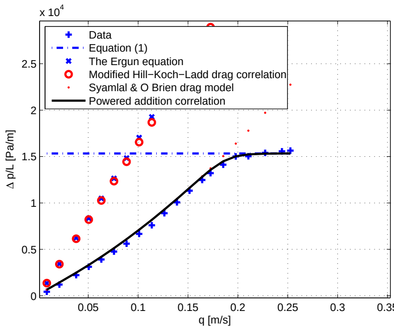
(b)

Figure 4: Investigation of effective particle diameter on the different drag-model predictions with the *ZrO*-powder. Drag models with an effective particle diameter of (a) 400 μm and (b) 600 μm respectively.

Hill-Koch-Ladd drag correlation (Benyahia, Syamlal & O'Brien (2006)). In all of the Figures the limiting condition for large values of q are represented by the dash-dot line (equation (2) is represented by (-·-)). These different drag models are represented in Figure 4 and 5 along side the pressure drop data acquired by using the *ZrO*-powder and glass-powder. For conciseness the models predictions are shown using the largest and smallest particle size of the distributions used. The mean value of the particle distributions were not used because it did not give good correlations in any of the cases investigated in the present study.

It is clear that the different models have varying accuracy with different values of the effective particle diameter. Previous research has found that in a fluidized bed, consisting of a particle size distribution, it is the smaller particle sizes that have the largest contribution (Jayarathna & Halvorsen (2009)). This followed from data that were collected by using different mixtures of particles. In the work by Jayarathna & Halvorsen (2009) it was found that only after about 40% of the mixture consisted of the larger particles did the minimum fluidization velocity differ considerably from the value for U_{mf} found with just the smaller particles. Even after 40% of the bed consisted of large particles the minimum fluidization velocity was closer to the smaller particles minimum fluidization velocity than to the value for the larger particles (Jayarathna & Halvorsen (2009)). Given the data as represented in Figure 4 it seems that the Syamlal O' Brien drag model (Syamlal, Rogers & O'Brien (1993)) produces the best correlation at a particle diameter close to the smallest particle diameter in the range. This is in agreement with previous research (Jayarathna & Halvorsen (2009)) and from Figure 2 it is clear that there were some particles with a diameter even smaller than 400 μm. This fact makes the good correlation found with the Syamlal O' Brien drag model (Syamlal, Rogers & O'Brien (1993)) even more feasible as 400 μm is not necessarily the smallest particle size in the distribution.

In all the cases investigated the Ergun equation (Kunii & Levenspiel (1991)) and the extended Hill-Koch-Ladd



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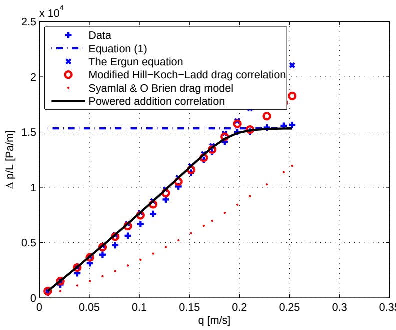
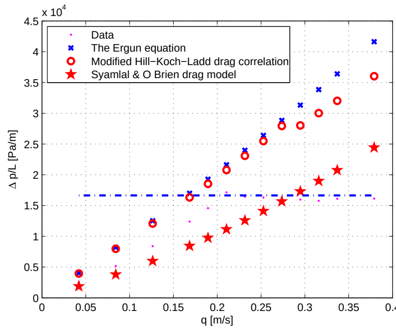
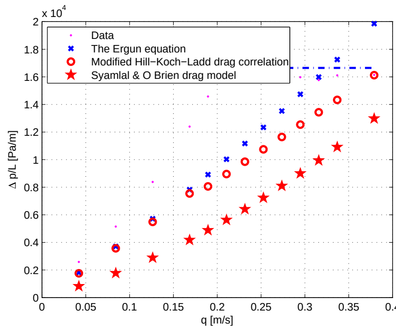


(b)

Figure 5: Investigation of effective particle diameter on the different drag-model predictions with just the glass powder. Drag models with an effective particle diameter of (a) 400 μm and (b) 600 μm respectively.



(a)



(b)

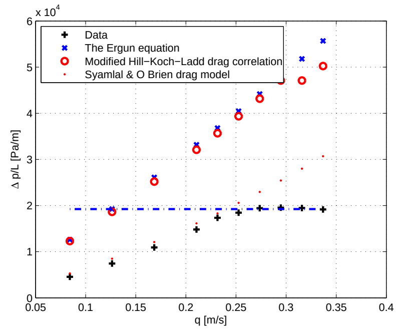
Figure 6: A mixture powder consisting of one third *ZrO*-powder and two thirds glass powder. Drag model comparisons to data assuming a particle size of (a) 400 μm and (b) 600 μm.

drag correlation (Benyahia, Syamlal & O'Brien (2006)) corresponded to the larger particle size in the distribution. This result agrees with previous findings by de Wet, Halvorsen & du Plessis (2009). The results for the glass particles and mixtures are similar to that given in Figure 4 for the *ZrO* and are given in Figure 5 to 7.

