A FIRST ORDER ASSESSMENT OF THE MOVING GRANULAR PANEL FILTER

by

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DECLARATION

I claim that this is my original work and that it has not been submitted in this or in a similar form for a degree at any University.

G D H Shaw

SUMMARY

The capabilities of a specific moving granular panel bed filter are investigated, highlighting FBC flue gas cleanup. The continuous operability offers some advantages over the discontinuous operating mode of the static filter bed but the effects of moving granular medium are found to introduce other disadvantages not originally anticipated.

Movement of granular medium resulted in significantly reduced filtration efficiencies with respect to static granular beds for the same operating conditions. Pressure drop analysis indicated voidage increases within the bed with granular movement but the increases were not considered significant enough to explain the reduced efficiencies suffered by the moving filter medium. Reentrainment at high Stokes numbers is considered to be the main contributor for reduced filter ability. An empirical correlations was developed to model the re-entrainment effects.

The causes and effects of non uniform granular flow were highlighted and resolved by modification to the panel design. The modifications involved the inclination of the panel a few degrees off the vertical. The precise angle was found to depend on fluid flowrate through the filter medium. The ability of the continuous medium replenishment to maintain low fluid pressure drop was demonstrated.

The concept of initial collection efficiency, used in static granular beds as a conservative measure of the filter efficiency in granular filter design, is shown not to be applicable to moving granular beds. Collection efficiencies were found to deteriorate progressively from initial clean granule conditions.

Increase in panel thickness is shown to be necessary to improve the filter efficiency to that required to meet the emission regulations. The resulting increased operating pressure drop is calculated to be in the order of that imposed by an equivalent electrostatic precipitator making the design, an attractive low pressure drop filter, assuming sufficient improvement in collection efficiency can be achieved with the increased panel thickness.

Order of magnitude cost estimates are presented at two different scales. Comparisons of costs with conventional filter designs show the design to be worth further investigation.

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NOMENCLATURE

B'	= Stokes-Einstein diffusion coefficient	$m^2.sec^{-1}$
B	= fluid inertia correction for filter efficiency	-
c	= particle concentration in flue gas	$g.(Nm)^{-3}$
Cin	= inlet dust concentration to filter	$g.(Nm)^{-3}$
$Cout$	= outlet dust concentration from filter	$g.(Nm)^{-3}$
Ci	= dust concentration in ith unit collector	$g.(Nm)^{-3}$
Cs	= Cunningham slip correction	-
d_g	= filter medium particle size	m
dp	= dust particle size	m
d_c^*	= dimensionless constriction diameter	-
E	= overall filter efficiency	%
F_s	= sand flow rate	$g.min^{-1}$
g	= gravity	$m.sec^{-2}$
Ga'	= Galileo No with modified gravity term	-
K_1	= constant Eq 17	-
l	= unit collector size	m
L	= bed depth	m
L_R	= ratio louvre thickness to panel length	-
m_p	= particle mass	kg
N	= number of unit cells	-
N_G	= gravitational parameter	-
N_I	= interception parameter	-
N_{Re}	= Reynolds number	-
N_{ST}	= Stokes number	-
P	= filter pressure drop	mmH_2O
Pe	= Peclet's No	-
R	= Boltzmann's constant	$J.kg^{-1}.K^{-1}$
t	= time coordinate	sec
T	= temperature	K
u_t	= terminal settling velocity	$m.sec^{-1}$
V_f	= fluidisation velocity	$m.sec^{-1}$
V_L	= flue gas velocity in louvre exit	$m.sec^{-1}$
V_{Lmax}	= flue gas velocity in louvre exit at failure	$m.sec^{-1}$

V_{max}	= limiting filter face velocity	$m.sec^{-1}$
Vs	= flue gas superficial velocity	$m.sec^{-1}$
Z	= average flow path length through filter	m
α	= louvre angle	degrees
γ	= adhesion probability	-
ε	= filter porosity	-
θ	= re-entrainment probability	-
η	= unit collector efficiency	%
ρ_p	= dust particle density	$kg.m^{-3}$
ρ_g	= flue gas density	$kg.m^{-3}$
μ	= flue gas viscosity	centipoise

CHAPTER 1

1.0 INTRODUCTION

Industrial air pollutants are recognised as major contributors to the degradation of atmospheric air quality. The impact of poor air quality is not well understood but growing awareness of its contribution to community health and welfare is reflected in the ever increasing air pollution regulations. As these regulations become more stringent they become progressively more difficult to meet and introduce additional capital and operating expenses the tighter they become. In addition, technology development has placed greater demands on the current cleanup techniques to an extent where existing equipment is unable to meet the requirements in particular cases.

A source of significant industrial pollution is fly ash particulate discharged from coal fired power plants. Although the control of air pollution on these plants is generally accepted as an integral part of the capital and operating costs, these costs are high, and tend to result in the bare minimum control required to meet the regulations being implemented. One cannot fault industry in its attempts to meet the prevailing regulations at minimum cost but it would be far easier to improve overall air quality if air pollution control costs were reduced. This is especially so for countries where the regulations are based on the principle of "best practical means" i.e. where the feasibility and cost of air pollution control ultimately determine the regulated pollution limits. These factors provide the incentive for research and development of aerosol filtration systems such as the project being tackled here.

High collection efficiencies for fine particles have been demonstrated for granular bed filters resulting in a number of designs being developed to make use of the good collection efficiency. This project is a first order assessment of one particular granular bed design.

The objectives of the work are to:

1) evaluate the use of a moving granular bed filter for fly ash capture.

A panel bed filter capable of continuous medium regeneration was chosen for the experiment. This specific filter design, known as the Moving Granular Panel Filter (MGPF), employed a thin granular filter bed, which was allowed to flow down in a panel retained between a series of angled louvres. Particles were caught in the filter medium as the dust laden stream was forced through the bed. Recycling the filter medium continuously through the panel and a subsequent regeneration step enabled continuous operation and to a certain extent, control of operating pressure drop.

2) assess filtration efficiency while maintaining low operating pressure drop with high granular recirculation rates.

Faster recirculation /regeneration can be employed to reduce particle loading in the filter bed, minimising air flow restriction and hence pressure drop through the filter. The lower pressure drop would reduce fan power costs, possibly improving the economics of this granular bed design. Lower particle loading within the granular medium improves granular flow characteristics but this operating condition is not typically suited to good granular bed filtration. Normally effective filtration through granular beds relies on building up an accumulation of particles on or within the bed and subsequent improved collection efficiencies of the loaded granular bed.

However, it is widely reported that a certain degree of particulate filtration can be achieved in granular beds without relying on particle loading in the granular medium. This is demonstrated by the collection efficiencies reported for the initial clean filter bed conditions. The ultimate objectives of most filtration research is to develop theoretical models for design and development under loaded filter bed conditions, but complexity introduced by deposit morphology has forced many investigations to simplify to clean filter bed conditions where models need not account for the effect of accumulated deposits within the granular matrix. Clean filter bed characteristics are thus well researched and numerous correlations have been developed to predict performance of beds in this condition. The understanding is that, although difficult to model during operation, filter efficiencies improve as particulate is accumulated within the bed and as the initial collection efficiency can be modelled with more reasonable accuracy it provides a simple but conservative filter design tool.

3) determine whether the granular filter design can be used as a primary separator ie with no precleaning

Generally, the granular filters are used in a secondary or tertiary role downstream of one or two cyclone stages. This project aims at using the granular bed as the primary cleanup device. With no upstream cleanup, the filter must cope with the total fly ash produced by the combustion process over the entire particle size range.

With regard to the overall viability of this specific granular filter, (MGPF), it should be reckoned that improvement of filtration economics through reduction of operating pressure drop is not unrealistic, as, for example, fan power costs over the life of a bag house filter, which also relies on accumulation of particulate deposit for effective filtration, can exceed twice the origanal capital cost of the filter. Before any definitive statement can be made concerning the specific granular filter under investigation, it is necessary to evaluate the performance of this design against that required by regulations and applications pertinent in South Africa. Probably a more realistic application for this filter is in the protection of turbine blades in combined cycle technology. The success of the combined cycle concept depends upon removal of particulate from combustor/gasifier flue gases without loss of temperature. This severe constraint precludes most conventional cleanup techniques where either the materials of selection of device internals are not suited to such high temperature operation (typically 900 - 1000 ^{o}C) or separation efficiencies fall off markedly at these high temperatures. The use of an inert granular material will allow this filter to tolerate these extreme conditions but its capture efficiency will need to be compared with that required for turbine operation. These two applications thus form the ultimate goal for this work.

This project was intended as a first order assessment of the MGPF for fly ash capture and thus experimental work was conducted accordingly. The experimental trial were divided into two sections, the first dealing with the technical development of a stable moving bed filter under cold operating conditions and the second, a study of the filter's performance under more realistic operating conditions, namely with real fly ash at higher temperatures.

The report that follows includes a review of commercially available filter designs aswell as designs still being developed for high temperature applications. The relevant theory on granular filtration is also reviewed to assist with the analysis of the filtration characteristics displayed by this particular design. As part of a first order assessment projections of the fullscale design and economics are investigated. Finally guidelines for the extension of this work are discussed.

CHAPTER 2

2.0 SURVEY OF THE LITERATURE

2.1 GRANULAR BED FILTRATION

Granular filter beds are described as particle separation devices using solid granules as filter media; the particle laden stream is passed through a dense bed of granular material where the particulate is separated. Generally, the granular medium is used to promote the formation of a particulate matrix generated by the accumulation of particulate from the stream flowing through the filter bed on or within the granular medium. The filtration ability attributed the granular medium is enhanced by the subsequent action of the particulate matrix formed. The formation of the particulate matrix is dependant on the clean granular material being able to trap particulate from the flow stream passing through. This process of particulate accumulation, can in itself, be considered as filtration. It has been shown that granular material can accumulate particulate from the clean condition, the rate of accumulation increasing with particle loading in the bed. It is this loaded bed condition and correspondingly higher filtration efficiencies upon which most granular filters are based. However, along with increasing filtration efficiency, the accumulated particulate also restricts fluid flow area and thereby increases pressure drop suffered by the fluid flowing through. In most granular filter designs, high pressure drop is generally accepted as part of the granular filter option as will be shown in the review that follows. Pressure drop is not always the most important consideration in a filter design, but can be where fan power costs required to overcome the resistance to flow, influence the overall economics.

The literature on aerosol filtration in granular beds can be split into two. The authors either tackle the theoretical modelling of granular filtration mechanisms or the mechanical details required to make use of the granular media. It is more appropriate to include the documented theoretical work in a separate chapter, so this chapter will be devoted to the mechanical configurations required to make use of the granular medium capabilities as a filter. The theory will be covered in the following chapter.

Granular bed filtration is a well known technique but is traditionally associated with cleanup of liquid streams(1,2,3). Its application to aerosol cleanup is relatively new and presently restricted to a few specific industries. The majority of these aerosol applications are found in dedusting of clinker cooler exhaust air generated in the cement industry. Gravel beds, so called because of the granular size used, have proven to be more superior in this application than other collectors (4).

Liquid and gas filtration through granular media involve the same filtration mechanisms, but the contribution by each individual mechanism varies for the two cases. Theoretically therefore, some meaningful information could be acquired through studying literature covering liquid filtration. It was felt that as the fluid properties of liquids and gases do alter the significance of the various filtration mechanisms, it would be more approriate to concentrate aerosol filtration in this project. Consequently, the bulk of the discussion that follows focuses on aerosol filtration.

The original configuration of a granular filter was a bed consisting of layers of gradually decreasing size granules. The larger sizes formed the base of the filter bed and supported layers of successively finer granules. During filtration the fluid was forced down through the granules, trapping particulate deep within the bed. The filtration mechanism responsible for the particulate capture was simply known as "depth filtration".(5) Essentially the same approach is employed in the more recent filters but, in terms of analysis, the "depth filtration" mechanism has been more carefully resolved and broken down into a number of different mechanisms. Being of a theoretical nature, detailed discussion of these mechanisms will be covered in the following chapter.

In simple terms, the depth filtration process relies on restricting the fluid and entrained particulate flow to narrow tortuous channels between the granules, where any contact with the filter medium along the way may cause the particle to stick to the granule surface.(6) Physical restriction, immobility, as well as the attractive forces between the particle and granule surface, cause the particle to be retained on the granule surface, resulting in capture. The ability of the filter medium to capture and retain the particulate is referred to as the filter efficiency. It is defined as the mass fraction of the inlet particulate that does not emerge downstream of the filter. Filtration efficiencies depend firstly on bringing the particulate into contact with the filter medium and then secondly ensuring that the particle sticks to the granule. The filtration efficiency is also affected by re-entrainment where particulate already trapped is re-suspended by the scouring action of the fluid flowing past. Particles in the 1-5 micron range are reported by numerous authors (1) to be the most difficult to trap. The reason for this is that particle/granule contact is hampered by the fact that these particles are able to follow the fluid flow streamlines, of which few actually approach the the granule surfaces close enough for capture and thereby reduce particle/granule contact. Without contact the particles are not captured and pass right through the bed. After contact between particle and granule the particle/granule adhesive forces prevent reentrainment of the particulate.(1) The adhesive forces need to be sufficient to keep the particle stationary between cleaning cycles or until the granule is removed from the filter in the moving granular bed filtration. Detailed description of the mechanisms and their effect on efficiency will be discussed in the chapter covering the filtration theory.

It is well known that filtration efficiencies vary over the duration of the filtration cycle. Initially, when the granular bed is clean, filtration efficiency is lowest. As particles are caught the flow channels become partially blocked and efficiency improves as the particulate matrix is established. Unfortunately the increasing pressure drop associated with the accumulated particulate eventually restricts flow to such an extent that cleaning or regeneration is required. The cleaning or regeneration cycle has to be regularly implemented when the pressure drop reaches a preset limit. As a result of these recurrent cleaning operations, the particle loading in the filter bed varies continuously. The intermittent filtration/regeneration technique is thus characterised by a fluctuating filtration efficiency and operating pressure drop. This condition is not ideal but overcome to some extent by using a number of filter units in parallel. By staggering the cleaning cycles the pressure drop and efficiency fluctuations are minimised. Numerous authors report using this filter configuration in the cement industry(4) but none report their use in coal combustion flue gas cleanup. No conclusive reason is given for the exclusion of granular filters from coal combustion flue gas clean up.

2.2 FILTER MEDIUM REGENERATION

In addition to the complication of trapping and retaining particulate in a granular filter bed, there is an increase in pressure drop associated with constriction of flow channels by the trapped particulate. High dust loadings increase the pressure drop very quickly making it necessary to clean or regenerate the filter medium for prolonged filter use. The cleaning of the granular medium is difficult and presents a major draw back to granular bed filtration.

The simplest cleaning technique used is back flushing (or reverse flow) of air. The back flushing is generally accompanied by agitation of the bed material to assist in dislodging the captured particulate. Agitation is generated by mechanical raking of the filter medium during the reverse flow cycle and, being effected in situ, cannot be done during the filtration cycle. The filter must therefore be isolated from the supply ducting during the cleaning cycle. This means the filter unit has intermittent cycles of filtration and regeneration. Cleaning is initiated when the capture of sufficient inflowing dust particles increases the pressure drop across the filter bed to a preset maximum. During the regeneration cycle, the removal of the particulate from the interstitial voids opens up the fluid flow channels and reduces the pressure drop.

A variation of the reverse flow regeneration technique is described by Squires.(7) The concept is similar in that flow reversal is required but the regeneration concept is entirely different. The simple reverse flow is used with deep bed filtration where the particulate is caught within the bed volume. The filter described by Squires is based on trapping the particulate on the surface of the filter bed. Careful selection of the filter media induces the particulate to build up or cake on the granular surface during the initial stages of the filtration cycle. The sieving action of the cake is then used to filter the fluid with very high efficiency. Squires reports being able to dislodge the filter cake by a short "puff back". The cake and a small amount of granular media are dislodged and a fresh filter surface exposed. Once again an intermittent operation of cleaning and filtering cycles is required. The filtration technique is not reported to be in use on any commercial units but has been shown to work on an experimental scale. (7)

A novel regeneration technique used in conjunction with reverse flow is reported to be part of a commercial filter made by Lurgi(Germany).(8) The filter design is aimed at emulating the high efficiencies of fabric filtration by the use of a very fine granular medium in the filter. The granules are supported in a narrow panel with filtration being accomplished using crossflow through the bed. When regeneration is required the granular media is circulated through the panel and up a central air lift pipe. The aerated circulation and reverse flow (as reported by Lurgi) produce an efficient filter regeneration technique. The concept is relatively new and designed as a retrofit for the downflow beds used in the cement industry but no installations have as yet been cited. The use of the fine granular filter medium is reported not to depend on dust agglomeration characteristics for filtration and thereby touted as a potential candidate for flue gas dedusting in coal fired power plants.(8) Little information is available on the precise filtration mechanisms involved but it would appear that as no cake formation is required, deep bed filtration mechanisms apply. This should then mean shorter cycle times but lower operating pressure drop.

