NATIONAL INSTITUTE OF TECHNOLOGY
ROURKELA

"DRY BENEFICIATION OF HIGH ASH NON-COKING COAL USING AN AIR DENSE MEDIUM FLUIDIZED BED"

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In partial fulfillment of the requirements of
Bachelor of Technology (Chemical Engineering)

SUBMITTED BY

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Under the Guidance of

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CERTIFICATE

This is to certify that the thesis entitled, "**DRY BENEFICIATION OF HIGH ASH NON-COKING COAL USING AN AIR DENSE MEDIUM FLUIDIZED BED**" submitted by Sri Asish Kumar Sahoo in partial fulfillments of the requirements for the award of Bachelor of Technology Degree in Chemical Engineering at National Institute of Technology, Rourkela (Deemed University) is an authentic work carried out by him under my supervision and guidance.

To the best of my knowledge, the matter embodied in the thesis has not been submitted to any other University / Institute for the award of any Degree or Diploma.

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ABSTRACT

In this project, dry beneficiation of high ash non-coking coal has been conducted by air dense medium fluidized bed. For dry separation, creating and sustaining the air dense medium is a complex process which requires intensive investigation. The dynamic stability of the bed which plays an important role in the sharpness of the separation has also been studied. Based on experimental data, four dimensionless groups i.e. Froude number, Reynolds number, ratio of density of fluid and solid and aspect ratio of the system are used to characterize the stability and quality of fluidization. The above stabilized bed is used to beneficiate coarse coal of -10 +0.1 mm size. The quality of separation is judged by ash analysis of the beneficiated coal samples collected from specified heights of the bed. Enrichment is represented as a function of different operating parameters represented by four dimensionless groups obtained from dimensional analysis approach. The values of enrichment calculated with the developed correlation have been tested which agrees fairly well with the experimental values of enrichment.

Keywords: Air dense medium fluidized bed, dry beneficiation of coal, Dynamic stability of bed, Coal enrichment.

1. SIGNIFICANCE OF THE PROJECT

Beneficiation of non-coking coals in India was not given due importance till last decade due to its low value or not being able to meet the cost of the process**. About 73% of the non-coking coal is used by the thermal power plants containing high ash ranging from 40-50%.** Recently, the Ministry of Environment and forest, govt. of India has made a regulation to use the coal having ash less than 34% for the thermal plants situated 1000 kms away from pit head or close to the urban/sensitive/critical area. These categories of power plants in India are around 60% of total power plants. In addition there is high demand for low ash content coal for sponge iron and blast furnace process. In view of this it is required to reduce the ash content in present Indian non-coking coals by any suitable beneficiation techniques so that the demand from steel plants (blast furnace operation) , sponge iron and power plants can be met.

Table-1 COAL RESERVES IN INDIA (as on 1.1.2007) in billion tonnes (data from Coal India Ltd)

	TOTAL RESERVE	PROVED RESERVE	INDICATED RESERVE	INFERRED RESERVE
COKING	32	17	13	2
NON-COKING	255	98	119	36
TOTAL	287	115	132	40

Due to the huge reserves of non-coking coal in India, in near future the dry beneficiation technology associated with such non-coking coal is gradually going to gain importance.

1.1 WHY DRY BENEFICIATION

Dry coal beneficiation has constantly lost its importance vis-à-vis hydraulic beneficiation during the last five decades, primarily because of the sharper separations achievable with the modern hydraulic beneficiation technologies, such as gravity separation and flotation for reduction of ash materials. Nevertheless, there are certain ***inherent advantages of dry separation methods*** which would give them advantages in the competitive market place to increase their thermal efficiency vis-à-vis hydraulic processes. These advantages are

- A dry product, resulting in a higher calorific value per ton.
- The water source problem is acute in semi-arid areas. Hydraulic processing of coal requires large quantity of water .Water is consumed as product moisture, tailings disposal and evaporation.
- Waste generated from hydraulic process after maximizing the recycle the water is unsuitable for disposal to water resources, because it contains good amount of waste solids fines, which causes the pollution of water bodies. Dry processes avoid the problems associated with treatment and storage of process waste water.
- The fines generated in dry processing are suitable for ideal fuel for fluidized bed combustor.

Water demand and pollution have opened up scope of research in the development of the process for dry cleaning of coal. Significant developmental work in dry beneficiation of coal is in progress in **Canada, China and India.**

There are different types of dry cleaning processes for coal beneficiation. Hand picking of gangue minerals or shale in coarse size is one of the simplest, oldest and labour intensive techniques of dry cleaning processes. The other dry cleaning techniques are mechanical methods (screening, classifier, gravity concentrations, heavy media separation etc.), magnetic separation, electrostatic separation, etc. These processes depend on the differences in physical properties between coal and gangue minerals such as density, size, shape, lusterness, magnetic conductivity, electric conductivity, radioactivity etc. These methods have both advantages and disadvantages. Air dense medium fluidized bed separation is one of the dry beneficiation processes that would offer benefits compared to other dry beneficiation processes .The results of economic evaluation for different processes is given in Table-2. The factor used to assess the main processes under consideration is the cost per heat unit delivered to the power station. This factor takes into account the benefit of reduced transport costs due to lower moisture product.

Table-2 Cost comparison of dry beneficiation processes

Process	Product quality, Kcal/kg	Yield, %	Process operating cost, $/t	Cost delivered to power station, $/Kcal
Conventional	5947.11	84.2	1.79	1.94
Rare Earth Magnetic Separator	6281.5	68.4	1.55	2.16
Air Dense Medium Fluidized Bed Separator	6281.5	80.6	1.91	1.91 (minimum)
Electrostatic Separator at Mine	6639.75	59.9	5.01	2.65
Electrostatic Separator at Power Station	6639.75	59.9	1.42	2.51
Air Table	6281.5	71.4	1.78	2.12

2. HISTORY AND DEVELOPMENTS

Major efforts have been made in China for developing high efficiency dry coal beneficiation methods. A dry beneficiation technology with air dense medium fluidized bed separation has been under development by Mineral Processing Research Centre of China University of Mining and Technology (CMUT), since 1984 by Q.Chen & Yufen Yang.

The first dry coal beneficiation plant in the world with air dense medium fluidized bed has been established by CUMT for beneficiation of -50 +6 mm size coal. The first commercial sized dense medium fluidized bed separator with **7,00,000 TPA of –50+6 mm coal size had been operational since 1994 in Heiongjiang province of China.**
It is claimed that high separation efficiency Ep value in the range of 0.05 to 0.07 is achievable for the coal in the range of size –50+6 mm.

Two kinds of dense medium were used for the system like

- magnetic pearl for low density medium.
- fine magnetite powder for high density medium.

In some cases fine coal powder have been added to the magnetite medium for creating a stable fluidizing bed.