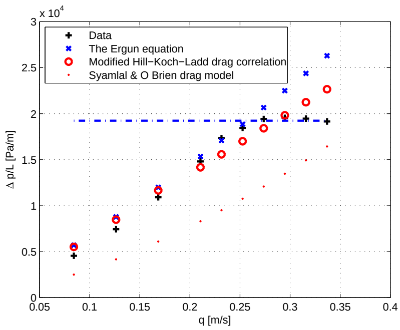
Results using powered addition. If we assume that the fully fluidized regime can be given by equation (2) then the following equation is produce using the powered addition technique

$$\frac{\Delta p}{L} = \left(Drag\ model^{-s} + equation\ (2)^{-s}\right)^{-\dfrac{1}{s}}, \quad (4)$$

were any adequate drag model can be used. The negative powers of s is because the data is a decreasing power of q. The powered additioned results are depicted in Figure 4, 5 and 8. First the correlation is given when the Syamlal O' Brien drag model (Syamlal, Rogers & O'Brien (1993)) is used and a particle diameter equal to 400 μm (Figure 4 and 5 (a)). Secondly the powered addition correlations are given using the Ergun equation and a particle diameter of 600 μm (Figure 4 and 5 (b)). In all the cases a value of 15 was used as the shifting parameter although values with in a range from 12 to about 20 would have serviced. These results seem to indicate that the shifting parameter s is insensitive to the density of the particles. In an attempt to confirm this suspicion the same correlations were made but now with the powder mixtures. In Figure 6 and 7 the correlation with the data are given with the larges and smallest particle sizes, namely 400 μm and 600 μm respectively. In both cases poor agreement were found between the drag models and the experimental data. In the case of the mixture



(a)



(b)

Figure 7: A mixture powder consisting of fifty percent *ZrO*-powder and fifty percent glass powder. Drag model comparisons to data assuming a particle size of (a) 400 μm and (b) 600 μm.

consisting of one third *ZrO* particles a relatively high void fraction value is obtained as presented in Table 1. This is considered a high value because the powders that this mixture is made up of has void fractions below 0.4. The main suspicion for this high value is due to segregation. The heaver *ZrO* particles can move to the bottom of the tower while the lighter glass particles can move to the top of the bed. As all of the drag models investigated has got some sort of assumption of only one particle size it is not a bad assumption that segregation could be the cause of this discrepancy. Although most of the investigated models should cover all porosities they still did not take into account the effect of particle size distributions and segregation. When a void fraction of about 0.38 was used (average void fraction of the base powders) for the powder consisting of one third *ZrO*, very good correlations were found with the data as presented in Figure 8. Much the same result as depicted in Figures 4 and 5. Thus the drag models predicted a more accurate drag with a lower void fraction supporting the suspicion that the drag models do not take into account the influence of segregation. The possible reason why the one powder mixture formed some segregated effects and the other powder mixture (50/50 mixture of glass and *ZrO*) did not is explainable in light of initial fluidization. If the superficial gas velocity, q, is slowly decreased after fluidization the heavy particles have time to settle to the bottom while the lighter particles are force upwards. Precaution was taken to avoid this, for example to close the gas inlet quickly after fluidization, but this approach is not guaranteed to always work equal effectively.

The aim of the present study is to investigate the shifting parameter, s. To be able to do so a fairly accurate prediction of the pressure drop in the fixed bed regime is required. Thus a porosity of 0.38 will be assumed for the mixture powder having a volume that consist of one third *ZrO*. In Figure 8 the correlation using the powered addition technique is given with the mixture powders' pressure drop data. In both cases the Syamlal O' Brien drag model (Syamlal, Rogers & O'Brien (1993)) were used with a particle diameter of 400 μm. In both cases a value of 15 were used for the shifting parameter, s. A possible explanation for the discrepancy of the Syamlal O' Brien drag model (Syamlal, Rogers & O'Brien (1993)) in the fixed bed regime of Figure 8 (b) be can be because the smallest particle size in the distribution range were used in the drag correlations and this particle size is just an assumption. The characteristics of a particle bed seems to be mainly determined by the smaller particles (Jayarathna & Halvorsen (2009)) but of course the larger particles will also still have an effect on the over all pressure drop. It is also possible that the mixture's particle size distribution did not consist out of an approximate bell-shaped curve but that there were more of the bigger particles than in the other powders that were investigated. A sieve analysis has to be performed to confirm or disconfirm this hypothesis.