In contrast to the intermittent cleaning operations discussed, several continuous filtration /regeneration concepts have been reported. The designs depend on removal of dust laden granules from the filter unit for cleaning in an external system. The continuous removal of the granules generates a moving bed in the filter unit. The granules are cleaned externally and then recycled back to the filter. The moving granular bed is thus able to filter continuously. It is also operated with a fixed pressure drop related to the particulate loading in the filter bed which is a function of the media circulation rate. High recirculation rates can be used to keep the bed cleaner, facilitating a lower pressure drop because of less interstitial obstructions. The mechanical details of various moving granular bed filter designs are documented in the literature.(9,10,11,12) The designs are based on deep bed filtration mechanisms and all attempt to avoid any caking of particulate in the bed. The prevention of surface cake formation, as used by Squires, is achieved by high granular recirculation rates. The cleaning of the dirty granules is carried out during pneumatic conveying of the media. The scouring action of the pneumatic air is reported to be sufficient to clean the media. Final collection of the dust is then accomplished by secondary filtration of the pneumatic exhaust air. The volume of pneumatic air is substantially smaller than the total effluent flow being filtered and a small baghouse filter is thus sufficient for cleanup of the pneumatic air stream(9).

In special circumstances, the difficulty of filter medium regeneration is reported to be avoided completely. The application is known as a Fixed Filter Bed and is limited to use on effluent streams with very low particle loadings. A large filter bed is used so that the gradual accumulation of particulate within the bed material does not increase the pressure drop very quickly. Filter beds are designed to operate for a period and then be replaced by new ones. Life spans of five years are reported to be possible, but these depend on low inlet particle loading(5). This design however does not have any relevance to this project because the dust loadings from coal fired installations are generally high.

A concept proposed by Dutkiewicz,(10) also allows the cleaning operation to be avoided. The concept relies on the particulate in the effluent stream being of some material value. The effluent stream is filtered through a bed of product material from the process concerned producing a saleable, blended product containing the captured fines. The application is designed to use a moving granular panel filter. The same concept is suggested using a raw material from the process. The raw material would be passed through the panel as the filter medium and trap particulate leaving the dust producing operation. The fines and raw material emerging from the filter would then be returned back to the process avoiding the difficulties of granular medium regeneration. This project, although not involved with product recovery, is the extension of Dutkiewicz's work to fly ash removal from FBC systems. The filter configuration used for this project is same as that used in Dutkiewicz's filter for metallurgical dust recovery.

2.3 INDUSTRIAL GRANULAR BED DESIGN

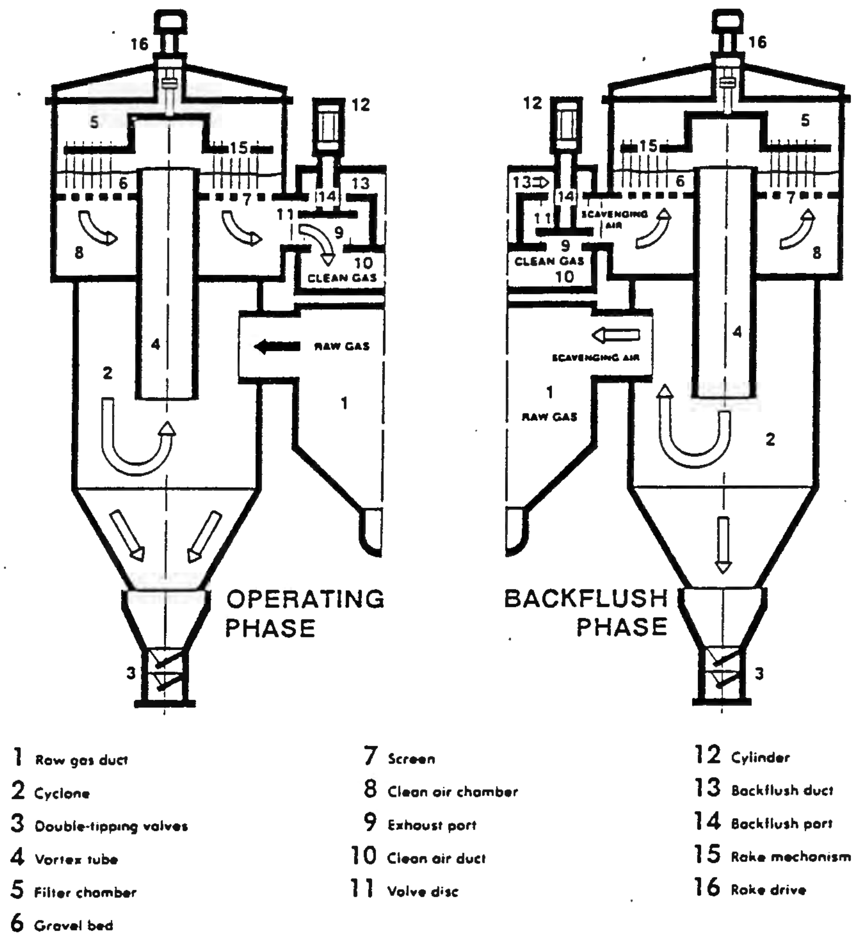
The original concept of a stratified filter medium has, in most cases, been replaced by an homogeneous bed of uniformly sized granules. The actual filter configuration used, i.e. the vessel design and operational parameters, has varied considerably since the original introduction of granular bed filtration. The different designs employed can be categorized by the fluid flow direction namely: downflow, up-flow (including fluidised filter beds) and crossflow. In general, different cleaning operations are suited to particular filter configurations. Continuous moving beds are typically in panel bed form relying on gravity to induce the solid flow down between two screens or series of louvres with crossflow of dust laden gases. The static filter bed configuration is generally a bed supported on a retaining screen with effluent flow either upward or downward through the filter bed. Reviewed below are a number of industrial applications currently in use.

2.3.1 HORIZONTAL FILTER BEDS (DOWN/UP-FLOW)

This particular granular filter is used extensively in the cement industry. Cement fines are removed from off gases by a bank of filter beds operating in parallel. The term horizontal refers to the orientation of the exposed filter face and is synonymous with down and up flow configurations. The filter bed is usually supported on a screen installed in a tubular vessel representing one filter module. Invariably more than one module is used in an installation which allows one unit to be isolated during periodic cleaning of the filter medium. Cleaning is usually done with reverse flow techniques assisted by mechanical agitation as described earlier. Included in each vessel is an integral upstream cyclone used for primary precleaning. Both the dust removed by the cyclone and the dust dislodged by the reverse flow cleaning operation are removed through a double-tipping valve at the base of this cyclone. In later models, a much larger external cyclone or bank of cyclones is used as a precleaner with improved results. The later model also includes a dedicated cyclone used to remove the particulate from the reverse flow air. In the larger scale units the vessel is modified to accommodate two filter beds one above the other to save on the floor space required.(4)

The present design limits gas temperatures to 400- 500°C and atmospheric pressures with cleanup capabilities down to 25 mg/Nm^3.(4) A diagram of the filter is shown below:

FIGURE 1: HORIZONTAL BED FILTER



drawing reproduced from Scheuler (8)

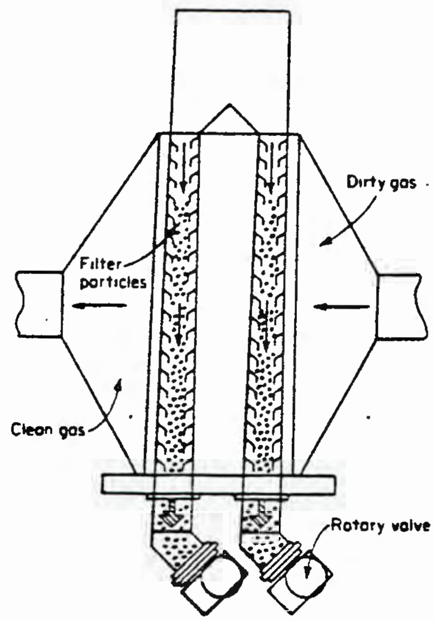
This filter design is reported to depend on particulate agglomeration to achieve high filtration efficiencies (13). For this reason the poor agglomeration characteristics of coal based fly ash (8) may be responsible for precluding its use on cleanup of coal based flue gases. During regeneration, the particle agglomerates carried in the reverse flow air, are of sufficient size that they can be removed from the reverse flow air stream by cyclonic action. Thus, without agglomeration, the reverse flow air stream cleanup would have to be modified.

2.3.2 PANEL BED FILTERS

The panel bed filter is a tall narrow panel of granular material. Filtration is achieved by crossflow of the dusty gas stream through the panel. The granules are retained by a mesh or a series of louvres to expose a vertically orientated face area. The granules are either kept stationary or allowed to move down through the panel during the filtration cycle. When stationary filter medium is used, the trapped particulate is removed by intermittent reverse flow similar to the principle used in the horizontal filter bed. When the medium is allowed to flow during the filtration cycle, the dust laden granules are removed from the base and cleaned externally. The cleaned filter medium is then returned to the top of the filter panel setting up a continually regenerated filter panel which is able to filter particulate from a gas stream continuously.

An example of a moving granular bed filter is the Dorfan Impingio (5) panel filter designed in 1950. The granules were retained in two vertical panels by a series of louvres. Dirty gas was forced through both panels in series during which time the particulate was trapped in the vertical filter beds. The dust laden bed was gradually regenerated by removal of granules from the base at a rate controlled by two rotary valves. The granules were then cleaned by screening or some other means and then returned to the top of the panel.(7). A sketch of the filter is shown below:

FIGURE 2: DORFAN IMPINGIO MOVING GRANULAR PANEL FILTER



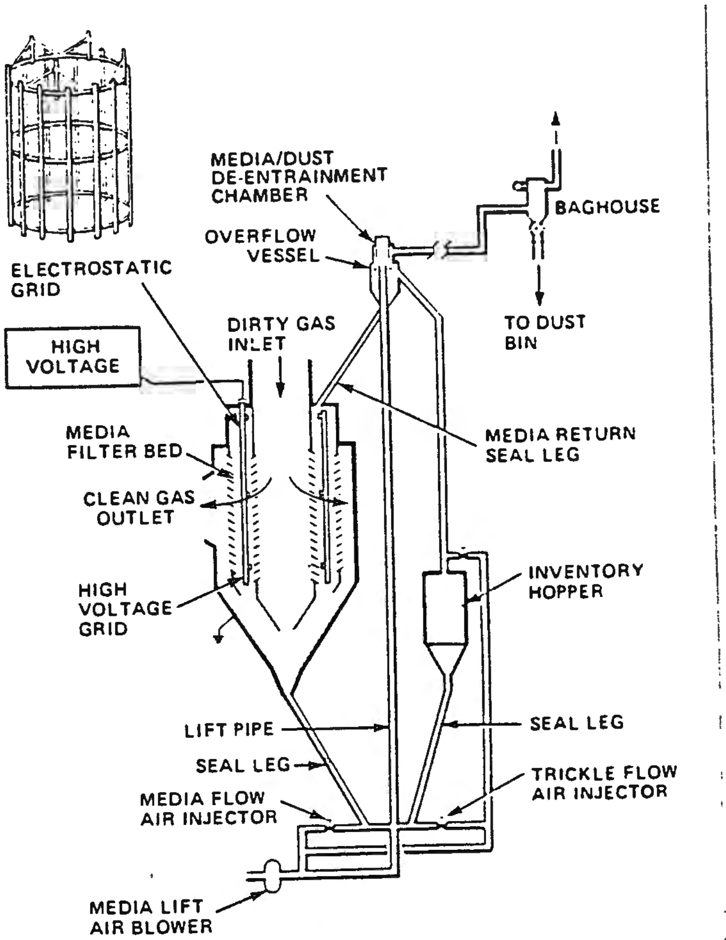
reproduced from Perry (5)

This project uses a filter concept similar to the Impingio design. The differences between the two are in the size of filter medium granules used and the louvre design. Little information on the performance of the Impingio design was found in the literature so it was impossible to determine if the filter was being operated with clean granule conditions but this does seem unlikely. Several authors made reference to the design (5,7) but reported conflicting results. In one case, the filter was reported efficiencies of 96% on asbestos dust(5) while tests carried out by the Environmental Protection Agency (EPA) in America indicated that the filter suffered high re-entrainment problems after steady-state conditions had been reached. Unfortunately the details of these reports were not available.

2.3.3 ELECTROSCRUBBER FILTER

Another example of a moving granular bed filter is the Electroscubber Filter made by Combustion Power Company in the USA. The filter was developed for flue gas cleanup on wood fired boiler installations and references to several successful operations are made in an article by Storms et al.(14) The design is an improvement over the Dorfan Impingio filter in that the granular bed is augmented by an induced electrostatic field. A diagram of the filter is shown below:

FIGURE 3: ELECTROSCRUBBER (TM) MOVING GRANULAR PANEL FILTER



drawing reproduced from von Reiche (11)

The inherent charges on the dust particles, together with an electrostatic field, provide a large improvement in collection efficiency of the fine particles (1-5 micron). The gravel, in this design, is continuously recycled through the panels via a pneumatic flow system which doubles as a granule cleaning system. During transport, the dirty gravel is cleaned by the turbulent air flow and the dust is re-entrained into the transport air. Clean gravel is returned to the top of the filter panel and the dust collected in an auxiliary baghouse filter by filtering the pneumatic transport air. The air volume is considerably less than the main stream with high dust concentrations allowing a small baghouse unit to be used for the cleanup.

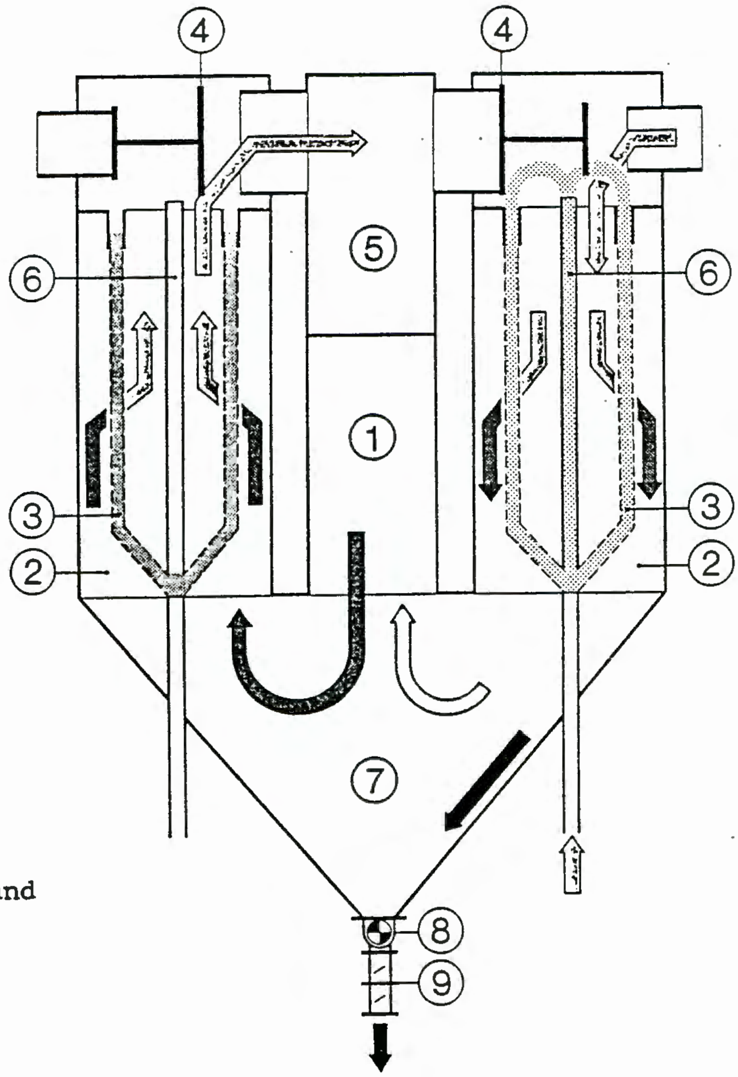
Improvement of this design is believed to be possible by increasing the surface charge on the particulate entering the filter and not relying on the inherent surface charges. The system contains a precharging section upstream of the Electroscrubber(TM). Research is currently being done in Israel, Japan and USA. (11)

The only drawback in the design is that although high collection efficiencies due to the electrostatic effects are reported, high pressure drop through the bed is also evident.

2.3.4 GRANULATE-TUBE-FILTER(TM)

Lurgi are currently marketing an additional design called the Granulate-Tube-Filter(TM).(8) This design uses a panel bed filter of very fine granules with a porosity reported to be comparable to fabric bag filters. The panel is designed as a retrofit for horizontal bed filters and also relies on the precleaning action of integral or upstream cyclones. The filter bed material does not move during the filtration cycle, as with the early Dorfan Impingo design. This design requires intermittent cleaning where the trapped dust is removed from the loaded bed by backflushing. During the cleaning cycle the fine granulate bed material is circulated in an aerated state up through a central air lift pipe and down the filter panels. The action results in an effective separation of collected dust from the granules. A schematic representation of the Lurgi design is shown below.

FIGURE 4: GRANULATE-TUBE-FILTER



Operating mode (left) and backflush mode (right)

1 Raw Gas Duct
2 Filter Plenum
3 GRANOTUBE (TM)
4 Disc Gate (Double Disc)
5 Clean Gas Duct
6 Central Pipe
7 Dust Hopper
8 Dust Screw-Conveyor
9 Double Tipping Gate

drawing reproduced from Lurgi pamphlet (8)

The reverse flow and propulsion air for the air lift system are returned to the main inlet stream and distributed over the remaining filter units. The precleaning cyclone doubles as a settling chamber during the reverse flow cleaning cycle and has a double tipping gate at the base through which the dust is removed. The reverse flowstream is returned to the main inlet duct so that dust that does not settle out during the cleaning is then recaptured along with the main dust load. No known industrial application of this design is reported as yet but reference to its ability for flue gas dedusting on power plants is made.(8)

2.4 EMISSION REGULATIONS

All the designs described above are used to reduce industrial effluent particulate emissions. Each different industrial process produces different quantities of particulate with a range of properties. The choice of filter type depends on the specific particulate being emitted aswell as the local emission regulations enforced. Emissions from coal combustion represent a significant contribution to overall pollution and have emission regulations relating specifically to their control. The regulations differ from country to country. A number of different emission regulations are reviewed below.