An air dense medium fluidized bed separator was designed and tested at Institute of Minerals and Material Testing lab, (CSIR) Bhubaneswar, India in 2003 for beneficiation of high ash Indian non-coking coal in the size range of -25+6 mm. The overall dimension of the unit is 1507 mm x1022 mm x 130 mm width. The capacity of the unit is 600 kg/ hr using high ash Indian coal. The ash percentage of feed coal is around 40%. The characteristics of this type coal is totally different in comparison with other coal. The near gravity material (NGM) is very high (15-20%). At the optimum condition, the ash percentage of the feed could be reduced from 40 to 34% with 70% yield of product.

3. AIR DENSE MEDIUM FLUIDIZED BED BENEFICIATION (SEPARATION) PROCESS--(ADMFB PROCESS)

The air dense medium separation uses a dense fluidized medium of air and fine magnetite particles for beneficiation of any material. By means of a two phase **gas-solid pseudo-fluid** separating medium, the light and heavy particles stratify in the fluidized bed according to their individual densities. The bed density is more or less same throughout fluidizing region. ***As the bed density of medium is presumed to be equal to the separating density and the distribution of pressure in fluidized bed is the same as in the static fluid***, the motion of particles in the bed has been considered to explain the mechanism of the beneficiation process.

3.1 PRINCIPLE OF AIR DENSE MEDIUM FLUIDIZED BED SEPARATOR

It is known that the general beds of particles fluidized by liquids behave very differently from those fluidized by gases. In case of fluidized by liquid, it almost behaves homogeneous manner with increasing liquid flow rate, whereas gas beds are characterized by existence of bubbles. When fine powder is fluidized by gases, the bed expands homogeneously up to a certain point over a range of gas velocity extending well beyond minimum fluidization at which bubbles start to appear. The void fraction of the dense phase is greater than at incipient fluidization where the velocity of the gas increases, then the homogeneous or particulate system transfers to heterogeneous or aggregate fluidization.

It has been reported that the addition of fines in a bed of small particles improves the quality of fluidization by increasing the minimum bubbling velocity and the extent of particulate expansion. Between minimum fluidization and minimum bubbling velocity, fluidized bed of fine powder can exhibit particulate expansion to a large extent.

The current stage of progress in establishing theoretical criteria for the transition from Particulate to aggregate fluidization has been reported by different mathematical forms 2, 3, 4, 5 & 6. An attempt was made in the present investigation on the stability of fluidized bed using magnetite fine powder by air in dense medium fluidized bed separator under particulate behaviour. The concept of this has been utilized later for beneficiation of non-coking coal in dry process.

3.2 MATHEMATICAL EQUATIONS FOR CHECKING DYNAMIC STABILITY OF ADMFB

A bed of particles resting on a distributor for a uniform up-flow of gas is considered. The onset of fluidization occurs when the drag force by upward moving gas is equal to the weight of the particles. It can be expressed mathematically as;

(Pressure drop across bed) (Cross sectional area of the bed) =

(Volume of the bed) (Fraction consisting of solids) (specific weight of solid)

$$\Delta p \times A_t = A_t L_{mf}\left(1-\varepsilon_{mf}\right)\left[\left(\rho_s-\rho_g\right)\frac{g}{g_c}\right] \quad (1)$$

The superficial velocity at the minimum fluidizing condition

$$U_{mf} = \frac{d_p{}^2\left(\rho_s-\rho_g\right)g}{150\mu}\frac{\varepsilon_{mf}{}^3\phi_s{}^2}{1-\varepsilon_{mf}}, \mathrm{Re}_{p,mf} < 20 \quad (2)$$

For fine particles, the values recommended by Wen et al[4], gives

$$\mathrm{Re}_{p,mf} = \left[(33.7)^2 + 0.0408Ar\right]^{0.5} - 33.7 \quad (3)$$

Where Ar = Archimedes number

$$Ar = \frac{d_p{}^3 \rho_g (\rho_s - \rho_g) g}{\mu^2} \tag{4}$$

The quality of the particulate fluidization can be predicted using the following expression (Froude number)(Reynolds number)(bed aspect ratio)(density ratio)<100

$$(Fr_{mf})(\mathrm{Re}_{mf})(L_{mf} / D_t)((\rho_s - \rho_g)/ \rho_g) < 100 \tag{5}$$

$$Fr_{mf}=u_{mf}^{2}/gL_{mf}$$

Abrahamsen et al.[9], represent the expansion of non-bubbling beds of fine powders by

$$\frac{\varepsilon^3}{1-\varepsilon}\frac{d_p{}^2(\rho_s - \rho_g)g}{\mu} = 210(U - U_{mf}) + \frac{\varepsilon_{mf}{}^3}{1-\varepsilon_{mf}}\frac{d_p{}^2(\rho_s - \rho_g)g}{\mu} \tag{6}$$

The plotting of $\epsilon^3/1- \epsilon$ against superficial air velocity gives a straight line on a logarithmic plot which will govern the dynamic stability of the bed.

3.3 SEPARATION MECHANISM WITH AIR DENSE MEDIUM FLUIDIZED BED

In dry air dense medium fluidized bed separator, a medium is created by suspending solid particles in an upward direction of air flow. This acts in the same way as hydraulic dense medium separator, allowing clean coal to float to the surface of the medium and rejects to sink**.** By means of a two phase **gas-solid pseudo-fluid** separating medium, the light and heavy particles stratify in the fluidized bed according to their individual densities. The bed density is more or less same throughout fluidizing region. ***As the bed density of medium is presumed to be equal to the separating density and the distribution of pressure in fluidized bed is the same as in the static fluid***, the motion of particles in the bed has been considered to explain the mechanism of the beneficiation process.

The forces affecting a coal particle immersed in a fluidized bed are gravity, pressure, frictional drag forces between air and coal particle, air and medium particle and medium and coal particle. It is the fluid like characteristics of the air dense medium fluidized bed that separates the heavy and light particles. Some of the characteristics of the bed are:

a) The air flow expands the bed surface,
b) The pressure drop between two points in the bed is almost equal to the difference between the hydrostatic heads of same points,

$$\Delta P = P_1 - P_2 = (h_1 - h_2)\rho \tag{7}$$

c) The bed demonstrates **pseudo fluid** flow characteristics.
d) Particles with relative density less than the bed density float to the top surface of the bed while particles denser than the bed density sink towards the bottom of the system, hence,

$$\rho_1 < \rho < \rho_2 \tag{8}$$

Where

$\rho =$ density of bed

$\rho_1 =$ density of light particle

ρ_2 = density of sink particle

Pseudo fluid characteristics in an air dense medium fluidized bed separator form a stable and uniform gas–solid suspension with a certain bulk density in which light particles float and heavy particles sink in the suspension medium.

The average density of the bed is ρ_b given by

$$\rho_b = (1-\epsilon)\rho_s + \epsilon\rho_a = W/(LAg) \tag{9}$$

Where,

ε = porosity of bed, %

ρ_s = density of the medium, kg/m^3

ρ_a = density of air, kg/m^3

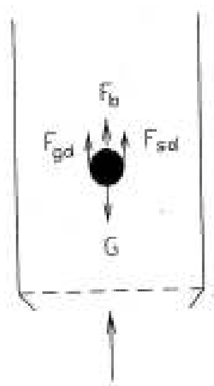
W = total weight of the medium, kg

L = depth of the bed, m

A = cross section area of the fluidized system, m^2

g = acceleration due to gravity, m/s^2

Archimedes principle is generally referred to explain the mechanism of separation of coal particles in the ADMFBS. But Archimedes principle can not explain the actual experimental results, in which a small amount of light and heavy products are misplaced during the separation process. Therefore, motion of the medium solids should be considered to study the mechanism of separation in ADMFBS.