Discussion

Effective particle size is of great importance when working with a particle size distribution. In a lot of practical applications fluidized beds consist of such powders. It is clear from the results produced in the present study and from results in previous work (Jayarathna & Halvorsen (2009)) that the smaller particle sizes in the distribution plays a bigger role in the estimation of the drag. More research is needed to find a better way of estimating the representative particle diameter for particle size distributions. I seems clear from the present study and from previous research done by de Wet, Halvorsen & du

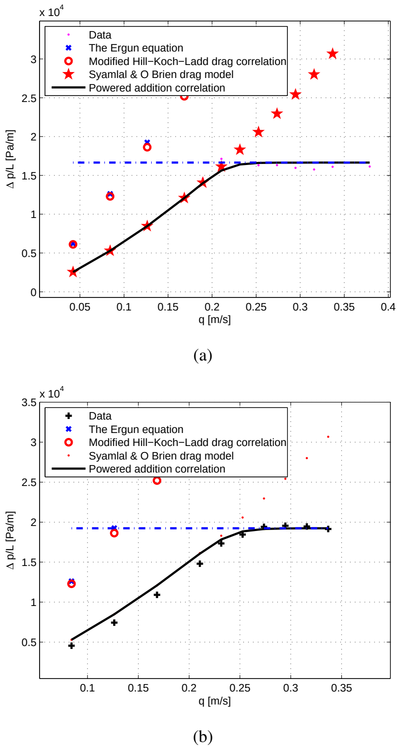


Figure 8: Powered addition correlation using the Syamlal O' Brien drag model (Syamlal, Rogers & O'Brien (1993)) and a particle size of 400 μm compared to (a) the mixture powder consisting of one third *ZrO* and two third glass and (b) the mixture powder consisting of fifty percent *ZrO* and fifty percent glass.

Plessis (2009) that a normal average representative value for the particle diameter is not appropriate. Using a particle size equal to 400 μm together with the Syamlal O' Brien drag model (Syamlal, Rogers & O'Brien (1993)) did however produce acceptable results.

In a case where data in the transition region is know a different approach can be taken to calculate the shifting parameter. At the point where the two asymptotes intersect a critical value is obtained (de Wet, Halvorsen & du Plessis (2009)). In the present study this point forms an estimation for the minimum fluidization velocity with the lower bound being a drag model and the upper bound equation (2). This can also be expressed as

$$y_o\{q\} = y_\infty\{q\}. \tag{5}$$

At this point equation (1) simply becomes

$$y_c^s = y_o^s + y_\infty^s = 2y_o^s = 2y_\infty^s, \tag{6}$$

with y_c the functional value at the critical point (de Wet, Halvorsen & du Plessis (2009)). Solving for *s* produces the following equation

$$s = \frac{ln\ 2}{ln\ y_c - ln\ y_\infty} = \frac{ln\ 2}{ln\ y_c - ln\ y_o}. \tag{7}$$

Thus if the functional value y_c is known a suitable value for the shifting parameter, *s*, can be calculated.

The problem with this procedure is that in the industry one usually wants to determine this transition regime not prescribe it. As the physical meaning of the shifting parameter is not known further research is needed to describe the value for *s* more precisely.

A possible explanation for the high pressures predicted by the Ergun equation could be found in the manner in which it was derived. It was derived on a fixed bed model and then later adapted empirically using fixed bed pressure drop data.

The inaccuracies of the Hill-Koch-Ladd drag correlation (Benyahia, Syamlal & O'Brien (2006)) could probable be based on the empirical way in which it was derived. Singling out the exact cause for the over estimation of the drag is not a trivial task.

It must be noted though that equation (2) did produce a reasonable result with a particulated bed void fraction equal to 0.42 in the case of the mixture consisting of a third *ZrO*. This actually leads to the same suspicion that the problem is mainly with the drag models in the case of a segregated bed.

Conclusions

In fitting an appropriate curve the shifting parameter is relatively insensitive as found by previous research (de Wet, Halvorsen & du Plessis (2009)). A range of values produce an appropriate result. This range can be anywhere between 12 and 20 but 15 was chosen in the present study. The shifting parameter, *s*, seems to be insensitive to changes in density. With all four powders investigated a value of 15 serviced. Further research is needed to determine the physical meaning of *s*, but it seems that a value of 15 is a good estimation for the shifting parameter in most practical application.

Thus an effective correlation is produced to give an adequate prediction of pressure drop data for a fluidized beds traversing from fixed to fluidized regime

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