Meetham (15) reports that instances of disease caused by dust inhalation in mines and factories are well understood and, under the general heading of Pneumoconiosis, include silicosis, asbestosis and other forms of reticulinosis. When inhaled, dust particles put many times their volume of lung out of action but Meethan indicates that there is no evidence of pneumoconiosis contracted by breathing street air.

However, in England, during the 1950's, 30000 people died every year through chronic bronchitis, a respiratory disease believed to be associated with the inhaling of air-borne pollutants. Although the damp foggy climate was partly to blame, there were sharp differences between the incidences of bronchitis in towns and rural areas. The comparison showed areas of heavy air pollution to be worst off indicating the adverse effects of air-borne pollutants. In the past three decades, introduction of regulations restricting emissions to the atmosphere has improved the situation considerably and far fewer deaths can now be directly attributed to air pollution. World wide emission regulations are however still constantly reviewed as additional detrimental effects of atmospheric pollution are recognised. Regulations restricting particulate emissions are included in the air pollution control because their effects are believed to be related to respiratory diseases in people employed within an industrialised sector(15).

In the USA, a fixed particulate emission standard is applied. (16) The Clean Air Act of 1970 limited particulate emission for coal-fired electric power plants to 43 mg/MJ of fuel input. In 1979, the Environmental Protection Agency (EPA) set out new standards, the New Source Performance Standards (NSPS), which reduced the particulate emission to 12.9 mg/MJ. Large reductions in permissible emissions have forced industry to improve standards of emission control and in many cases alternative cleanup processes are now required to meet the new standards. It is also noted, by Shannon (16), that future restrictions on dust emission could be aimed at limitation of the less than 5 micron size fraction rather than a total effluent restriction. This is based on the concern that the smaller size fraction is responsible for related human health problems such as pulmonary disease. A restriction like this would change the economic viability of several cleanup options and in particular, costs of the Electrostatic Precipitator, a collector used extensively on fly ash, would become prohibitively large. (16)

Fixed emissions standards also apply generally in the European Economic Community but in the UK and South Africa, pollution control is based on the principle of 'the best practical means'. Unlike the fixed standards, the regulations are flexible and the pollution source emission level is negotiable depending on the capability and economics of available cleanup systems. The principle also considers the effect of a pollution source in conjunction with the extent of pollution already present in the area. Presumptive limits for coal fired installations exist as part of the regulations but should not be seen as absolute limits. The function of these limits is simply to imply, by compliance, that the best practical means of control is in operation(16). Where pollution would not pose an obvious threat, the regulations are relaxed. The presumptive limits for particulate emissions within the UK and SA are shown below:

BRITISH (17)

1. 0.2% of coal fired for greater than 140 MWt output (equivalent to 115 mg/Sm^3)

2. 0.5% of coal fired for greater than 30 MWt output.

3. 1.0% of coal fired for less than 1.5 MWt output.

SOUTH AFRICA (18)

1. 50 mg/m^3 for coal consumption greater than 20 tons/h.

2. 120 mg/m^3 for coal consumption less than 20 tons/hr.

3. 400 mg/m^3 for coal consumption less than 10 tons/hr.

These limits are lenient when compared with the fixed standards imposed in the USA. By comparison, the SA presumptive limit of 50 mg/m^3 represents a permissible emission of roughly three times the American standard. The South African limits also become more lenient for smaller installations while the American fixed standard applies across the scale. As a result in S A, the effluent from a small industrial boiler burning 10 tons/hr of coal may be as high as 19 times the permissible limit in the USA.

24 coal fired power plants exist in South Africa today. B11 of these plants are small installations and, 13, large installations. The small installations are fitted with cyclones for particle capture with efficiencies of up to 80%. This figure is exceedingly low but because the plants are small and pollution regulations lenient, high collection efficiencies are not required. In addition, of these 11, 10 use chain grate combustion, a technique which produces relatively little fly ash. The chain grate combustion is typically used on smaller coal burning installations and the Pulverised Fuel combustion (PF) on larger plants. The PF combustion uses a finely powdered coal which produces large quantities of very fine fly ash. ESP is reported to be installed on all the large power plants achieving a national average cleanup efficiency

of 95.2%. This efficiency is also low but attributed to older plant designs.(19)

A program is underway to reduce existing power plant emission levels with the use of flue gas conditioning agents. In the worst cases new ESP units are to be installed. The high resistivity of the fly ash produced when burning low sulphur coals, characteristic to SA, affects the ESP capability. Control of the agent dosage allows resistivity to be reduced to optimum levels for ESP performance. It is hoped to increase the national average ESP collection efficiency to 98% by use of a sulphur trioxide conditioning agent.(19)

The cost of various particle capture systems is more dependant on required collection efficiency than others. ESP is one particular system whose cost is dependant on collection efficiency. Changes in emission regulations enforce higher filter efficiency requirements and may alter filtration costs considerably. The most economical option for large scale installations in SA is currently ESP but being sensitive to emission regulations this could change with the introduction of tighter regulations. However, revision of the regulations are not expected at least until the year 2000 so ESP will probably be used for some time to come. (20)

2.5 DEVELOPMENTS IN FILTER TECHNOLOGY

Although demand for the development of more economic filters, required to meet emission regulations, exists, the bulk of current filtration development is aimed at an entirely different requirement. Current filtration research is focused on the needs of advanced coal combustion techniques.

2.5.1 ADVANCES IN COAL COMBUSTION TECHNIQUES

The move away from conventional combustion techniques is related to current emission regulations but not necessarily those for particulates. It is the reduction of sulphur and nitrogen oxides from coal combustion flue gases that is of primary concern because it is believed that these pollutants are instrumental in the formation of the acid rain which is responsible for killing off of large forests in the northern hemisphere.

In addition, acid rain is believed to be responsible for the acidification of inland lakes in the same areas. The original idea that tall flue stacks would allow dispersion of the sulphur and nitrogen oxide emissions is now disputed and restrictions on these oxide emissions are now being introduced. Emission limits introduced are forcing industries to consider different cleanup technologies, different combustion techniques and even different fuel sources (with low sulphur content). One such combustion technique, Fluidised Bed Combustion (FBC) is an attractive concept, as it has the capability of reducing sulphur and nitrogen oxide emissions during combustion.

2.5.2 FLUIDISED BED COMBUSTION

The use of Fluidised bed combustion (FBC) is seen as one of the most promising techniques for coal combustion because it is capable of burning the most poor grade fuels while controlling both sulphur and nitrogen oxide emissions. Sulphur retention of 90% is possible by injection of calcareous sorbents into the combustor. Nitrogen oxide levels are kept low by the low combustion temperatures used in FBC (typically 850^{o}C). However the reduced sulphur content in the combustion flue gases, produces fly ash with high resistivity characteristics, making it difficult to capture by electrostatic precipitation.(39) Several different FB designs have been proposed, with the largest units being in the region of 200 MWe. The use of the FB combustion technique in these units has moved cleanup away from ESP to baghouse filtration because bag filtration does not depend on resistivity characteristics of the fly ash.(21)

2.5.3 COMBINED CYCLE FLUE GAS CLEANUP REQUIREMENTS

FB technology is used in several coal conversion/combustion techniques because of the advantages it affords in emission control. It is well known that one of these systems, the coal fired combined cycle power plant, demonstrates improved overall plant performance. The system is based on expanding hot, high pressure, combustion gases from a fluid bed combustor/gasifier through an open cycle gas turbine before raising steam in a conventional steam cycle.(22) The combined cycle technique is used successfully on liquid fuel combustion (without FBC) but is not quite as advanced for coal combustion. The stumbling block is the gas cleanup required upstream of the gas turbine. An alternative concept is to cool the gas prior to cleanup reducing the temperatures to a point where conventional filter systems can operate. This method is feasible, but at a loss of the thermal efficiency that makes the combined cycle so attractive in the first place. The high temperature requirements in the combined cycle designs range between 400-1000°C which is unsuitable for the conventional fabrics used in baghouse filtration. This has encouraged a number of different filter research programs of which granular bed filtration is one of the more promising options (23). Granular media can tolerate high temperatures but high collection efficiencies have yet to be effectively demonstrated under such high temperature/high pressure constraints.

Reed (22) describes a combined cycle plant design proposed by British Coal. Both a fluid bed gasifier and fluid bed combustor generate high pressure combustion gases used to drive a gas turbine. The design is anticipated to have high overall efficiencies while keeping sulphur emissions low by exploiting the fluid bed ability to control sulphur oxides by sorbent retention. Gas at temperatures of 1000°C, leave the gasifier with dust concentrations too high for the combustion gas turbine specifications. Thus, to retain high efficiencies, gas cleanup has to be carried out without cooling. At expected turbine inlet temperatures of 1300°C Reed quotes desirable inlet gas specifications as shown in table 1 below:

TABLE 1: EXPANSION GAS QUALITY FOR GAS TURBINES (REED) (22)

Particle size (microns)	None $>$ 10 99% $\leq$ 6
Particle loading (mg/Nm^3)	4

British Coal have concentrated their research into ceramic fibre filters as they believe this to be the only system capable of meeting the specifications. Reed reports ESP to be unsuitable because of the very low resistivity of the gasifier dust material. Normal ESP operation depends on resistivities of 10^9 Ohm-cm to 10^{10} Ohm-cm. The British Coal system produces a dust with resistivity in the order of 10 Ohm-cm which would require uneconomically large plate areas to overcome reentrainment. Cyclone cleanup is reported to be insufficient on its own but would be used as a precleaner to remove the bulk of the dust load.

Performance of granular panel beds is also reported by Reed (22), to be disappointing in capture of gasifier dust. The reasons are said to be due to electrostatic neutrality in the electrically conductive gasifier dust which is then unable to form a satisfactory filter cake necessary for high filtration efficiencies. What these effects would have on deep bed filtration are not reported. It is however noted that efficient filtration would require finer sand than practical restraints would allow but no figures for effective granule size were quoted.

The turbine specification adopted by British Coal is an extremely tight one. The requirements for different vendors quoted in the literature, vary from 3.3 to 200 ppmw total limit for solids entering in the inlet gases. An analysis by Stringer and Drenker (24), based on work done by General Electric, indicated that most of the damage done by a typical dusty combustion gas was done by small percentage of larger particles (20-50 micron). Stringer calculated damage contributions for each size fraction included in the typical dust. Particles of less than 3 micron were found to do no damage at all. Stringer and Drenker then suggested the following inlet specification based on a reasonable expander life span.

<u>TABLE 2: EXPANSION GAS QUALITY (STRINGER)(24)</u>

Not more than	.1 ppm	larger than 20 micron
Not more than	1 ppm	in the range 10-20 micron
Not more than	10 ppm	in the range 4-10 micron

Stringer reports that although this specification is on the borderline of cyclone capabilities, a more advanced form of cleanup will probably be required to ensure adequate expander life span. Roberts (25) on the contrary suggests that cyclones are capable of protecting turbine blades for a commercially-acceptable life span based on a 1000 hr evaluation trial.

Rubow (23) also reports unofficial turbine tolerances limits for different turbine manufacturers. The limit for GE turbines is similar to that proposed by Stringer but a total limit of 200 ppmw is included. Limits for Stal-Laval are much the same with a 200 ppmw total limit and 12 ppmw allowable for particulate greater than 5 micron. Westinghouse reports a total limit of 40-100 ppmw.

The wide range of specifications is a result of uncertainty due to lack of experience. So far no full size turbine has been operated downstream of a PFBC. (24)

2.5.4 DEVELOPMENTS IN FILTER DESIGN

Advantages of hot gas cleanup in combined cycle plant design affords a two percentage point increase in overall efficiency, over that attained with cold cleaning.(22) The improvement in efficiency represents enormous economic advantages for this design. As a result extensive research programs are currently underway to develop a filtration system suitable for particulate cleanup from coal combustor/gasifier systems.

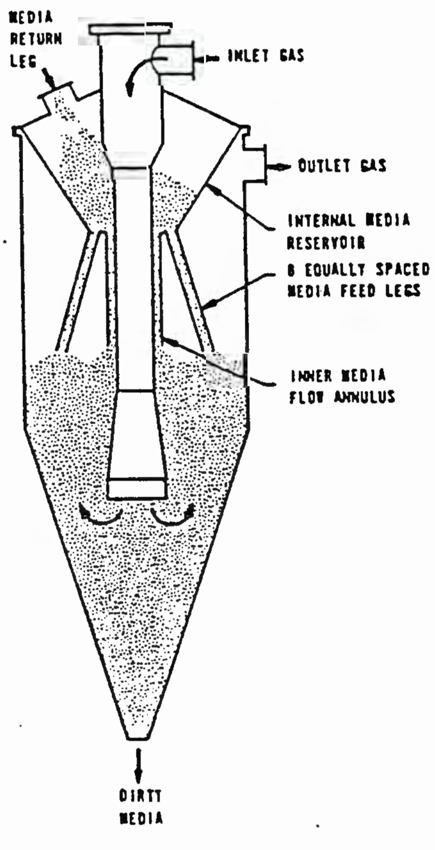
Rubow (23) reviewed an investigation done by Gilbert/Commonwealth for the US Department of Energy (DOE). The investigation covered a comprehensive evaluation of ten commercial filter designs for high temperature/high pressure service at different stages of development. The objectives were to establish a priority for each design based on performance, cost and general viability. Conceptually, the results of Rubow's analysis ranked specific commercial developments in the following order:

1. Electrostatic precipitator
2. Moving granular bed filter
3. Ceramic filter (rigid)
4. Woven ceramic filter
5. Acoustic aggolomeration

These techniques have been used extensively in lower temperature applications but problems relating to each, make their high temperature application uncertain. Nevertheless granular bed filtration is believed to be one of the most promising.

The moving granular bed filter described in the review by Rubow is of particular interest to this project. The design is one developed by the Combustion Power Company, the same group who developed the Electroscrubber, described earlier.

FIGURE 5: MOVING GRANULAR BED FILTER (CPC)



drawing taken from Rubow (23)

The filter bed is not configured in a panel but is designed to be screenless to avoid the problems of buildup on the louvres. The filter bed is housed in a tubular vessel with a conical base. The media is introduced to the top of the vessel and removed along with the trapped dust from the base. A number of media feed legs keep the vessel roughly half full of filter media. The dirty gas stream is then fed into the bed via a central downpipe which is emersed below the surface of the filter media. The gas stream is forced out within the filter media, concurrent to the granular flow. A diagram of the filter unit is shown below. The unit has been developed for high pressure and temperature applications and prolonged test runs are reported to have achieved overall efficiencies of 99%. The filter does not have an electrostatic field as the earlier Electroscrubber design.(14)

2.6 SUMMARY OF FILTER DESIGN PARAMETERS:

A summary of the filter design parameters extracted from the literature are shown below:

TABLE 3: SUMMARY OF GRANULAR FILTER DESIGN PARAMETERS

FILTER	TYPE fluid flow solid flow	GRANULE SIZE mm	FACE VEL m/sec	PRESSURE DROP kPa	BED DEPTH m	INLET LOADING g/m3n	EFFICIENCY %	REFERENCE
HORIZONTAL	downflow static	2-3	0.5-0.7	*	0.12	14	99.0-99.7	4
Fixed Granular Bed Filter	static	.3-75	.025	2.8	up to 2.7	very low	99.98	5
SQUIRES PANEL BED FILTER	crossflow static	0.3-0.4	.11	1.5	0.09	8	99.7-99.9	7,36
GRANULATE-TUBE-FILTER (LURGI)	crossflow static	fine	*	*	*	2	99.5	8
DORFAN IMPINGIO	crossflow moving	13-38	1.8	*	0.3	14	96.7	5,7
ELECTROSCUBBER	crossflow moving	6-9	0.4-0.7	2.0	0.75	1.5	93	9,11,12,14
SCREENLESS GB FILTER (CPC)	upflow moving	2	*	6-7	6	*	99	12,23

CHAPTER 3

3.0 GRANULAR FILTER THEORY

The design of an industrial filter has to be based on some predictive calculations, obtained, in some cases, from pilot or bench scale tests or from a theoretical model or, in most cases, a hybrid of the two. To reduce the dependency on physical testing, correlations and models have been developed to predict filter performance. A clean filter bed design as proposed by this current work would need similar analysis to establish design tools. Intuitively, moving granular bed filtration characteristics should be similar to that of the fixed bed filter models which are reviewed below. It is necessary to follow the development of these models to establish whether any modifications to suit moving bed conditions are necessary.

Filtration model development requires a fundamental understanding of the filtration processes involved and quantitative evaluation of the influence of each process. Granular bed filtration is by nature a transient process where the collection efficiency and pressure drop vary with time.(26) This variation is a function of the accumulation of captured particulate, and hence changes in collection efficiency occur during the course of filtration itself.