Compressed air

Fig. 1 Forces exerted on a spherical coal particle in an ADMFB

Therefore, making the force balance on the coal particle

$$G = (\pi / 6)\rho_c d_c^3 (a + g) = F_b + F_{gd} + F_{sd} \tag{10}$$

Where,

$G =$ gravitational force on coal particle, Newton

F_b = effective buoyancy force exerted on the coal particle, Newton

F_{gd} = friction drag force of gas exerted on the coal particle, Newton

F_{sd} = drag force of air dense medium exerted on the coal particle, Newton

ρ_c = density of the coal particle, kg/m^3

d_c = equivalent diameter of the coal particle, m

a = acceleration of the coal particle, m/s^2

g = gravitational acceleration , m/s^2

μ = viscosity of medium, kg/m.s

The axial pressure distribution of fluidized bed can be expressed as

$$-dp/dx = \left[(1-\varepsilon)\rho_s + \varepsilon\rho_a\right]g = \rho_b g \quad (11)$$

Where,

ρ_b = density of bed, kg/m^3

ρ_a = density of air, kg/m^3

ρ_s = density of medium solids, kg/m^3

ε = void fraction in the bed, %

The effective buoyancy force is

$$F_b = (\pi/6)\rho_s d_c^3 g \quad (12)$$

The general drag formula of fluid, F_{sd} can be expressed as

$$F_{sd} = (\pi/6)C\rho_b d_c^2 |U_r| U_r \quad (13)$$

Where,

C = drag co-efficient,

U_r = relative velocity between coal particle and dense medium particles, m/s

As the diameter of a coal particle is much larger than that of the medium solid, the friction drag force of the gas exerted on coal particle can be neglected.

The equation on simplification becomes

Acceleration, $a = (\dfrac{\rho_b - \rho_c}{\rho_c})g + \dfrac{C\rho_b}{\rho_c d_c}|U_r|U_r$ (14)

Let terminal velocity $U_t = \sqrt{\dfrac{d_c(\rho_b - \rho_c)g}{C\rho_b}}$ (15)

U_t is the terminal velocity of coal particle in a fluidized bed when there is no relative motion between the coal particle and the medium .

When U_r is 0 i.e., there is no relative velocity between coal particle and fluidizing medium, acceleration $a = (\rho_b - \rho_c)g / \rho_c$ the coal particles can be perfectly separated according to the bed density,

and if $\rho_b = \rho_c$, $a = C\rho_b / (\rho_c d_c)|U_r|U_r$.

The position of coal particles in the bed depends on the relative velocity between the coal particle and medium solids.

Separation efficiency is generally decided by the motion of the coal particles (i.e., misplacing effect of viscosity) and the velocity of medium solids (misplacing effect of motion). At too low gas velocity the activity of medium solids is weak, which results in a higher misplacing effect of viscosity.

Moreover, when the medium of solids are not well dispersed and gas can not be uniformly distributed in the bed, the misplacing effect of motion is also larger in some areas of bed. At too high gas velocity, back mixing of the medium solids caused by gas bubbles is enhanced and the misplacing effect of motion will also be intensified. With decrease of coal size, the specific surface increases and the terminal velocity of coal particle decreases, resulting in the increase of ratio of drag to gravity exerted on the coal particle and the increase of the misplacing effect of both viscosity and motion. Therefore, the gas velocity of fluidization should be chosen according to the optimal velocity for the separation of feed stock with minimum size.

4. MATERIALS AND UTILITIES

4.1 Material :

- The magnetite powder was collected from UCIL, Jaduguda. The true density of the magnetite powder is 4.7 gm/cc .
- The coal used is taken from Mahanadi coal fields ltd, Ib Valley, Brajrajnagar. The proximate analysis of the coal sample gives **43.9% ash content.**

4.2 Utilities :

- The filter cloths were used as air distributor supplied by M/s Weaverbird Fabrics, Nagpur and M/s Popat Brothers, Mumbai.

Table-3 : Technical Specification of filter cloth

Commercial name	Mesh approximately	Pore size in micron approximately	Air permeability*
NLN-80	450	8-10	0.23
TLN-1001	450	8-10	NA
PL-2510	500	8-10	5.2
PL-2511	550	3-5	4.9
P260B	500	5-8	1.6
1405	NA	NA	0.053
2525	NA	NA	0.18
2620	NA	NA	0.14

*Air permeability data is supported by manufacturer

- Here we are checking PL-2510, 1405, 2525, 2620. Among all the above PL-2511 is the best distributor but as it is not available, so we'll use the next better one i.e. 2620.
- Compressor with air accumulator connected through silica gel tower.
- Rotameter (air) 0-100 lpm range
- Manometer 2 nos.(Hg and CCl_4)
- A strong magnet of wide field for separating coal and magnetite particles.

5. EXPERIMENTAL WORK

5.1 EXPERIMENTAL SET UP

The experimental apparatus used in this study consists of three main parts; the air supply, beneficiation apparatus and the pressure measurement of the bed. Two different set ups were fabricated using perspex tube having diameter 50 and l00mm with height of 1250 and 1400 mm respectively. The air flow rate was measured using rotameter with range of 0-100 litre/minute. The filter cloths were used as air distributor. The pressure drop across the distributor was measured with a manometer between one tapping placed on plenum chamber and other just above the air distributor.

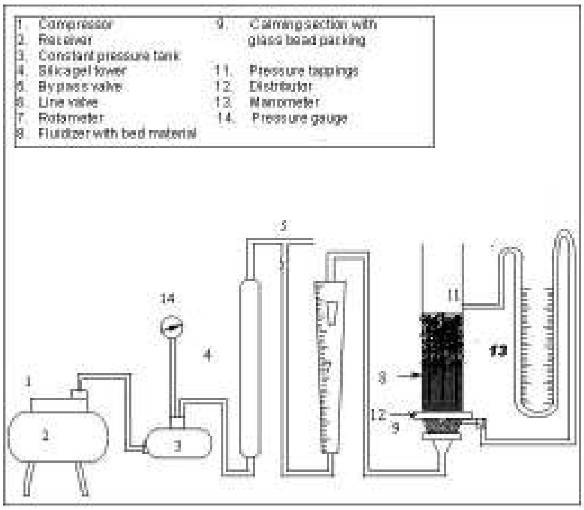


Fig-2 Schematic representation of the experimental air dense medium fluidization set up

Fig-3 Setup used for the ADMFB experiment

5.2 EXPERIMENTAL PROCEDURE

5.2.1 FOR CHECKING THE BED STABILITY

Different filter cloths were used as distributor in 50 mm set up. Based on the bed expansion and onset of bubbling, the 2620 filter cloth was used as air distributor for further study. The compressed air was used in the system without any magnetite powder to determine the pressure drop across the distributor only in 50 mm set up. Then the dried magnetite powder was used to study the bed expansion at different air flow rate. The quantity of magnetite powder was varied in the succesive experiments(refer to sample tables: 4 and 5 and corresponding figures 8(a) and 9(a). The expansion of the bed was studied at different air flow rate till the minimum bubbling starts in the bed.