For aerosol collection in granular beds, it is known that collection efficiencies improve over the course of filtration but that the varying deposit morphology makes such deposition very difficult to model. As a result, most attempts to model granular bed filtration assume that the granules are clean and free of any already deposited material. Theoretical models and empirical correlations for granular bed collection efficiency are thus only applicable to initial stages of the filtration cycle when the granules are still clean. Because of the lower collection efficiency during the initial stages, the initial collection efficiency is often used as a conservative estimate in fixed bed design calculations. A theoretical model or empirical correlation of the initial collection efficiency with the pertinent filtration parameters is thus of great value to the design of an industrial filter. If the initial collection efficiency models for fixed bed filters apply to the moving bed case over the full range of operation, it should be possible to model the proposed clean bed filter design accurately doing away with the conservative design approach used in fixed bed filters.

This chapter reviews the fundamental processes occurring in granular bed filtration and development of models used to predict their effects on filtration efficiency. A purely theoretical treatment has, to date, been unable to predict the filtration characteristics fully and hence a number of semi empirical correlations have been developed. Most of the published correlations on granular filtration relate to the initial collection efficiency.

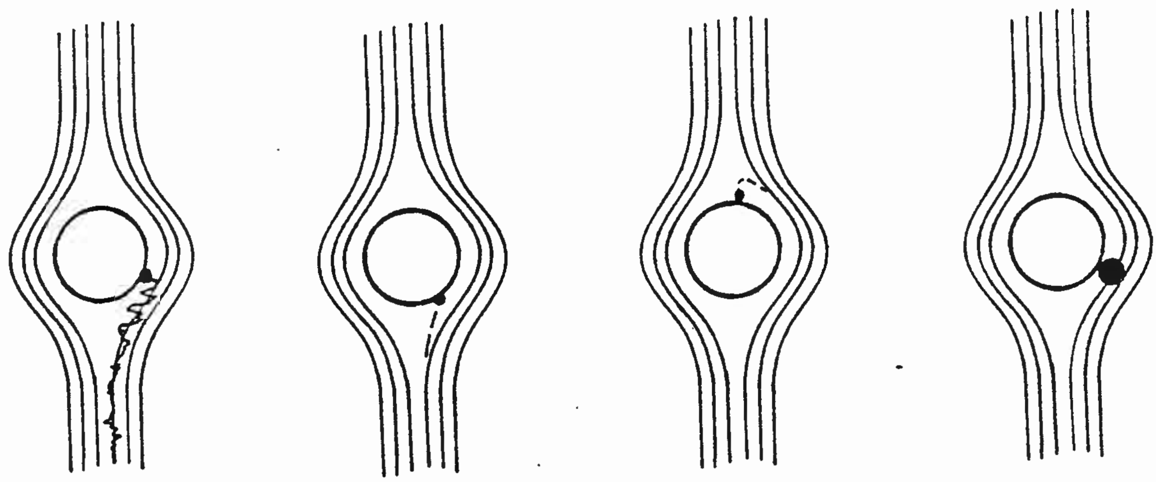
3.1 SEMI EMIPIRICAL CORRELATIONS

The majority of correlations reported have been derived on the basis of a number of filter mechanisms which collectively contribute to the overall filter ability. In the previous chapter the mechanisms were lumped under the label of "depth filtration", analysis of which identifies four mechanical processes involved during the passage of the fluid and particulate through the granular interstices. The mechanisms have been described by Clift et al (6) as the following:

1) Diffusion
2) Inertia
3) Gravity
4) Interception

A schematic representation of each mechanism is shown below:

FIGURE 6: FILTER MECHANISMS



Diffusion Inertia Gravity Interception

Two distinct processes are involved in granular filtration, namely collection and retention. The mechanisms described here are all part of the "collection" process relating to particle/granule contact. Having established contact, "retention" refers to the processes which retain the particulate on the collector and prevent re-entrainment. The efficiency of both these processes has to be high for effective filter performance.

In the preceding chapter it was noted that particles following streamlines could almost avoid any granular contact and thereby avoid capture. With the exception of interception, the above mentioned capture mechanisms relate to influences that cause the air-borne particulate to deviate from these fluid streamlines. Such deviation improves the probability of particle/granule contact and thereby increases particle collection efficiency.

Intuitively, the relatively slow movement of the granular collectors, with respect to the fluid velocity through the bed, should not affect the collection mechanisms unless the granular packing is altered. If bulk movement of the granules can be achieved there should be no variation between moving and stationary granular filtration characteristics because the flow field around the granules should not be altered significantly. However if intergranular movement is introduced in the moving bed case some variations can be expected. Intergranular movement could alter bed voidage where collector granules are not perfectly spherical. This in turn would alter the flow paths of the gas stream through the bed and thereby affect the significance of the various collection mechanisms. The effect of intergranular movement on the retention mechanisms for small particles should not be affected as the fraction of collector surface in contact with adjacent collectors is very small. Thus it is surmised that the dislodging effects from intergranular friction should not be significant especially during initial filtration when the filter bed is still clean. This however would not hold for larger particulate or particle agglomerates. Dislodging of the larger particles would not be independent of intergranular movement as retention mechanisms of the larger particles wedged in crevices would be affected by intergranular movement. Documentation on retention mechanisms is not as extensive as that covering the collection mechanisms.

Diffusional effects are a result of what is usually called Brownian motion. Brownian motion describes the erratic movement of a very small particle (<2 micron) resulting from successive random collisions with gas molecules. The effect is restricted to very small particle sizes because the higher inertia of larger particles is significantly greater than the inertia of the gas molecules. Erratic movement of the particle uses deviations from the fluid streamlines thereby increasing particle to granule contact. The effect is described by the Stokes-Einstein diffusion coefficient B′ :

$$B^1 = \frac{RT}{3\pi\mu \, d_p C_s} \qquad (1)$$

By dividing B′ by the particle diameter, d_g, the mean velocity caused by Brownian motion is obtained. The diffusion effect is expressed by the ratio of Brownian velocity (B'/d_g) to flow velocity, and is given by the inverse of the Peclet Number.

$$\frac{1}{P_e} = \frac{B}{d_g V_s} = \frac{RT}{3\pi\mu\, d_g V_s d_p} C_s \qquad (2)$$

Temperature has a strong influence on Brownian motion. The influence of pressure is reported to be much weaker and only becomes perceptible for particles smaller than 2 micron at high temperatures.

The inertial capture mechanism is associated with larger particle sizes (<10 micron) (1) where the inertia of the particle introduces some deviation from the streamlines and hence increases the chances of contact with the surface of the collector. Inertial effects are usually correlated in terms of the dimensionless Stokes Number given as:

$$N_{st} = \frac{d_p^2 V_s \rho_p C_s}{9\mu d_g} \qquad (3)$$

The Stokes No. represents the ratio of the "stopping distance" of a particle by viscous forces of the fluid to the diameter of the collector granule. (6) The Stokes No is sensitive to temperature because of the combined effect of temperature on gas viscosity and density but relatively insensitive to pressure which only affects the gas density.

Gravity effects on a particle can also cause deviation from fluid streamlines and induce contact with the collector surface. The effect of gravity is normally expressed as the ratio of the terminal settling velocity of the particle to the superficial velocity through the filter bed:

$$N_G = \frac{U_t}{V_s} \qquad (4)$$

The terminal settling velocity is derived from Stokes' law as:

$$U_t = \frac{C_s\, g\, d_p^2\, (\rho_p - \rho_g)}{18\mu} \tag{5}$$

The gravitational parameter thus becomes:

$$N_G = \frac{C_s\, g\, d_p^2\, (\rho_p - \rho_g)}{18\mu\, V_s}. \tag{6}$$

Once again the effect of temperature on the gravitational parameter through the gas viscosity and density terms is more significant than the effect of pressure.

The fourth mechanism, direct interception, does not describe particular forces which cause deviations from streamlines, but rather takes into account the finite size of the particle and its flow path. A particle would make contact with the collector surface if it comes within half a particle diameter of the surface. The collection mechanism is usually correlated by the "interception parameter" which is the ratio of the particle and granule diameters:

$$N_I = \frac{d_p}{d_g} \tag{7}$$

In addition to the four mechanical mechanisms described above there are other electrophoretic, diffusiophoretic and thermophoretic effects found to influence filtration efficiency. These effects are described by Clift et al (6) and also by Ives (1). They are by no means considered unimportant but the analysis of their influence is complex and considered beyond the scope of the project.

All four mechanisms operate simultaneously on the particulate travelling through the filter bed, each providing a contribution to the overall collection efficiency of the filter. The total efficiency is then usually calculated by summing the efficiencies of the individual mechanisms:

$$E = E_D + E_I + E_G + E_{DI} \qquad (8)$$

where the suffixes refer to diffusion, inertia, gravity and direct interception. Most models do not consider all the mechanisms. They are usually tailored to make one or two of the mechanisms dominant to highlight the effect of that particular mechanism. Correlations are thus developed by identifying the influence of each mechanism on the filter efficiency over a range of conditions.

Intuitively, moving granule conditions would not affect any of the parameters if granular movement was kept very small in relation to gas and particulate flow. As a first approximation intergranular movement will be assumed to be negligable, with one exception. Intergranular movement is expected to have an effect on the bed voidage by altering the packing density of the granular charge. Several empirical correlations published in the literature include voidage in their model which should allow for the moving bed case by appropriate substitution of the moving bed voidage. The number of different correlations published is quite extensive and for the sake of brevity three of the most recent and comprehensive have been selected for investigation. The correlations are based on a combination of theoretical work using particle trajectory analysis as described below.

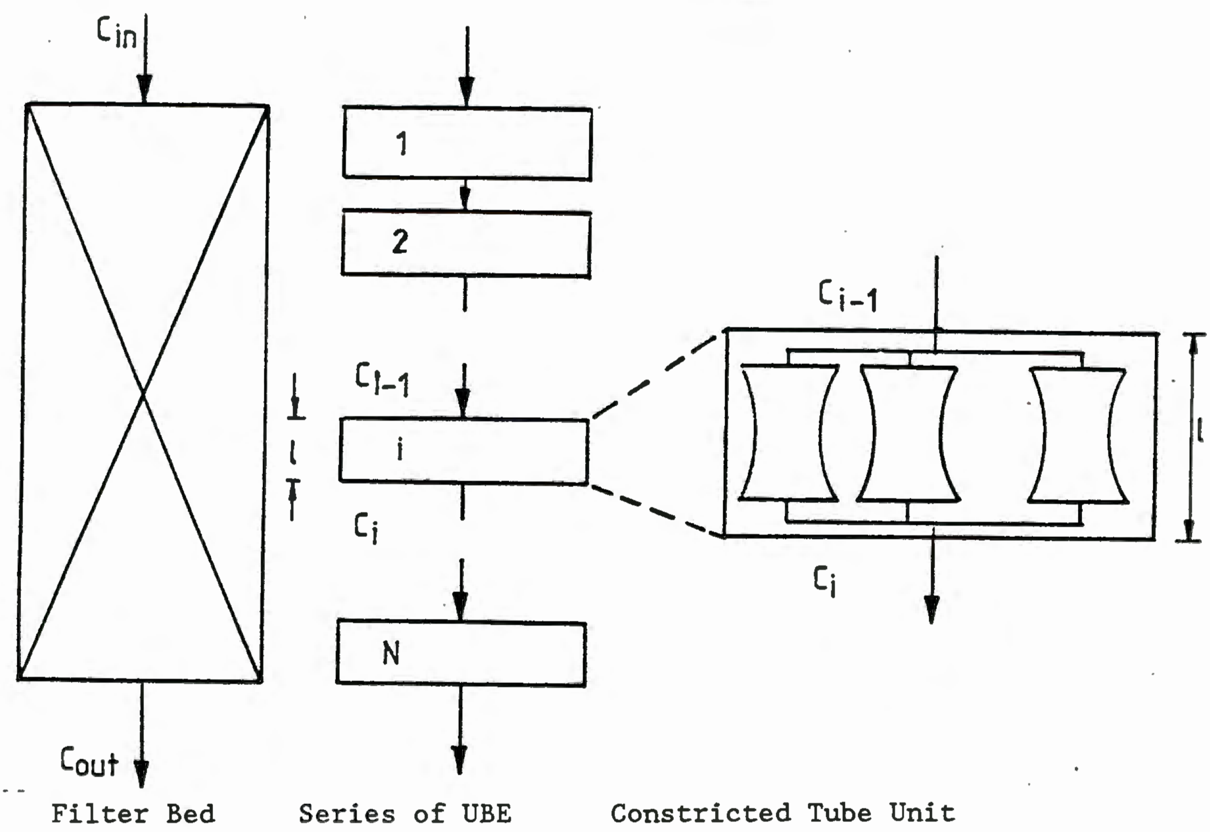
3.2 THEORETICAL TREATMENT

The theoretical treatment is based on developing equations to describe the particle flow field through the granules. Knowledge of particle trajectories allows prediction of any contact with the collector surface and hence determination of the theoretical collection efficiency. The actual flow path through the bed of granules is usually so complicated that simplified models have to be used. The models developed, can be classified as either internal flow models, where the fluid is assumed to flow through interstices in the granular bed, or external flow models, where the fluid is assumed to flow around the granules. The complexity of the theoretical models does not permit analytical solution and invariably the solution has to be found numerically. Several authors have used theoretical models to assist in the development of semi empirical models, two of which are described below.

3.2.1 INTERNAL FLOW MODEL

Pendse and Tien (26) consider the granular bed to be composed of a number of unit bed elements (UBE) in series, each of which comprises a number of geometrically similar constricted tube collectors (27) (unit cells) as shown below:

FIGURE 7: SCHEMATIC REPRESENTATION OF A GRANULAR BED



The overall filter efficiency is determined by the collective efficiency of each UBE. As each UBE is made up of a number of unit cells operating in parallel it is sufficient to calculate the unit cell efficiency to determine the UBE efficiency. This draws attention to the unit cell efficiency, as knowing this would allow the overall bed efficiency to be calculated. The collection efficiency of a unit cell is modelled by setting up equations to describe the particle trajectory through the constricted tube geometry (27). The model, known as the constricted-tube model, uses the geometry of the unit cell walls to describe the interstitial collector surface created by two adjacent filter granules. Fluid is assumed to flow through the unit cell and particle capture to occur with contact on the cell boundaries.

The development of the theoretical model starts with the general equation for particle motion which describes the dynamic equilibrium of a number of forces experienced by the particle as it flows along a fluid streamline through the granular medium. The forces considered are particle inertia and fluid drag.

$$mp \frac{d^2 x}{d_t^2} = \frac{3\pi\mu}{C_s} \cdot d_p \left(V_s - \frac{dx}{dt}\right) - F_{ext} \qquad (9)$$

As with purely empirical solutions, various external forces aswell as the contribution of interception need to be included in the equation for particle motion. The gravitational and diffusional contributions being external influences can be introduced through the F_{ext} term. The interception effect, taking into account the finite size of the particle, is introduced through boundary conditions. The equation is then rewritten in dimensionless form incorporating the relevant dimensionless groups mentioned previously. The solution to these equations establish the trajectories for each particle size. Of these, the limiting trajectory, the trajectory that just misses the collector, is found and its distance b, from the symmetrical axis, recorded. Assuming the particles entering the cell at distance greater than b from the symmetrical axis touch the collector and are caught, the cell collection efficiency for each particle size can be calculated. Overall filtration efficiencies are then readily calculated.

Pendse and Tien (26) describe a theoretical model where inertia dominates particle capture. Numerical solution of the equations of particle motion demonstrated that a combination of parabolic and sinusoidal geometries, for the shape of the cell wall, produced better correlation between theory and experimental data than each of the individual geometries on their own. Correlation with experimental results also showed that fluid inertia was important to the interception parameter and an empirical correction was introduced to account for this. The final form of the semi empirical equation is:

$$\eta = B\left[N_{st} + 0.48\,(4 - \frac{4N_I}{d_c^*} + \frac{N_I^2}{d_c^{*2}})^{\frac{1}{2}} \cdot \frac{N_I^{1.041}}{d_c^*}\right] \qquad (10)$$

where B is the correlation introduced to account for the fluid inertia. It is defined as the ratio of unit collector efficiency predicted by the model to the value calculated under creeping flow conditions. The exact form of B is:

$$B = 1 + 0.04N_{Re} \qquad (11)$$

However Yoshida and Tien (28) report that Eq. 10 overestimates the collection efficiency and proposed an alternative expression for B. The correlation is also the ratio of the unit collector efficiency calculated by trajectory analysis and the value calculated under creeping flow conditions. The exact form of B is:

$$B = 7 - 6\exp(-0.0065N_{Re}) \qquad (12)$$

The relationship between the granular medium and the unit bed element is given by the following equation.

$$l = d_g\left[\frac{\pi}{6}\frac{1}{(1-\epsilon)}\right]^{1/3} \qquad (13)$$

With a bed height, L, made up of unit cells of length l, the overall collection efficiency, E, becomes:

$$E = 1 - \frac{C_{out}}{C_{in}} = 1 - \frac{C_1}{C_{in}}\frac{C_2}{C_1} \;\cdots\cdots\; \frac{C_{out}}{C_{N-1}} \tag{14}$$

$$= 1 - (1-\eta_1)\,(1-\eta_2)\;\cdots\cdots\;(1-\eta_N)$$

For initial conditions in the bed $\eta_1 = \eta_2 = \cdots\cdots = \eta_0$. Combining this with the above equations the following represents the equation for the initial unit cell collection efficiency:

$$\eta_0 = 1 - (1-E)^{l/L} \tag{15}$$

Rewriting the equation as shown below provides the basis for the calculation of the unit cell collection efficiency from experimental data.