Then same procedure was followed for 100 mm setup. The density of bed at different air flow rates and corresponding voidage were determined by using the measuring parameters of height of the expansion bed, true density of the solid and weight of the sample.

Table-4 Bed height vs air flow rate for varying quantities of 100µm magnetite powder in 50 mm setup.

	Height of bed, cm			
air flow rate, lpm	200 gm of magnetite	300 gm of magnetite	400 gm of magnetite	500 gm of magnetite
0	6	9	12	15
5	7	10.5	13	17
10	8	11.5	15	19
15	9	13	16	20.5
20	10	14	18	23
25	10.5	15.5	19	24
30	11	16	20	25
40	13	18	24	29
50	14	18.5	26	30

Table-5 Bed expansion vs bed density for varying quantities of 100µm magnetite powder in 50 mm setup.

air flow rate, lpm	200 gm of magnetite		300 gm of magnetite		400 gm of magnetite		500 gm of magnetite	
	Bed expansion cm	Bed density gm/cc	Bed expansion cm	Bed density gm/cc	Bed expansion cm	Bed density gm/cc	Bed expansion cm	Bed density gm/cc
0	0	1.698	0	1.698	0	1.698	0	1.698
5	1	1.456	1.5	1.4558	1	1.568	2	1.572
10	2	1.274	2.5	1.329	3	1.359	4	1.341
15	3	1.132	4	1.176	4	1.274	5.5	1.243
20	4	1.019	5	1.092	6	1.132	8	1.108
25	4.5	0.970	6.5	0.986	7	1.073	9	1.061
30	5	0.926	7	0.955	8	1.019	10	1.019
40	7	0.784	9	0.849	12	0.849	14	0.878
50	8	0.728	9.5	0.826	14	0.784	15	0.849

5.2.2 FOR DRY BENEFICIATION OF COARSE HIGH ASH COAL

After a uniform and stable ADMFB is formed, the coal of given size was put into the fluidized bed. After 20 minutes, the air supply was cut off and bed samples were taken from four equidistant bed heights (s1, s2, s3, s4) (Fig-6) using a vaccum pump. After separation of magnetite, the ash of the coal samples from four equidistant bed heights were analyzed by taking one gm of coal sample in a open silica crucible and heating in muffle furnace at temp 750±15ºC for 1and ½ hrs. This was repeated with varying the air flow rate and the sizes and amount of coal and magnetite. The scope of the experiment is presented in table-6.

Table-6 SCOPE OF EXPERIMENT

d_{pc}	d_{pm}	Wc	Wm	Air flow rate	D_t
1 mm	100µm	150 gm	100 gm	20 lpm	50 mm
5 mm	80µm	175 gm	100 gm	25 lpm	100 mm
10 mm	60µm	200 gm	100 gm	30 lpm	
	45µm	225 gm	100 gm	35 lpm	

5.2.3 SAMPLE COLLECTION PROCEDURE

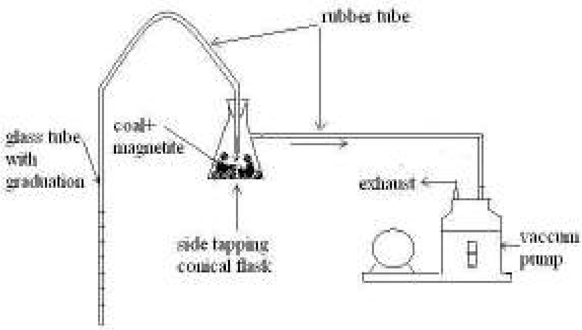


Fig-4 Primary design for collecting coal and magnetite sample from fluidization column.

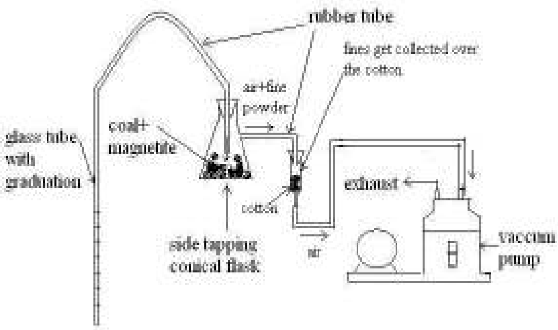


Fig-5 Modified design for collecting coal and magnetite sample from fluidization column.

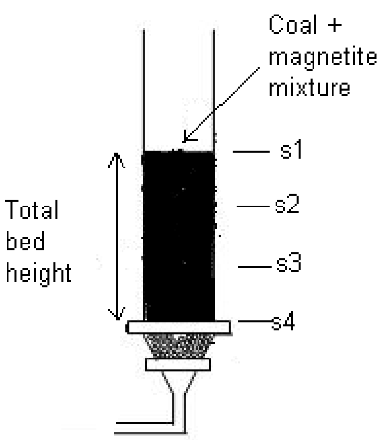


Fig-6 points for collection of coal and magnetite mixture sample after beneficiation.

In the primary design (Fig-4) while collecting the sample, a significant amount of fines entered into the suction line connected to the vaccum pump resulting chocking of the pump. So to rectify it, collection design was modified (as shown in Fig-5) so that the fines instead of entering into the suction line got arrested over the cotton packing.

6. INTERPRETATION OF EXPERIMENTAL DATA

6.1 STUDY OF DISTRIBUTOR BEHAVIOUR

The pressure drop across the air distributor using 2620 filter cloth was measured. The result of this study is given in Fig-7. The graph is almost a straight line which governs the uniform behaviour of the distributor.

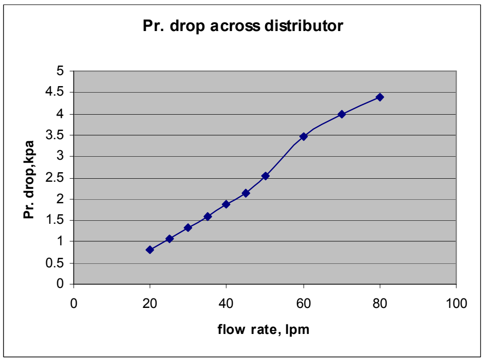


Fig-7: Pressure drop across the distributor without any material in 50 mm setup.