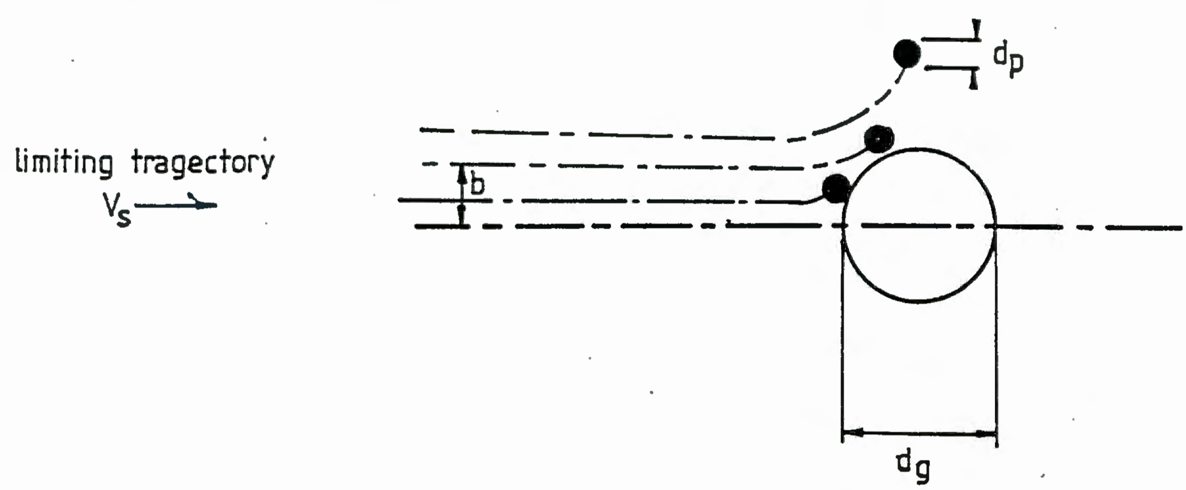
$$\eta_0 = 1 - \left(\frac{C_{out}}{C_{in}}\right)^{l/L} \tag{16}$$

Strictly speaking, the Pendse and Tien correlation should be modified for the moving bed case. The constriction diameter in Eq 10 would be affected by granular packing arrangement but as only an approximate value for this variable is cited in the literature it is assumed that the same diameter would hold for the moving bed (26). The effect of voidage variation in the moving bed would be accounted for in the estimation of UBE length, l in Eq 13. This model will be used in an attempt to describe the action of the moving granular bed.

3.2.2 EXTERNAL FLOW MODEL

The external model (29) also assumes the filter bed to be subdivided into unit cell collectors. The unit cell collection efficiency is then integrated over the bed depth to calculate the overall collection efficiency. The difference between the internal and external unit cell models arises out of the different definition of the unit cell collector. The unit cell, used in the external flow model, is a single sphere around which particle trajectories can be calculated. For the purposes of the analysis of the single cell collection efficiency the cell is in most cases assumed to be a spherical granule with a free surface (no sphere is touching another) located in the centre of a cell of flowing fluid, as shown below:

FIGURE 8: SCHEMATIC REPRESENTATION OF A SINGLE CELL COLLECTOR



The radius of the cell is calculated so that the void fraction is equal to the void fraction in the granular bed. Equations describing particle trajectories around one filter element within the cell, are developed and solved numerically, to determine the limits of the flow field within which a suspended particle will be caught. Establishing the limiting trajectory allows the theoretical collection effciency to be calculated.

The overall efficiency is calculated using the single cell efficiency by the following:

$$E = 1-\exp(-k_1(1-\varepsilon)(L/d_g)\eta) \tag{17}$$

where K_1 is a constant. Several values for K_1 are given in a review by Gal et al (30) of which, 1.5 seems to be the most commonly used. The equation can be rewritten to allow single cell efficiencies to be calculated from experimental data :

$$\eta = \ln(C_{out}/C_{in})/-k_1(1-\varepsilon)(L/d_g) \tag{18}$$

Several different flow fields around the collector have been developed.(29) The fluid flow boundary conditions depend on the effect of one collector on another and are hence based on the assumptions of the location of one collector relative to others. The simplest of these assumptions considers the cell to be in a fluid which extends to infinity. The model does not fit the packed bed filter but may sometimes fit a fluidised bed filter. Further it relates to a filter bed with high porosity and is often used to test more complicated solutions which should converge to flow around a sphere in an infinite fluid when porosity approaches unity.

Tardos et al (29) cite several other well known approximations to the solutions for flow around a sphere. The interaction of collectors on each other is accounted for, in most of the later models, by considering the porosity of the filter medium. The theoretical models are difficult to validate without going through the whole numerical solution using one's own data. However the semi empirical models developed from the theoretical exercises can be used far more easily.

Gal et al (30) have published a semi empirical correlation based on theoretical trajectory calculations. The model considers inertial effects to be dominant and assumes capture of particulate to occur if the particle trajectory comes within a distance equivalent to the particle radius of the collector surface. The theoretical model is based on an external flow where the collectors are viewed to touch one another in contrast to the case where a collector is viewed in an infinite fluid. A three dimensional flow model was used instead of the two dimensional flow field assumed in most other models.

The trajectory calculations are far more complicated in three dimensions but the collection efficiencies are still based on establishing limiting trajectories. The collection efficiencies predicted by this model are much higher than the sphere-in-cell models because the effect of the contact points between collectors is taken into account. The semi empirical model developed by linear regression of results from the trajectory analysis is shown below:

$$\eta = 2 \cdot N_{st}'^{\,3.9} \Big/ \left[4.3 \times 10^{-6} + N_{st}'^{\,3.9}\right] \qquad (19)$$

$$0,1 > St' > 0,03$$

where St′ is a modified Stokes No allowing for the hydrodynamic effect of the fluid. The modified Stokes No is found by the following:

$$N_{st}' = N_{st}\left[1.0 + 1.75\, N_{Re} \,/\, 150\,(1-\varepsilon)\right] \qquad (20)$$

where Re is the Reynolds Number representing the hydrodynamic properties of the fluid.

Comparison of predicted and experimental results is covered in section 5.3. Additional theoretical development is included along with the results because the results highlight the relevance of the modifications to the theory.

CHAPTER 4

EXPERIMENTAL EQUIPMENT

Earlier work by Dutkiewicz (10) reported the capability of a vertical granular filter bed in the recovery of valuable product material dusts in the metallurgical field. The effects of granular movement necessary for filter medium recirculation were however not discussed. In this current project the aim has been to establish the ability of the same filter design to trap fly ash derived from coal combustion with continuous regeneration of the filter medium. A similar panel configuration to that used by Dutkiewicz was used in all the experimentation. Dutkiewicz however did not identify whether his results were representative of clean granule conditions or not. Nor did his work assess the ability of the filter panel for the capture of coal combustion fly ash. This project can be considered an extension of his work in so far as the same panel configuration was used but the objective was to use the regeneration control as a means of reducing pressure drop, while maintaining desirable filtration efficiencies.

Numerous experiments (26,28,29,30) have been conducted to assess the initial collection efficiency of granular filter beds and to develop models for this condition. The initial collection efficiency is considered the least effective condition in static granular bed filtration and the models are developed solely to provide a conservative measure of filtration efficiency. However, the clean granule efficiency, does illustrate that clean granules have a significant filtration ability.(26,28,29 30) This filtration ability is considered inferior to that achievable with a filter cake but it does have one advantage over the latter, namely low operating fluid pressure drop and the ability to predict consistent behaviour. Low pressure drop is not always of major concern but in large filter installations typically used in the coal fired power generation industry, fan power costs, a function of pressure drop, constitute a major operating cost. Reduction of operating pressure drop without affecting overall collection efficiencies would represent an attractive economic advantage for granular filter operation in general.

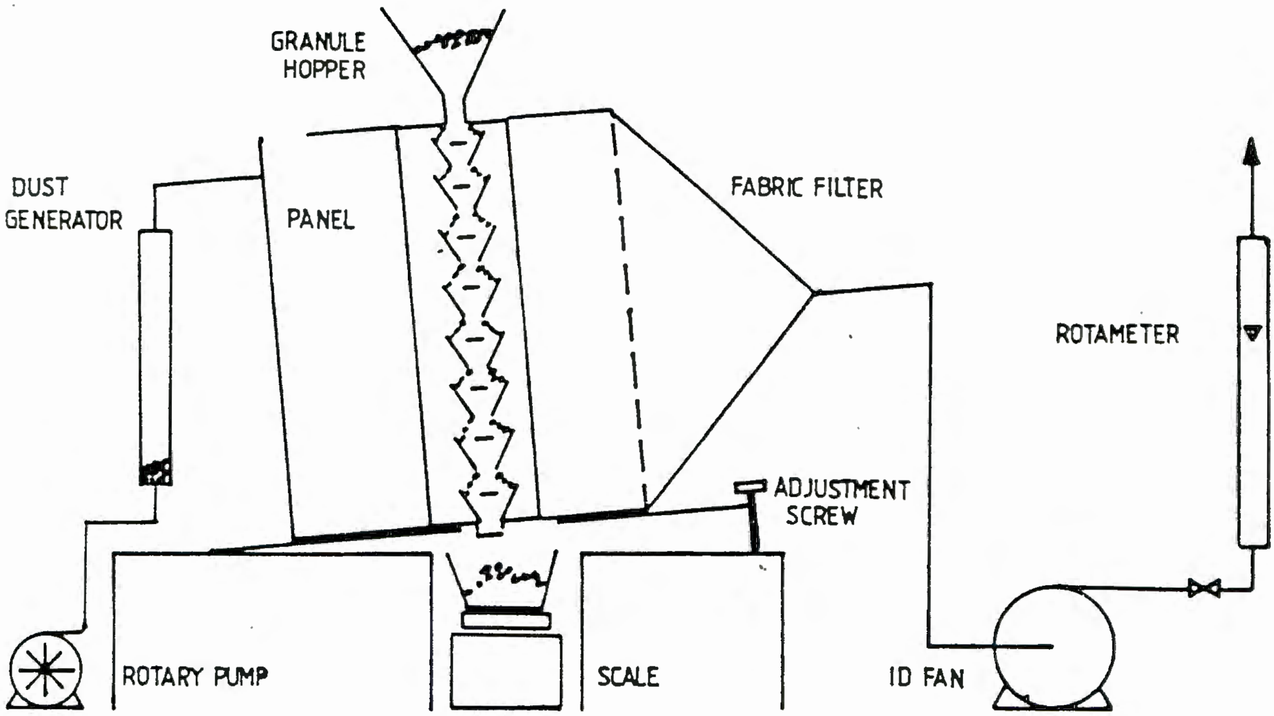
The moving granular bed has the ability to control pressure drop by varying the trapped dust concentration in the filter bed. Increases in granular circulation flow reduce the dust accumulation within the bed and thereby generate less resistance to fluid flow through the filter bed. However the granular movement may affect the filtration efficiency. Part of this experiment is thus to assess the effects of granular movement on granular filtration efficiency.

4.1 PERSPEX MODEL

As some problems in solid medium flow control were anticipated, it was necessary to visually observe granular flow patterns to optimise the panel design. It was decided to conduct these observations at room temperature to enable transparent perspex material to be used. Emphasis, was placed on regeneration of the granular surfaces and reduction of stagnant areas within the granular flow. The low temperature experimentation was important in development of a working model but the filter could only really be tested under more realistic conditions. Consequently, a second experimentation program was conducted at higher temperatures using a fluidised bed combustion flue gas stream from a small experimental rig described later.

The apparatus for the preliminary experimental work was constructed in clear perspex to enable visual observation of the granular flow. The filter panel itself, also built in clear perspex, had a face area measuring 150x400 mm. The louvre and baffle arrangement used was to scale which presented a filter thickness of 4-5 cms to the gas stream flowing through. The small cross-sectional area of filter panel presented to the dust flow was thus considered to represent a small section of an industrial panel module. The filter section was built into a perspex manifold with an outlet connection to an I.D. fan and flow meter. The apparatus is shown below.

FIGURE 9: LOW TEMPERATURE EXPERIMENTAL RIG



The I D fan was used to draw air through the test filter section while a valve on the discharge controlled the ultimate velocity of the air flow being drawn through the panel. The superficial gas velocity through the filter was established by reading the total air flow from the flow meter connected to the blower discharge and dividing that by the face area of the filter. A hopper was fitted directly above the filter panel to house the sand filter medium.

Granular medium flow through the panel was controlled by adjusting the aperture produced by a hinged flap at the base of the panel. Sand (740 micron) pouring freely out of the flap was caught in a container placed on top of an electronic balance allowing incremental mass discharge rates to be measured. Pressure drop across the filter element was measured with a manometer connected to tappings in the upstream and downstream manifolds.

The apparatus mentioned above was used to study granular medium flow patterns and the effect of fluid flow, louvre orientation and flowrate on flow patterns. Pressure drop characteristics of stationary and moving granular medium were also investigated.

A dust generator was connected to the upstream manifold. This comprised a glass tube connected to a small rotary vane pump. The tube was blocked at both ends with piping connecting the pump, the tube and the inlet manifold. A shallow layer of sand in the base of the glass tube was used to enhance the entrainment of metered dust samples into the fluidising air flow. Dust samples used were collected from an experimental FBC rig. The metered dust samples were added to the dust generator every 5 - 10 min. Once air borne, the dust particles were fed to the inlet manifold of the perspex filter rig. The dust was allowed to mix with the inlet air stream before being drawn through the panel bed. Effluent dust samples were collected by a fabric filter in the downstream manifold.

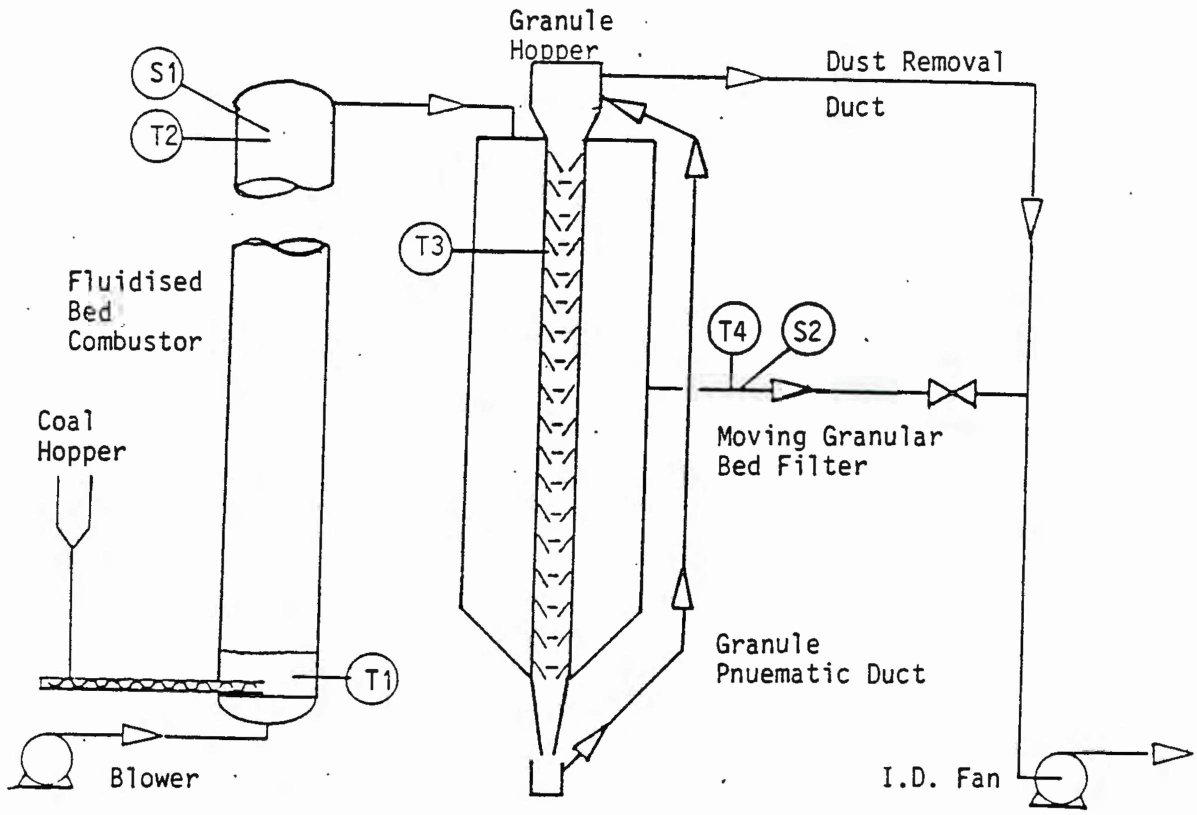
The collection efficiency was established by weighing the samples of fly ash added to the dust generator, the dust settling in the upstream manifold and the dust trapped by the fabric filter downstream of the granular bed. No attempt was made to weigh the dust emerging with the granules drawn from the base of the filter.

Initial tests showed solid flow patterns to be adversely influenced by the air stream flow through the panel. The crossflowing air stream slowed and eventually stopped the granular flow on the downstream face. A corrective measure was found to be accomplished by tilting the panel off the vertical. To accomplish this the entire perspex rig was mounted on a platform with an adjustable screw one end. The adjustable platform allowed the panel to be tilted to overcome granular flow problems as discussed in section 5.1.2. See Figure 9

4.2 HIGH TEMPERATURE APPARATUS

The intention of the high temperature work was to extend the work done at lower temperatures and produce more meaningful results by establishing collection efficiencies of a moving clean granular bed using hot flue gases. Most large scale power plants are based on burning the coal in pulverised form which produces large quantities of flyash characteristic of that combustion technique. Combined cycle technology incorporates fluid bed combustion/gasification for coal conversion which generates flyash with differing properties to that of the PF process. However the simpler construction and control of small scale FB combustors prompted the choice of FBC over PF as a flue gas generator. Difficulties in the production of a real fly ash from pulverised fuel combustion at laboratory scale were realised and thus not attempted. The results from the FBC filtration experiments will not strictly apply to PF flue gas cleanup but should suffice as a first order assessment. Ultimate proof of the filter's ability would have to be demonstrated using a flue gas slipstream from an industrial PF installation. A complete diagram of the experimental FBC and filtration equipment is shown in Figure 10.

<u>FIGURE 10: HIGH TEMPERATURE EXPERIMENTAL RIG</u>



The FBC used in this experiment was a 200 kW coal fired unit, 4500 mm high and diameter 296 mm. Bed depths were typically 200 - 600 mm, supported by a nozzle-type distributor. A blower connected to the base of the vessel below the support plate supplied air to fluidise the granular bed material. Firing of the coal required the use of an LPG preheating system to raise the bed temperatures above the auto-ignition temperature of the coal. Precrushed coal (< 10 mm) was then screw fed into the base of the sand bed. Five thermocouples installed in the bed, freeboard and flue ducting were used to monitor temperatures during operation. Oxygen concentration in the flue gas was monitored using an online Beckman 7003 analyser tied into a sample line from the ducting between the FBC and the filter. Coal feedrate was manually controlled to keep bed temperature and excess oxygen levels consistent with typical operating conditions found in industrial installations using fluid bed combustors. The FBC and filter housing were lagged with ceramic fibre insulation to reduce heat losses from the flue gas during its flow through the connecting ducting.

Dust concentrations produced by the fluid bed were measured with an isokinetic probe and sampling system installed in the top of the FBC vessel where dust samples were representative of elutriation.(31) Samples downstream of the filter (MGPF) were taken by a total stream fabric filter. A flap in the ducting enabled the fabric filter bag to be installed in the ducting during operation without undue complications, allowing more than one sample to be taken during each test run. The size distribution of the collected dust samples was analysed using a Malvern 2600/3600 Laser Diffraction Particle Sizer.

4.2.1 FILTER PANEL DESIGN

Initially a panel, 2 meters long by 0,5 meters wide, was built for the high temperature experiments. This size panel was envisaged to be representative of a full scale module, a number of which, collectively, would make up large areas required in industrial filter units. The louvre configuration, also full scale, was identical to that used in the preliminary tests. Silica sand of average size 0,74 mm (size analysis in Appendix E) was used as a filter medium throughout all the high temperature work.

The panel was constructed in four sections bolted on top of each other to make up one 2 m long section. After exploratory investigation it was found that it would be would be more appropriate to reduce face area than increase fluidisation velocity to achieve higher face velocities. As a result the face area was reduced to one quarter of the original size by use of suitable blanking plates.

Upstream and downstream manifolds were attached to the panel frame with hinges and bolts to allow quick and easy access to the panel during test work. The manifolds were encased in refractory panels to minimise heat losses. The upstream manifold was also fitted with a drain valve to allow settled particulate to be drawn out of the system. A settling rate was determined from the sample drained out of the upstream manifold. This figure was then used to correct the inlet particle loading predicted by the isokinetic sample. This allowed panel efficiencies to be distinguished from the total filter efficiencies.

Testwork consisted of running the FBC and filter simultaneously. Samples of particulate were collected from upstream and downstream ducting. Samples were weighed and analysed to provide data for collection efficiencies calculations. The filter medium sand was initially introduced to the hopper mounted directly above the panel. Sand flowrates through the panel section were varied by altering the removal rate through the rotary valve at the base of the panel but erratic flow characteristics of dust laden granules tended to inhibit accurate control. Below the rotary valve, a pneumatic transport line was used to convey both the dust and the granules up into the hopper above the filter panel. The transport line could be diverted and sand flow allowed to fall into a container to measure sand flow rates intermittently.

The transport air flow, as well as the filtered effluent disposal, were handled by a single I.D. fan. The suction developed by the fan was used to draw the granules back up to the hopper above the panel. The dust and granules were separated during transport and further by impingement on a baffle installed in the hopper. The dust laden air was then routed to the I.D. fan without cleaning. The experiment was designed to test the filter panel efficiency so no provision was made to clean up the transport air stream.

CHAPTER 5

RESULTS AND DISCUSSION

5.1 LOW TEMPERATURE EXPERIMENTS

Dutkiewicz (10) reported filtration results using lime ammonium nitrate dusts and silica sand filter medium in a similar panel design. The filter panel variant used in his experiment made use of a mesh to support the downstream face, not louvres as in this case. Long term blockage problems associated with a mesh support promoted the concept of louvres on the downstream face. Preliminary experiments with a louvred panel were conducted to compare the mesh and louvre support on the downstream face.

Dutkiewicz did not report granular flow rates for his experiments nor any difficulties with granular flow control. The high face velocities reported by Dutkiewicz could not be achieved without blowing the filter medium out of the panel setting a limitation on the fluid velocities for louvre support at the downstream face. Fluid flow through the panel was found to adversely affect the granular flow on the downstream side of the panel to an extent where flow was stopped. These difficulties prompted a detailed investigation of granular flow patterns through the louvre configuration and the effect of fluid flow. As already mentioned, a transparent perspex rig was built for the work. The same apparatus was used to observe the effect of accumulation of dust in the panel and the effect of granule recirculation on filtration efficiencies. Operational parameters could thus be established for the high temperature work where limited visual observations could be conducted. The following discussion relates to the results of the preliminary experimentation. The results of the high temperature experimentation are covered in Section 5.2.

5.1.1 LOUVRE ORIENTATION

Successful operation of this particular moving granular bed filter is based on continuous regeneration of the filter medium, thereby maintaining a relatively clean filter bed free of accumulation/caking on the granular surface or within its bulk. Ideally the flow of granules through the filter bed should have been free of any stagnant areas ie where granules were stationary. Stagnant regions were potential blockage areas which, by accumulation of particulate, would eventually prevent fluid flow through that particular zone. The simplest vertical panel design, free of any stagnant zones, would be one retained between two meshes but the potential blockage problems on the mesh surfaces preclude its use in many applications. Continuous renewal of the granular filter face was thus of equal importance to regeneration of the medium in the filter bed itself.

In this filter design the panel exposed a free, granular surface to the fluid stream. The filter was then designed to renew this free granular surface continuously to maintain a cake-free filter face. To enable free granular surface exposure, the panel was made up of a combination of angled louvres and horizontal baffles. The louvres retained the medium in a panel format exploiting the natural free angle of repose of the filter medium as shown in Figure 12. The inclined angle of the louvre could have been anything from 0 to 85 degrees to the horizontal but several conflicting requirements forced the use of the angle adopted.

Inclining the louvre had both advantages and disadvantages. Steeper angles ensured more uniform solid flow eliminating stagnant granular zones. However increasing the inclination of the louvres reduced the exposed free granular surface area and therefore the area for fluid flow at both the up and downstream filter faces. Reduced flow area generated a higher exit velocity on the downstream face according to the following relationship:

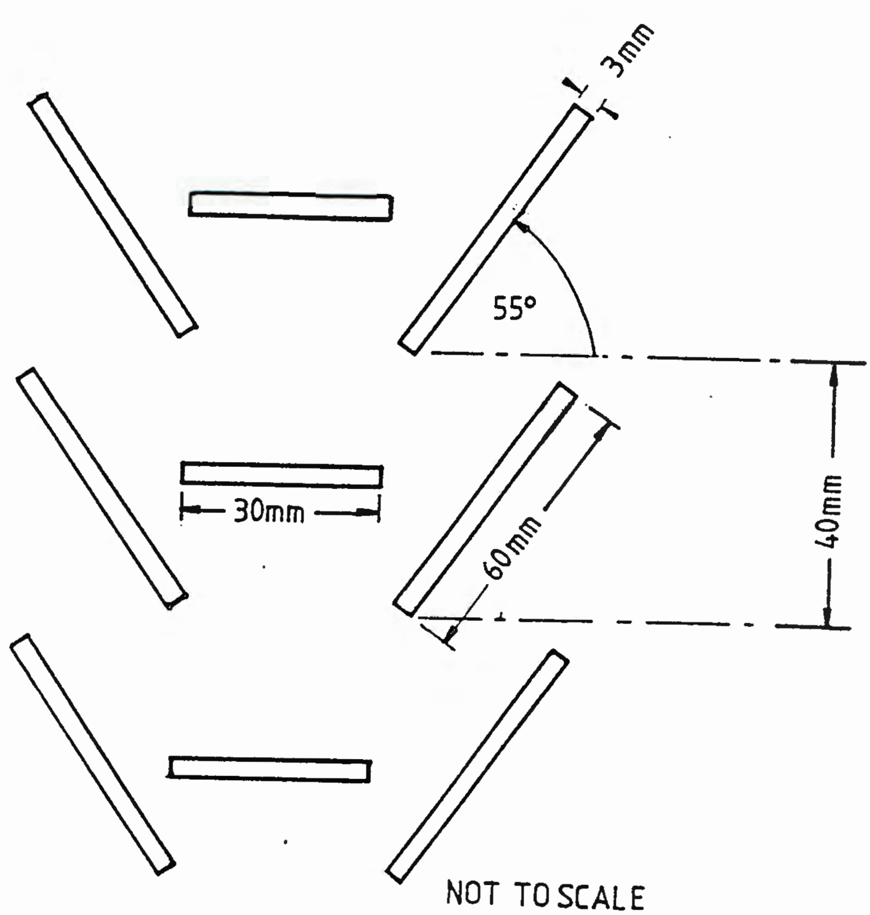
$$V_L = \frac{V_f}{\cos\alpha - L_R} \qquad (21)$$

$$\alpha < 90$$

$$L_R < \cos\alpha$$

where α is the louvre angle with the horizontal and L_R the fraction of the face area taken up by the louvre thickness. The higher exit velocity associated with steep louvre inclination was found to limit operating flowrates by causing premature failure of the granular surface on the downstream face. To allow higher fluid flowrates through the panel the louvre angle was reduced trading off good granular flow patterns against fluid flow limitations. The angle of repose for the sand used in this experiment was estimated to be 35°. Initial work done by Dutkiewicz using similar material made use of a series of louvres angled at 55° to the horizontal. (10) The same angles were used in this design and represent a compromise between solid flow requirements and limiting face velocity. The final louvre arrangement used to make up the panel is shown below:

FIGURE 11: LOUVRE CONFIGURATION

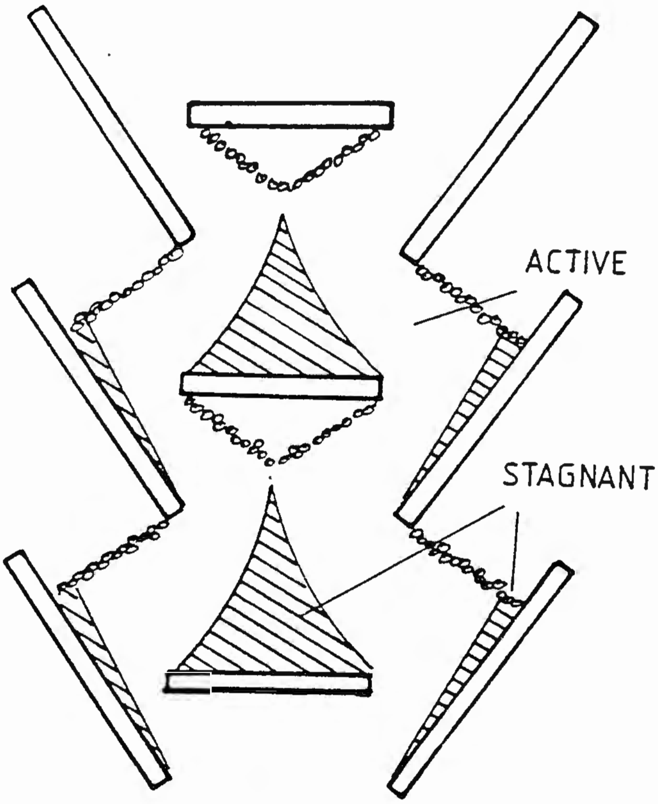


Without any external forces acting on the granular medium will flow much like a viscous liquid until the angle of the solid surface with the horizontal is equal to the natural angle of repose. The basis of filter face/surface renewal is the capacity of the granular medium to maintain its angle of repose automatically. To assist in the surface renewal, and to force granular flow through the louvre sections a horizontal baffle was inserted in the bed between the angle louvres. The obstruction to granular flow created by the baffle forced flow outwards toward the filter surfaces enhancing the surface renewal. In addition, the void space created below the baffle provided a shorter flow path for the fluid, and thereby reduced the variation in flow path length associated with the uneven filter surface created by the louvres. The filter bed developed by the louvres has a narrow section which would be the natural path chosen by the fluid flow being forced through. The central baffle, by created a void directly beneath its length, served as an alternative path for the flue gas passing through the thicker section of the panel, thereby distributing the fluid flow move evenly down the panel length.

5.1.2 VISUAL OBSERVATIONS

The perspex filter section allowed visual observation of the granular flow during filtration. Upstream filter face regeneration was demonstrated but stagnation of a thin layer of granules on the louvre surface was seen to persist along with other regions within the filter bed. The figure below identifies stagnant flow areas and active flow areas as seen from the observations.

FIGURE 12: STAGNANT AND ACTIVE FLOW AREAS

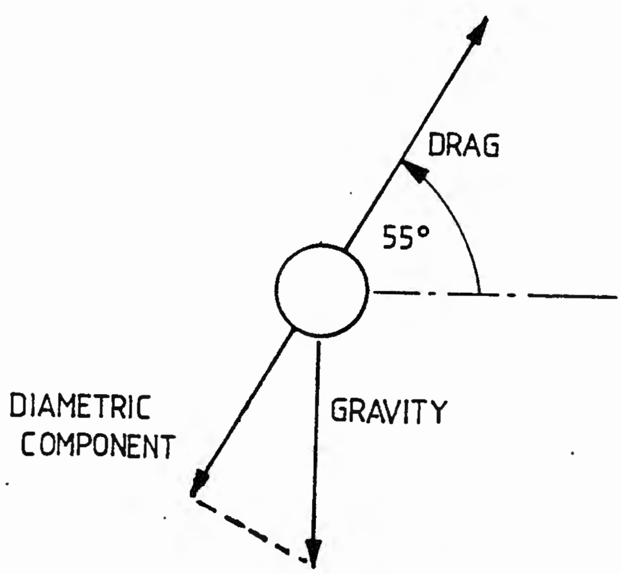


The horizontal baffle was found to generate a large stagnant section directly above it. However, without the baffle, the granular flow was found to channel down between the louvres leaving both upstream and downstream exposed granular surfaces stationary. The baffle and louvre arrangement developed two granular flow streams descending independently down either side of the baffle. The action of the granular flow around the baffle, outward toward the filter faces, assisted in maintaining filter face renewal. As expected, the stagnant region above the horizontal baffle was found to cake up with dust and eventually block. The effect did not increase pressure drop significantly because the corresponding void space created below each baffle provided an unrestricted flow path for the fluid.

5.1.3 PHYSICAL FAILURE

The maximum air flow through the filter panel was an important parameter to ascertain, as it put a limit on the minimum filter area permissible, thereby affecting the physical size and cost of the filter. The maximum face velocity is almost unlimited in designs where the filter bed is supported by a screen. The use of the louvred panel does not incorporate a support screen so that, in these experiments, high velocities would eventually remove material from the panel. The expected failure point was initially envisaged to be related to the minimum fluidisation velocity of the filter bed material ie the velocity at which the pressure drop across the granular bed is equal to the bed weight. An estimate of the expected limiting velocity was calculated using the minimum fluidising velocity equation from Wen and Yu as reported by Yates (33). At ambient conditions the minimum fluidisation velocity was calculated to be 0,37 m/sec. In practice the failure point was found to be at a face velocity of 0,16 m/s, which, because of the reduced air flow area created by the angled louvres, represents an exit velocity in the louvre section of 0,32 m/s. This fluid velocity in the louvres represents 80% of the minimum fluidisation velocity of the granular material. In classical fluidisation the pressure drop and gravity forces are vertically opposed but the angled louvres is this design would alter the orientation of these forces as shown below. It would therefore be incorrect to consider fluidisation as the cause for failure. In this case, failure would occur in the direction of the pressure drop when the component of the gravity, diametrically opposed to the pressure drop, was exceeded by the fluid flow pressure losses.

FIGURE 13: FORCES ON FILTER GRANULE AT THE POINT OF FAILURE



Taking into account the orientation of the force balance, the failure of the granular panel was calculated to be 0,31 m/s, which then matches the actual louvre exit velocity measured at failure point.