6.2 BED STABILITY ANALYSIS IN 50 MM DIA SETUP

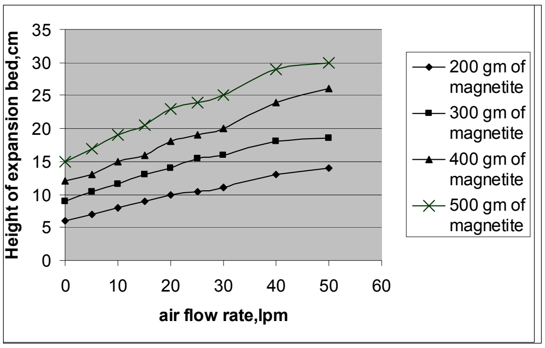


Fig.8a: Bed expansion at different air flow rate for 100µm magnetite

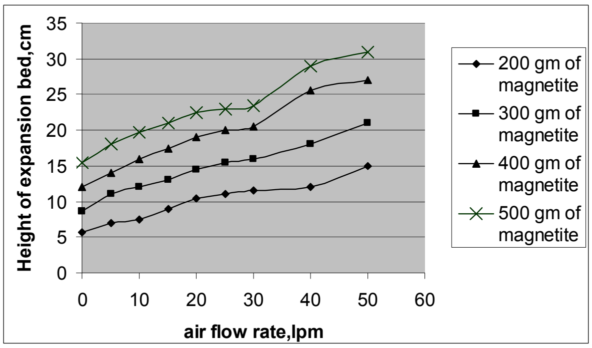


Fig-8b: Bed expansion at different air flow rate for 80µm magnetite

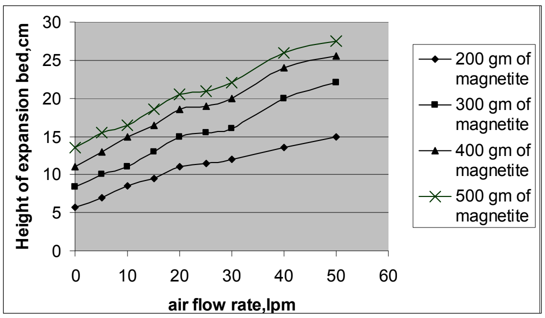


Fig-8c:Bed expansion at different air flow rate for 60 µm magnetite

As it can be observed from fig.8(a),(b) and (c) for 100µm, 80µm and 60µm magnetite powder (in 50 mm dia setup) a stable region (almost constant bed expansion) is obtained at flow rates of **20-30 lpm**, so we can say that this is the particulate fluidization region and operating under non-bubbling condition.

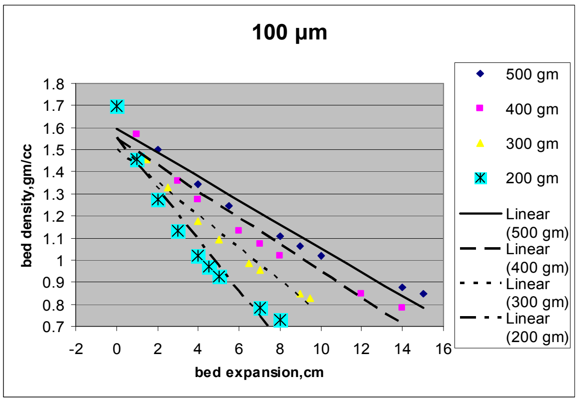


Fig-9(a) Bed expansion versus bed density for 100µm magnetite

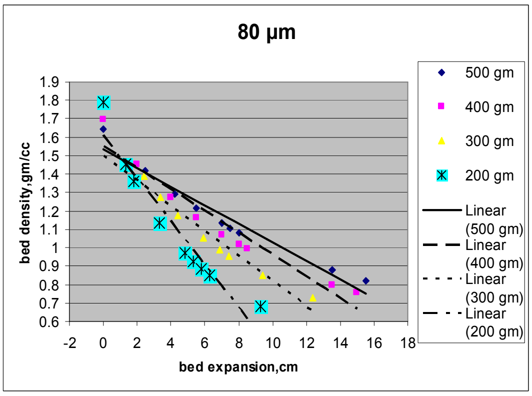


Fig-9(b) Bed expansion versus bed density for 80µm magnetite

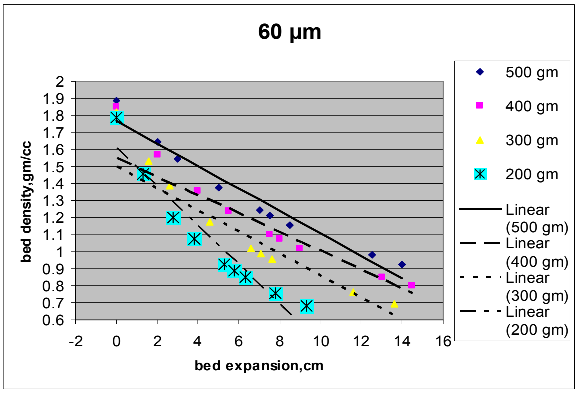


Fig-9(c) Bed expansion versus bed density for 60µm magnetite

Further for 100µm magnetite, more uniform curves (straight lines) are obtained when bed expansion is plotted against bed density (fig.9(a)). So **100µm magnetite** is selected. It shows that the bed expansion is linear at different quantities of magnetite powder, hence the system is in non-bubbling condition.

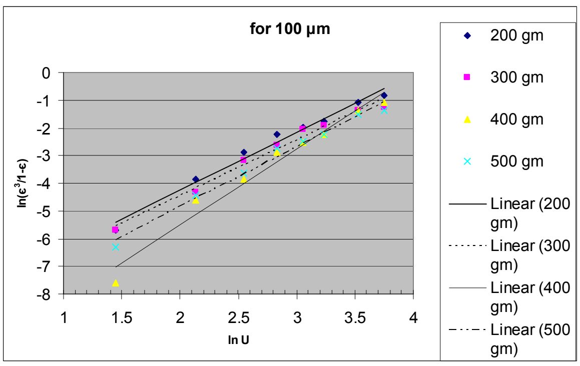


Fig-10(a)Fluidization characteristics for 100µm magnetite in 50 mm dia setup

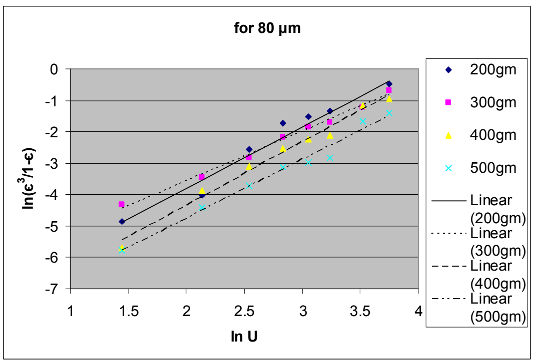
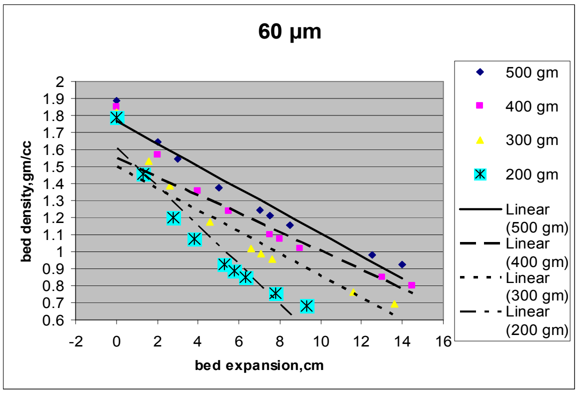


Fig-10(b) Fluidization characteristics for 80µm magnetite in 50 mm setup



10(c) Fluidization characteristics for 60µm magnetite in 50 mm dia setup

Again the characteristics of the fluidized bed was examined using equation (6) and the plotting of $\epsilon^3/1-\epsilon$ against superficial air velocity gives a straight line on a logarithmic plot. The results are given in Fig-10(a), (b) & (c).For 100µm magnetite, the converging trend indicates that the Fig-system operates in non-bubbling condition.