The failure point of the filter panel is analogous to the point at which minimum fluidisation occurs from the point of view of equating of opposing forces. Development of an appropriate failure point equation has thus been modelled along the lines of minimum fluidisation. The final equations for failure point shown below, are very similar to those for minimum fluidisation, but incorporate the angle of opposing forces to derive limiting face velocity in terms of the fluid properties, filter medium and louvre angle:

$$V_{Lmax} = \mu/d_p\rho_g.((33,7)^2+0,0408\ Ga')^{\frac{1}{2}}-33,7 \qquad (22)$$

where Ga′ represents the Galileo No modified for angled fluid flow. Ga′is modified as shown below:

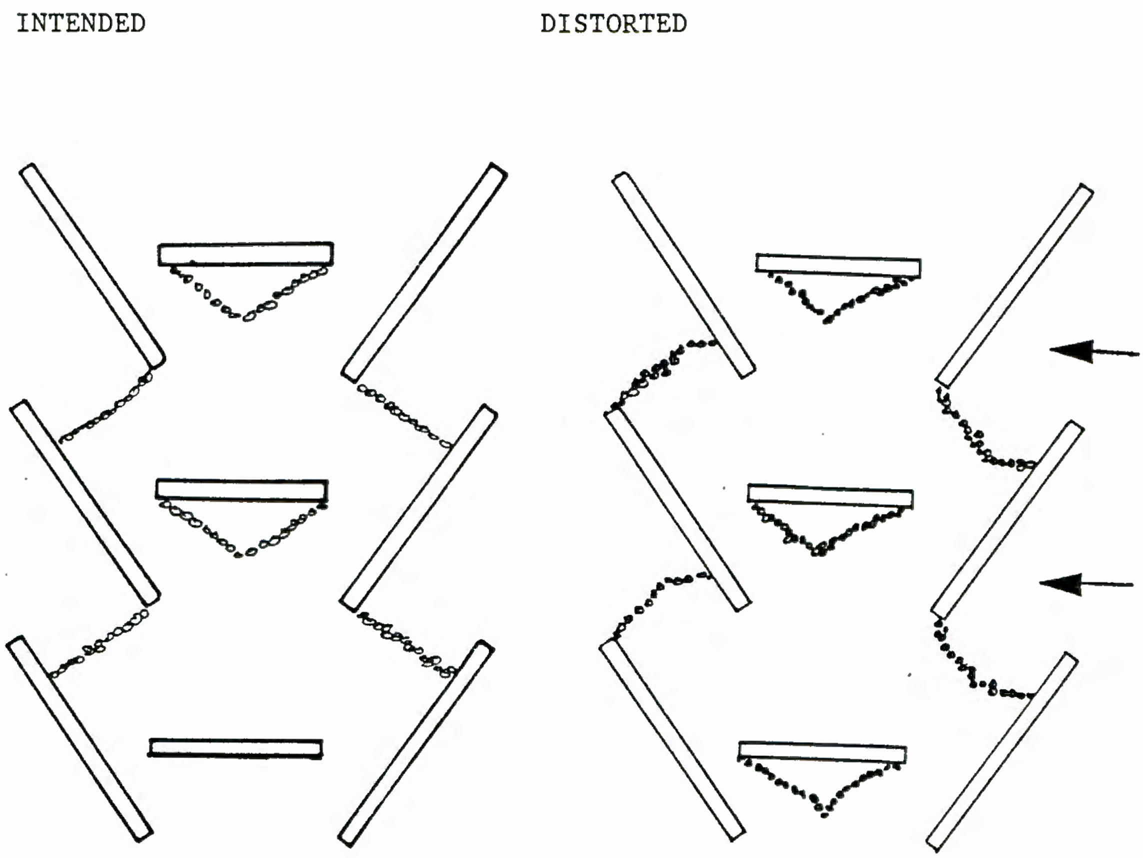
$$Ga' = \frac{d_p^3\ \rho_g(\rho_s-\rho_g)g\ \sin\alpha}{\mu^2} \qquad (23)$$

Louvre exit velocity is related to the limiting face velocity by the relationship:

$$Vmax = V_{Lmax}(\cos\alpha-L_R) \qquad (24)$$

The studies of solids movement with fluid flow through the bed identified further problems before actual failure. The flow of fluid through the granules distorted the intended cross section as per Fig 14.

FIGURE 14: PANEL CROSS-SECTION SHOWING DISTORTION

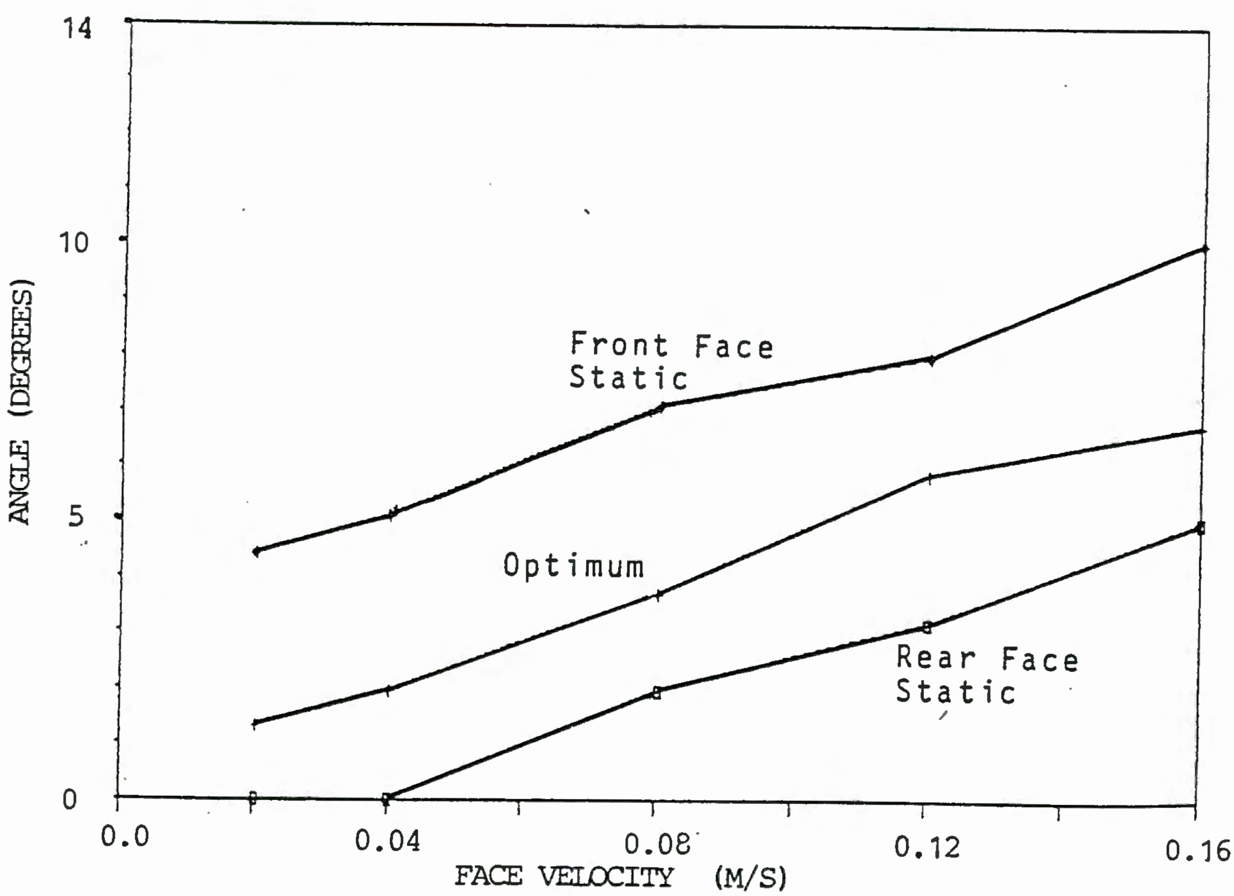


Some distortion was found to occur at velocities as low as 0,08 m/s. The associated stagnation of filter medium at the downstream face of the filter was inherently problematical for a moving granular filter design. The stagnant region was a potential blockage area which was later substantiated in filtration studies both hot and cold. Distortion of the filter cross section was found to begin at the point at which the downstream granular flow stopped. A more careful study of the solid flow patterns identified two distinct flow paths down each side of the filter panel. These two paths flowed almost independently down both faces of the filter bed without any fluid flow through the panel. As the fluid flow through the panel was increased, the two flow paths were found to flow at differing rates, with upstream flowing faster and downstream slower to a point where no flow occurred. The solid flow patterns were found to depend on face velocity with higher velocities aggravating the flow differential on the two filter faces.

5.1.4 INCLINED PANEL

By trial and error, it was found that inclination of the panel from a vertical position could be used to readjust the flowrates on the two filter faces during fluid flow through the bed. The granular flow patterns were shown to be a function of both fluid velocity and panel angle. In the vertical position the downstream flowpath would be brought to a standstill by increasing the face velocity to 0,04 m/s. By inclining the panel at this velocity the upstream and downstream flow paths could be induced to flow at similar rates. Further inclination off the vertical was found to reduce the upstream flow to a standstill. The variation in angle for the two extreme flow conditions was roughly 5 degrees making the operation very sensitive. The relationship between face velocity and off-vertical angle required is shown below with the optimum being when flowrates of the upstream and downstream faces being equivalent.

FIGURE 15: OFF-VERTICAL ANGLE REQUIRED FOR OPTIMUM FLOW



The complexity of these flow dynamics was not resolved. Increasing the louvre angles, in particular that of the downstream louvres, had no significant effect because the reduction in fluid flow area increased the fluid exit velocities increasing fluid drag. It is believed that the off vertical orientation of the panel allowed effective louvre angle correction (with respect to gravity forces) without increasing the exit velocities. From these findings it was necessary to incline the panel to avoid stagnation of the rear face and subsequent filter clogging.

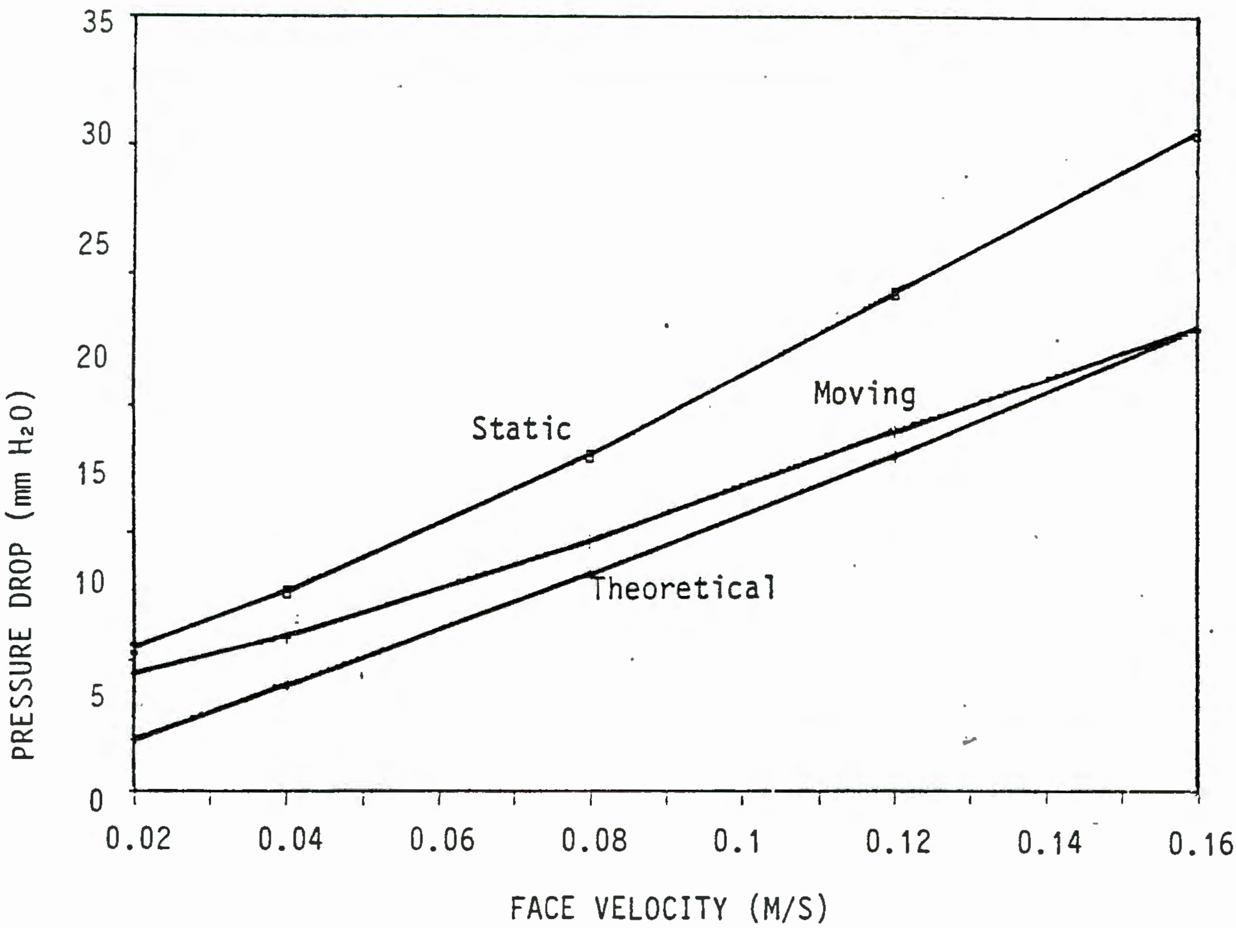
As mentioned earlier the exact mechanism of air/granule flow interaction was not investigated. The effect of a faster flowing upstream filter face on filter performance was also not investigated. This however, could be important as reference is made to improved efficiencies with higher upstream solid flowrates. (12)

5.1.5 PRESSURE DROP

Pressure drop measured across the filter thickness for various face velocities is shown below. Both static bed and moving bed conditions along with the theoretical pressure drop, are included. Theoretical pressure drop was calculated using the modified Ergun equation shown below:

$$\Delta P/L = 150(1-\varepsilon)^2 V_s \mu / d_g^2 \varepsilon^3 + 1,75(1-\varepsilon) V_s \rho_g / d_g \varepsilon^3 \qquad (25)$$

FIGURE 16: GRANULAR PANEL BED PRESSURE DROP RESULTS



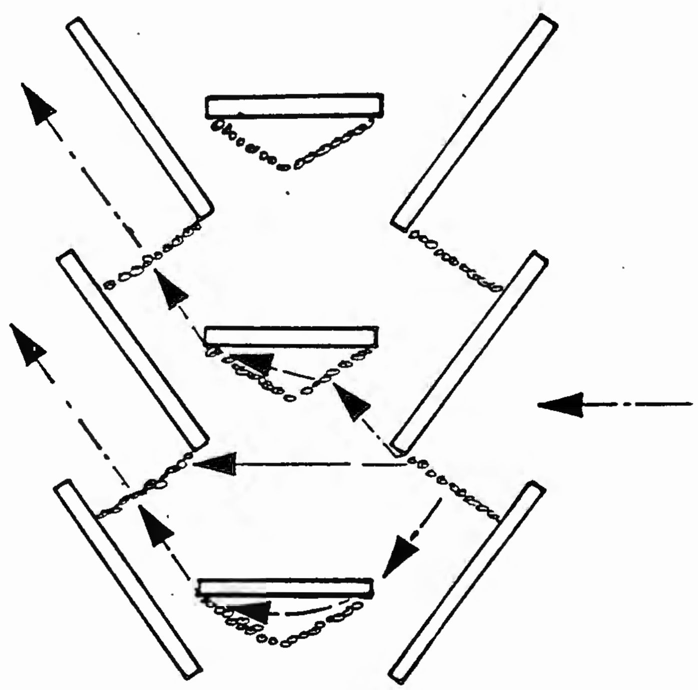
5.1.6 VOIDAGE

The pressure drop through the filter was found to decrease as the filter bed was allowed to move. However, the pressure drop was not further affected by sand flowrate over the interval tested in these experiments. The pressure drop variation could only be due to a change in panel thickness or a change in the area available for flow. Since appreciable changes to the bed cross-sectional shape and thickness did not occur, especially at low face velocities, a change in the flow area was suspected. The change in flow area would be caused by an alteration of the granular voidage induced by movement. An indirect method was used to estimate the moving bed voidage as this is an important parameter in the efficiency calculations discussed later on.

The pressure drop measured across a static bed over the range of face velocities was used to estimate the effective path length selected by the fluid flowing through the panel (as described above). Then assuming that little change in the filter cross section occurred once the bed started moving, (for a given face velocity) a change in voidage was assumed to be responsible for the change in pressure drop. The exercise was really only valid at velocities below 0,1 m/s where distortion of the filter bed cross-section was visually undetectable. The voidage calculated with this procedure for the full range of face velocities are shown in Table 4. It is concluded that the voidage does alter with granular movement and remains consistently between 0,42 - 0,43 over the range of velocities and granular flowrates used.

The complex cross-section of the granular bed, created by the louvres and baffle, presented some difficulties in pressure drop calculations. As seen in Figure 17, the sand bed presented a variable flow path length to the air stream which would naturally select the shortest route through the panel. Although one of the reasons the baffle was inserted was to make the flow distribution more uniform, some variation in path length was inevitable. At first, the average flow path length was calculated by dividing the volume of granular material in the panel by the face area of the filter. This method was found to be ineffective. The exact route selected by the fluid has not been established due to the complex configuration of the louvres and resultant cross section but a reasonable estimate of the flow path length for various flowrates has been estimated as shown below.