6.3 BED STABILITY ANALYSIS IN 100 MM DIA SETUP

To see the effect of diameter on bed stability and thereafter on beneficiation, the larger dia (100 mm) fluidized bed was used for a mixture of coal and 100µm magnetite powder in varying amounts.

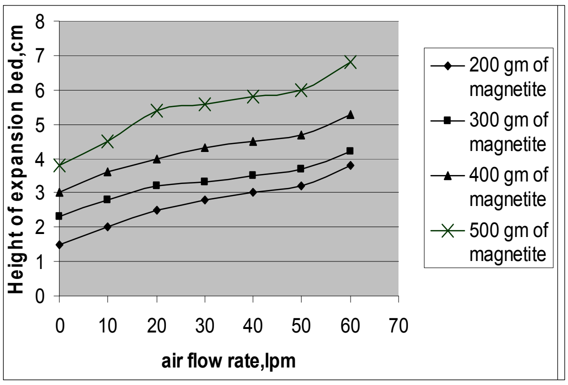


Fig-11: Bed expansion vs air flow rate for 100µm magnetite powder (100 mm dia setup)

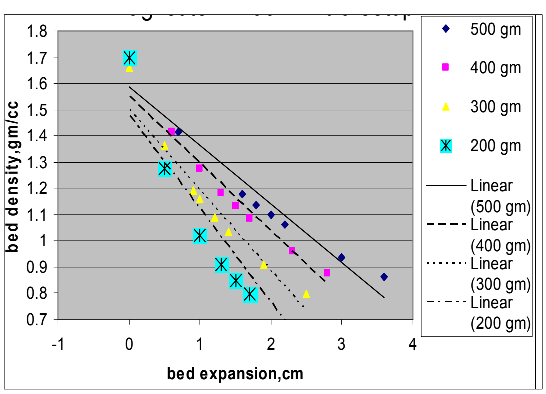


Fig-12: Bed expansion vs bed density for 100µm magnetite powder (100 mm dia setup)

As it can be observed for 100 mm dia setup almost constant bed heights are obtained for air flow rates in the range 25-50 lpm. Further for 100µm magnetite powder the bed expansion vs bed density plot (Fig-12) and Fluidization characteristics plot ($\ln(\epsilon^3/1-\epsilon)$ vs ln U) (Fig-13) give satisfactory results.

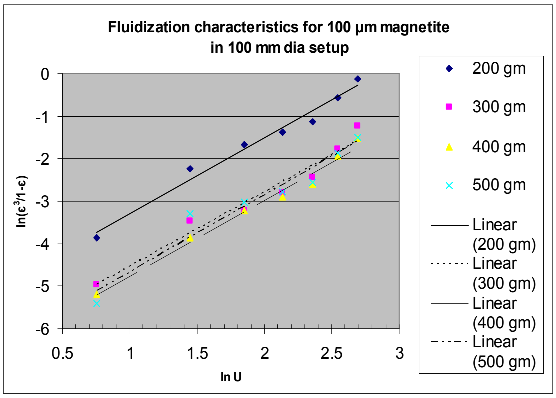


Fig-13: Fluidization characteristics for 100µm magnetite powder in 100 mm dia setup

Now let's calculate the minimum fluidizing velocity for the 50 mm dia setup.

Wt. of magnetite=400 gm

Initial bed height $L_{initial}$=12.0 cm

Mean dia. d_p=100µm=0.01cm

Air viscosity µ=0.00019 gm/cm.sec

Particle density ρ_s=4.7 gm/cc

Air density ρ_g=0.0012928 gm/cc

Dia of column D_t=5.0 cm

Bed height at minimum fluidization L_{mf}=16 cm

L_{mf} / D_t =16/5=3.2

$(\rho_s - \rho_g)/ \rho_g$ =3634.52

Ar = Archimedes number $Ar = \dfrac{d_p{}^3 \rho_g (\rho_s - \rho_g) g}{\mu^2}$ =165.07

For fine particles, the values recommended by Wen et al;

$$\mathrm{Re}_{p,mf} = \left[(33.7)^2 + 0.0408 Ar\right]^{0.5} - 33.7 = 0.0998$$

ε_{mf}=(L_{mf} - $L_{initial}$)/ $L_{initial}$ =0.25

As Rep,mf<20

Hence ,the superficial velocity at the minimum fluidizing condition,

$$U_{mf} = \frac{d_p{}^2 (\rho_s - \rho_g) g}{150\mu} \frac{\varepsilon_{mf}{}^3 \phi_s{}^2}{1-\varepsilon_{mf}} = 0.336 \text{ cm/sec or } 0.4 \text{ lpm}$$

So we need to operate above this flow rate.

Froude number $Fr_{mf}=u_{mf}^2/gL_{mf}$= $7.19*10^{-6}$

The quality of the particulate fluidization can be predicted.

$\left(Fr_{mf}\right)\left(\mathrm{Re}_{mf}\right)\left(L_{mf}/D_t\right)\left(\rho_s-\rho_g\right)/\rho_g$ =0.0083 <100

Hence well with in the particulate fluidization range.

Similarly the minimum fluidization velocity for the bed (coal + magnetite) was calculated to be U_{mf}=0.784 cm/sec =0.923 lpm and checked for particulate fluidization.

Air flow rate used=25 lpm i.e. around 30 times U_{mf}.

6.4 COAL ENRICHMENT ANALYSIS

The various coal samples were taken by varying Wc/Wm, d_{pc}/D_t, d_{pc}/d_{pm} and $(G_f-G_{mf})/G_{mf}$.

Table-7 Ash content of beneficiated coal (A_{bc}) Initial ash content A_c=43.9 %

Run no	d_{pc} mm	d_{pm} µm	Wc gm	Wm gm	Air flow rate, lpm	Column dia,mm	Ash content, A_{bc} %			
							S1	S2	S3	S4
1	1	100	200	100	25	50	32.5	34.5	37.5	40
2	1	80	200	100	25	50	36	36.5	38	39
3	1	60	200	100	25	50	37	37	37.5	41
4	5	100	200	100	25	50	34.5	37	38.5	40
5	5	80	200	100	25	50	36.5	38	38	41
6	5	60	200	100	25	50	38	39	39.5	41.5
7	10	100	200	100	25	50	35	38.2	38.7	41.1
8	10	80	200	100	25	50	37	38.3	38.8	41.4
9	10	60	200	100	25	50	38	39	40.6	42
10	1	45	200	100	25	50	36.4	37.6	39.3	41
11	1	100	150	100	25	50	36	36.5	38.5	40.1
12	1	100	175	100	25	50	36	37.1	37.4	41.3
13	1	100	225	100	25	50	32.9	35.2	37.4	40
14	1	100	200	100	20	50	34.6	35.2	37.1	41.4
15	1	100	200	100	30	50	32	35.1	36.8	41
16	1	100	200	100	35	50	32.3	35	36.5	41
17	1	100	200	100	25	100	32.6	33.4	35.2	36

The magnetite particles from the samples were separated by magnetic separation and the coal samples were subjected to ash analysis. The ash content of the above coal samples after dry beneficiation is given in Table-7 from which the percentage Enrichments were calculated using equation (16).

$$E= (A_c-A_{bc})/A_c*100= (43.9- A_{bc})/43.9*100 \quad (16)$$

Table-8 Enrichment of coal after dry beneficiation in ADMFB

Run no	Ash content, A_{bc} %				Enrichment, E %			
	S1	S2	S3	S4	S1	S2	S3	S4
1	32.5	34.5	37.5	40	26	21.4	14.6	8.88
2	36	36.5	38	39	18	16.8	13.44	11.16
3	37	37	37.5	41	15.72	15.72	14.58	6.6
4	34.5	37	38.5	40	21.4	15.72	12.3	8.88
5	36.5	38	38	41	16.86	13.44	13.44	6.6
6	38	39	39.5	41.5	13.44	11.16	10.02	5.46
7	35	38.2	38.7	41.1	20.27	12.98	11.84	6.38
8	37	38.3	38.8	41.4	15.72	12.76	11.6	5.69
9	38	39	40.6	42	13.44	11.16	7.52	4.33
10	36.4	37.6	39.3	41	17.08	14.35	10.48	6.6
11	36	36.5	38.5	40.1	18	16.86	12.3	8.65
12	36	37.1	37.4	41.3	18	15.49	14.8	5.9
13	32.9	35.2	37.4	40	25.06	19.82	14.8	8.88
14	34.6	35.2	37.1	41.4	21.18	19.82	15.49	5.69
15	32	35.1	36.8	41	27.11	20.04	16.17	6.6
16	32.3	35	36.5	41	26.4	20.27	16.86	6.6
17	32.6	33.4	35.2	36	25.74	23.9	19.82	18

7. RESULTS AND DISCUSSIONS

7.1 DEVELOPMENT OF CORRELATION

Considering acceptable value of ash in the beneficiated samples as 36 %, the values of Enrichment for the first two sample points in case of each of the runs have been used in the development of correlation and its testing with the additional experimental points.

The enrichment of coal (E) can be expressed as a function of various system and operating parameters as:

$$E= f(Wc , Wm, D_t, d_{pc}, d_{pm}, G_f, G_{mf}) \qquad (17)$$

From Dimensional analysis approach , the above equation can be represented in terms of dimensionless groups as

$$E= k\left[\left(\frac{W_c}{W_m}\right)^a\left(\frac{d_{pc}}{d_{pm}}\right)^b\left(\frac{d_{pc}}{D_t}\right)^c\left(\frac{G_f-G_{mf}}{G_{mf}}\right)^d\right]^n = k\,(\text{product})^n \qquad (18)$$

Where a, b, c and d are the exponents of the individual groups calculated from log-log plots of E vs. the groups. (Fig.14, 15, 16 and 17 respectively)

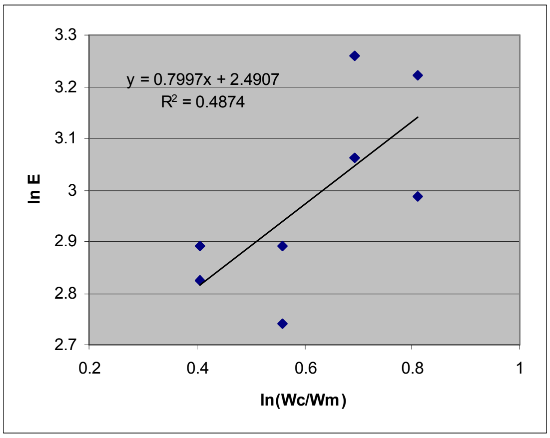


Fig.14: ln E vs ln (Wc/Wm)

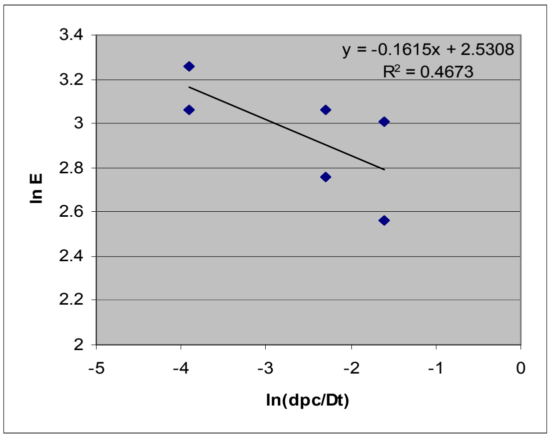
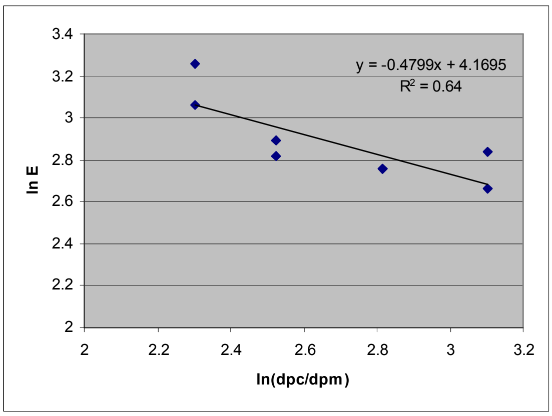


Fig.16: ln E vs ln (dpc/Dt)



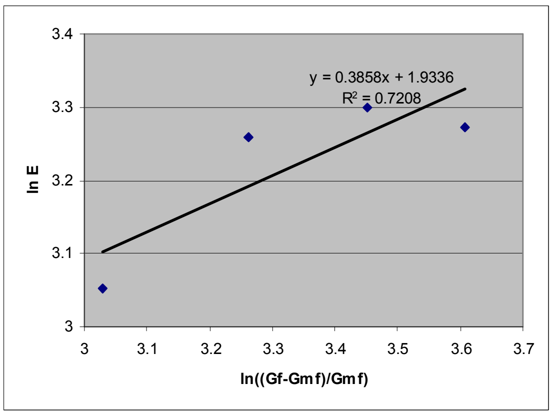


Fig.15: ln E vs ln (dpc/dpm)

Fig.17: ln E vs ln (Gf-Gmf)/Gmf)

Values of the exponents obtained are a=0.800 b=-0.480 c= -0.162 d=0.386

The final correlation is obtained by plotting the product $\left[\left(\frac{W_c}{W_m}\right)^{0.8}\left(\frac{d_{pc}}{d_{pm}}\right)^{-0.48}\left(\frac{d_{pc}}{D_t}\right)^{-0.162}\left(\frac{G_f-G_{mf}}{G_{mf}}\right)^{0.386}\right]$ against E on a log-log plot (as shown in fig.18) .The data which are used for plotting Fig-18 are given in Table-9.

So we get, k=7.715 n=0.8179

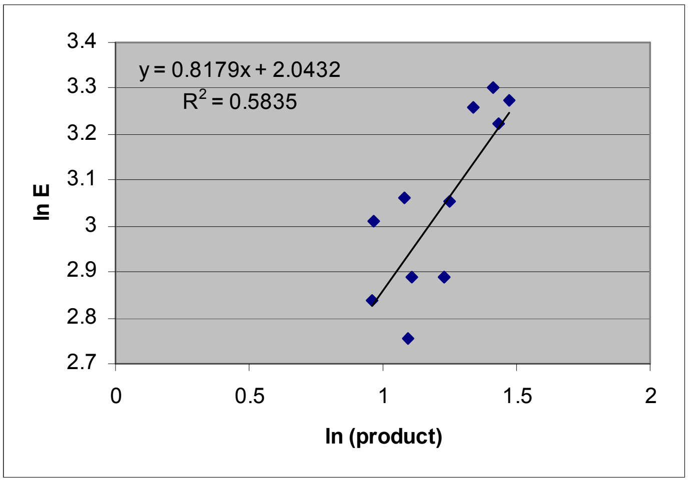


Fig.18: ln E vs ln (product)

Table-9 Data used for plotting ln E vs ln product

Wc/Wm	dpc/dpm	(Gf-Gmf)/Gmf	dpc/Dt	Eexpt
1.5	10	26.08	0.02	18
1.75	10	26.08	0.02	18
2	10	26.08	0.02	26
2.25	10	26.08	0.02	25.06
2	12.5	26.08	0.02	18
2	16.67	26.08	0.02	15.72
2	22.22	26.08	0.02	17.08
2	10	20.68	0.02	21.18
2	10	31.52	0.02	27.11
2	10	36.88	0.02	26.4
2	10	26.08	0.1	21.4
2	10	26.08	0.2	20.27

Thus the final correlation obtained is:

$$E=7.715\left[\left(\frac{W_c}{W_m}\right)^{0.6541}\left(\frac{d_{pc}}{d_{pm}}\right)^{-0.3925}\left(\frac{d_{pc}}{D_t}\right)^{-0.1321}\left(\frac{G_f-G_{mf}}{G_{mf}}\right)^{0.3155}\right] \tag{19}$$

Now, the rest experimental enrichment data (other than used for calculating the correlation) were used to verify the above correlation.

The enrichment values calculated from the above correlation have been compared with the experimental values of enrichment (fig.19) and fairly good agreement has been found.

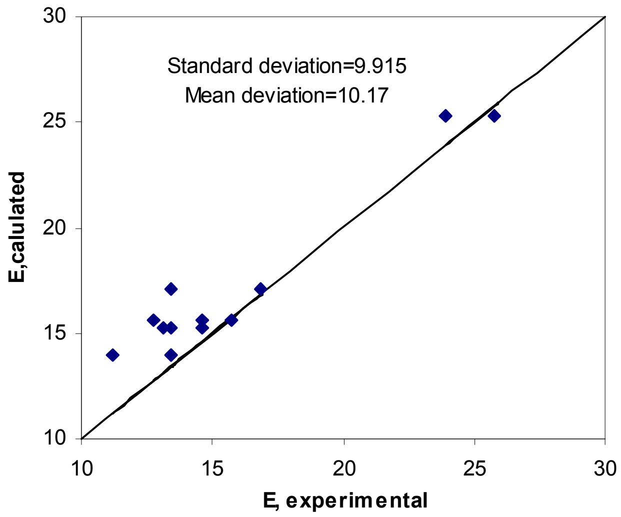


Fig-19 Comparison plot between the experimental and the calculated values of enrichment

7.2 DISCUSSION ON RESULTS AND CONCLUSION

It has been found that enrichment is a direct function of coal to magnetite and excess velocity ratios. But both of the above variables Contribute to high bed pressure drop marginally with the increase in power consumption. Enrichment is inversely proportional to coal particle size as indicated by the negative exponent of the of the other two groups but there is a limiting size from the operational convenience as fine coal particles will create environmental problem and the beneficiated small-sized coal particles will have limited downstream uses.

From the above considerations, coal and magnetite particles of sizes 1 mm and 100μm respectively in the amount ratio 2:1 with 30 times the minimum fluidization velocity has been found to be effective for enrichment in a 50 mm diameter fluidized bed reducing the ash content of the coals of Mahanadi coal fields ltd, Ib Valley, Brajrajnagar from 43.9% to nearly 36% on an average with 60 percent yield of product. The single test in a large diameter (100 mm) fluidized bed also corroborates the above conclusion.

8. POTENTIAL APPLICATION OF THE PRESENT PROJECT WORK

In view of appreciable beneficiation achieved with respect to the reduction of ash content obtained in the present experimental studies and tested for a relatively larger size equipment, there is an excellent potential for the application of this result ***for further scale up in the pilot plant/semi-commercial and commercial unit*** for the ultimate use of the presently-discarded high ash coal with dry beneficiation for various process applications viz. thermal power plant, sponge iron unit and coke ovens.

9. NOTATIONS

A_c=ash content of coal before beneficiation, percentage

A_{bc}=ash content of beneficiated coal, percentage

Ar = Archimedes number

$Ar = \frac{d_p{}^3 \rho_g (\rho_s - \rho_g) g}{\mu^2}$, dimensionless

A_t= Area of cross section of the fluidized bed, cm^2

d_p = particle diameter, cm

d_{pc}= coal particle diameter, cm

d_{pm}=magnetite particle diameter, cm

D_t = Diameter of the fluidization column, cm

E= Enrichment, percentage

Fr_{mf}= Fraud number at minimum fluidization

=u_{mf}^2/gL_{mf} , dimensionless

g =Acceleration due to gravity, cm/sec^2

G_f=mass velocity of air at any flow rate, $gm/cm^2.sec$

G_{mf}= mass velocity of air at minimum fluidization velocity, $gm/cm^2.sec$

L = Bed height, cm

L_{mf} = Bed height at minimum fluidization, cm

Re_{mf}= Reynolds number at minimum fluidization

=$(d_p U_{mf} \rho_g)/ \mu$, dimensionless

U_{mf}= superficial velocity at minimum fluidization, cm/sec

W_c=Weight of coal taken, gms

W_m= Weight of magnetite taken, gms

ϵ_{mf}= Bed voidage at minimum fluidization

ρ_s= Particle density, gm/cc

ρ_g = air density, gm/cc

Δp = pressure drop across bed, bar

Φ_s= sphericity of particle

μ= Air viscocity, gm/cm.sec

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