FIGURE 17: ASSUMED FLUID FLOW PATTERNS



The method used to estimate the flow path length adopted by the fluid over a range of face velocities is based on pressure drop measurements. It is assumed that the air flow path length is a function of the air flow to the filter panel. The higher the throughput the shorter the average path length used. The Ergun equation, Eq 25, is used to establish the path length encountered. The estimates of the average path used by the fluid are shown in the table below. These figures are based on a voidage calculated from the specific and bulk densities quoted by the supplier of the granular material according to the equation:

$$\varepsilon = 1\text{-bulk density/specific density} \qquad (26)$$

TABLE 4: ESTIMATES OF THE AVERAGE FLUID PATH LENGTH

FACE VELOCITY	STATIC BED VIODAGE	PATH LENGTH	MOVING BED VOIDAGE
cm/s	-	m	-
4	.4	.06	.42
8	.4	.05	.43
12	.4	.048	.43
16	.4	.047	.43

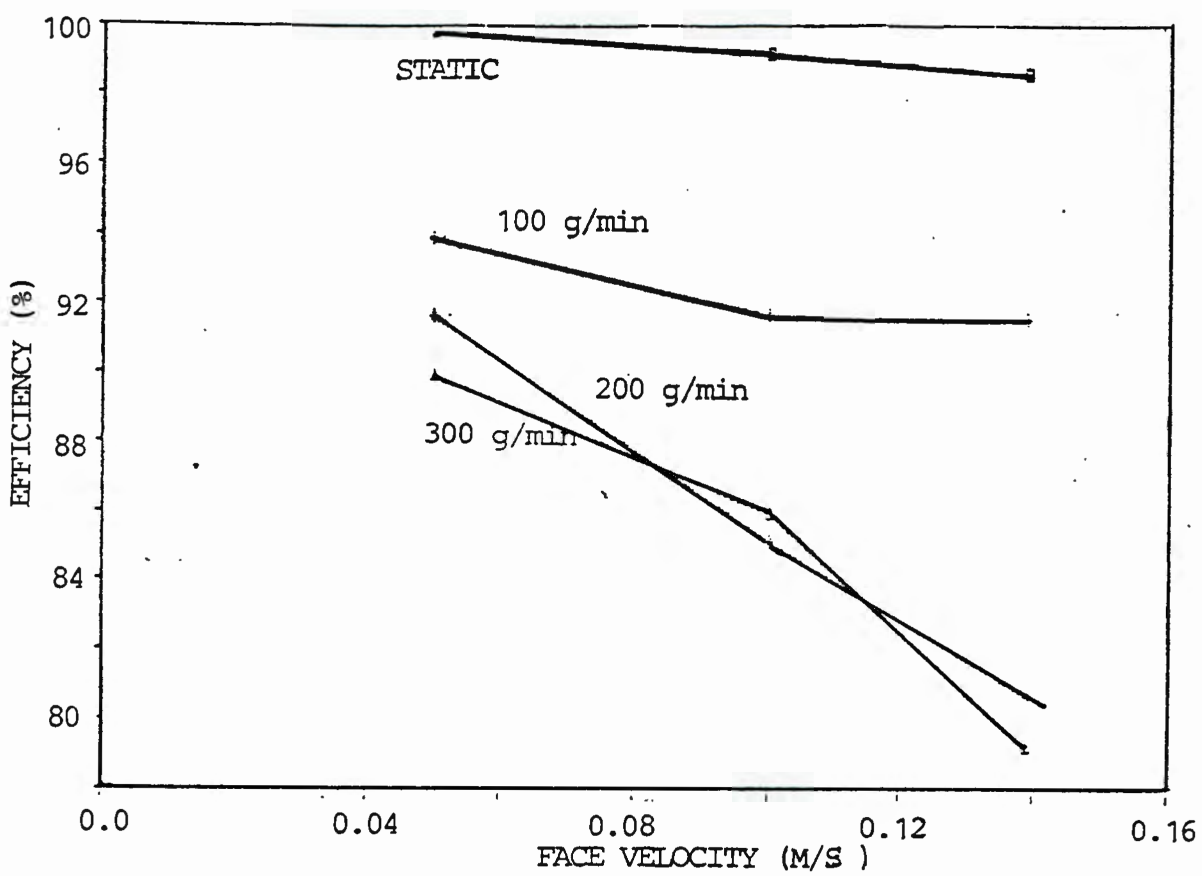
5.2 COLLECTION EFFICIENCY

The collection efficiency of granular bed filters is a function of accumulation of particulate in the bed but the initial collection efficiency is of particular interest, as it represents the worst collection efficiency of the filter. Design of granular filters often makes use of this conservative collection efficiency knowing that filtration efficiencies will improve as particulate accumulates in the bed. For design purposes, it is important to know whether the same characteristics apply to the MGPF. Experiments were thus conducted to measure initial collection efficiencies and then to determine whether the collection efficiencies improved with accumulation of particulate in the filter bed.

5.2.1 INITIAL COLLECTION EFFICIENCY

Initial collection efficiencies were measured over a range of granular flowrates, starting from a clean bed condition, and recording efficiencies for the first 10 minutes of operation. Dust used in this experiment was collected from a coal fired FBC cyclone discharge. The dust was passed through a sieve to separate the <75 micron fraction which was used in the experiments. The graph below illustrates the effects of sand flow on initial collection efficiency.

FIGURE 18: EFFECT OF SAND MOVEMENT ON INITIAL FILTER EFFICIENCY

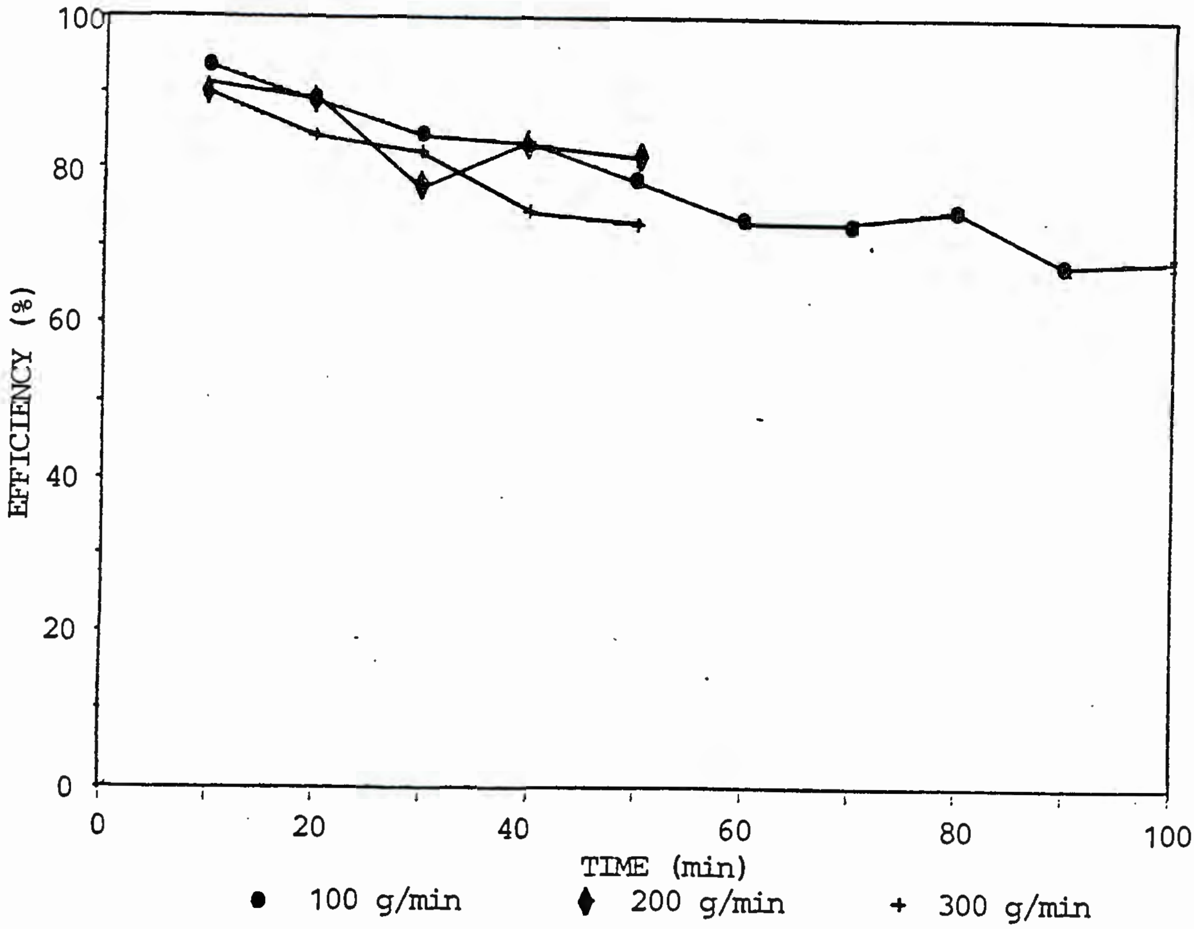


Static medium conditions are included to demonstrate the effect of movement of the filter medium on filtration efficiency. The results show that the efficiency is drastically reduced by filter bed movement. In addition, the efficiency deteriorates with increased filter medium flowrate. Increases in face velocity also reduce the collection efficiency but more so at the higher sand flows tested, indicating an improved operation with slower sand flow. The limit to this would be a stationary filter bed resulting in clogging of the filter and high pressure drop. Thus it is necessary to trade off efficiency against the increased pressure drop. This project aimed to maintain a low operating pressure drop by increasing granular medium flow keeping dust concentrations in the bed low. The operating point was chosen to establish low pressure drop collection efficiencies and because this condition is a good approximation of the initial bed conditions where modelling is more accurate.

5.2.2 INCREMENTAL FILTER EXPERIMENTATION

The poor initial collection efficiency of the moving granular bed indicated inferior particulate cleanup ability in this mode. However it is well known that efficiencies improve with extended operation where granules are not as clean as fresh filter medium. In this series of tests, filter medium cleaning was not attempted and extended operation was achieved by continuous addition of clean sand. The efficiency of the moving granular bed design was not expected to improve as a clean condition was maintained by the continual replenishment of the filter medium. The clean filter bed was expected to maintain initial collection efficiencies and experiments were thus conducted to establish this. The experiment was conducted over a series of short time increments with filtration efficiency measured for every increment. In this way it was possible to map the collection efficiency from the initial clean state. The data was taken every 10 minutes and is presented graphically below.

FIGURE 19: EFFECT OF DUST ACCUMULATION ON FILTER EFFICIENCY



The MGPF collection efficiency was found to decrease with time and not maintain the initial clean bed capabilities. None of the incremental tests were run for sufficient length of time to establish an equilibrium between accumulation and re-entrainment of particulate. The implication of the decrease in efficiency during the approach to equilibrium was that the initial collection efficiency now had no relevance to moving granular beds. However the decrease in collection efficiency was not seen as a problem of particle capture but more one of severe reentrainment initiated by the intergranular movement.

The experiments illustrated that efficiencies, in the case of a moving bed, deteriorate from the initial clean granule conditions. Extended test runs were thus necessary to reach equilibrium to evaluate the final performance of the panel section but due to the lack of automated granular recirculation facilities, the equipment did not lend itself to extended test runs. The extended operation was therefore scheduled for the high temperature work where the equipment was automated to a greater extent.

5.3 HIGH TEMPERATURE EXPERIMENTS

The high temperature operation aimed to demonstrate the filter's performance as a FBC flue gas filter. It was intended to operate the filter panel over the range of fluid bed operating conditions and monitor the filtration efficiency and pressure drop whilst attempting to maintain clean granule conditions. The experience gained from the preliminary ambient temperature work was considered invaluable, producing knowledge of the filter's operation that might well have gone unnoticed, in particular the fluid flow effects on granular movement. The results of the preliminary work were used to improve the filter design for the performance runs, the objectives of which were:

1) To evaluate whether granular recirculation could be used to maintain clean granule conditions and keep pressure drop representative of that condition.

2) To document collection efficiencies over a range of FBC operating conditions whilst maintaining clean filter conditions.

3) To investigate the mechanisms involved in the deterioration of filter efficiency associated with a moving granular bed.

4) To establish improvements necessary to this specific design to increase filtration efficiencies sufficient to meet air pollution and turbine inlet specifications.

The test filter unit was designed to operate with a certain degree of automation so that extended test runs could be accomplished. As already acertained in the preliminary work, filter efficiencies deteriorated from clean conditions and extended test runs were deemed necessary to reach some equilibrium in the operation.

5.3.1 DATA COLLECTION

As this experiment was intended as a first order assessment of the MGPF, the use of fabric filtration to collect particulate samples was considered accurate enough for measurement of overall collection efficiencies. The collection efficiency of the fabric filter was tested to support this assumption. A dust generator was connected to a test filter bag and a sample blown into it at a rate similar to that used in the experimental work. The weight of the dust generator before and after the experiment allowed the total weight of dust produced to be calculated. The increase in weight of the fabric filter before and after the experiment represented the mass of dust collected by the filter. Comparing dust produced and collected demonstrated 100% collection efficiency to within 0.01g for a sample weight of 112g.

Flue gas temperatures at the filter inlet were too high to effect sampling by fabric filtration. Measurement of dust concentrations here was based on isokinetic sampling. The particle size range of the fly ash (submicron - 400 micron) necessitated the use of isokinetic sampling to reduce inaccuracies due to inertial effects during the drawing of the aerosol sample. The isokinetic probe was fitted to the freeboard of the FBC where flow conditions were best suited to isokinetic sampling. The absence of cyclonic precleaning generated high dust loadings at the filter inlet, an order of magnitude higher than the downstream dust loading. Meaurement of the filtration efficiency was thus not as sensitive to the higher upstream concentration as the downstream concentration. This condition suited the sampling configuration being used as the reduced sensitivity of the upstream measurement could accommodate the wide scatter in isokinetic sampling data typical of FBC elutriation rates as reported by Wen and Chen (32). The more sensitive downstream measurement was measured using the more accurate fabric filter. Apart from using a more accurate sampling system, additional efforts were made to improve the measurement of the downstream dust concentration. The improvemnts included increasing sampling frequency, filtering the total sample rather than an isokinetic sample and extending the sample time for as long as the filter element pressure drop could be accommodated. It was fortunate that accuracy was not dependant on the upstream sample because the isokinetic sampling proved to be cumbersome, resulting in a lower sampling frequency, to an extent where, in several cases, one upstream sample had to be used for a number of collection efficiency calculations, within one test run. The upstream dust loading was considered only that dust which actually entered the filter medium so that dust settling out in the upstream cowling had to be collected and subtracted from the isokinetic results.

Pressure drop experienced by the flue gases passing through the granular filter was measured using a water manometer with tappings in the upstream and downstream cowlings. An effort was also made to keep the pressure inside the upstream cowling close to atmospheric pressure by controlling the draft generated by the ID fan. Control of this pressure was done to eliminate ingress of air into the upstream side of the filter aswell as to eliminate losses of flue gases to the atmosphere.

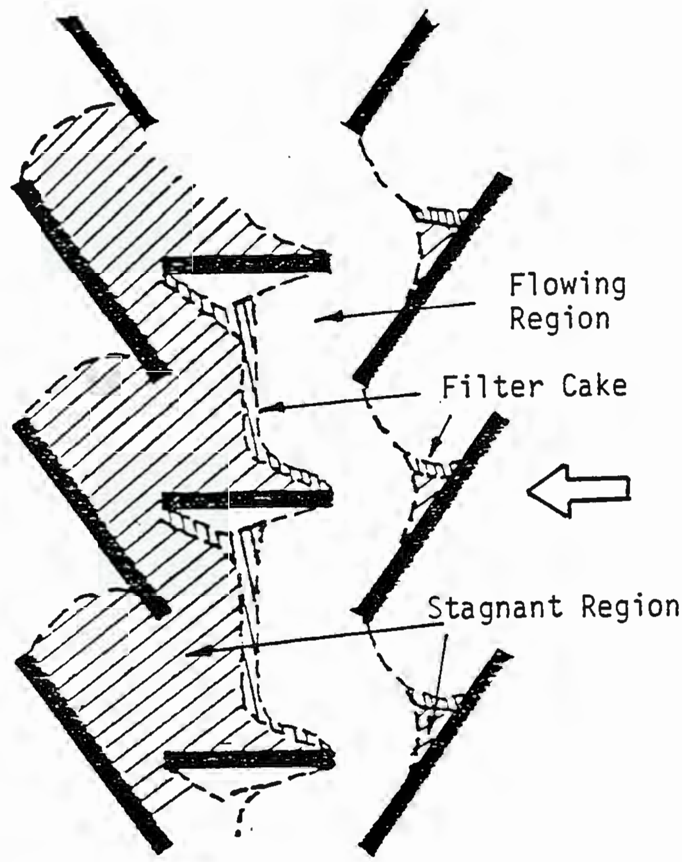
The data collected for the test runs conducted over a range of operating conditions are tabled in Appendix A.

5.3.2 CAKE FORMATION

Early experimentation with the filter panel in a vertical orientation is documented to demonstrate the effect of poor granular flow. The results of these experiments are tabled in Appendix B.
These initial testruns were characterised by high filtration efficiencies and increasing pressure drop which could not be regulated by granular flow. Eventual blowthrough of the panel forced termination of each run and prompted the detailed study of granular flow patterns within the panel.

The detailed study as explained in the previous chapter showed that under certain conditions flue gas flow through the granular bed reduced the granular flow on the downstream face of the filter to a standstill. Particulate then accumulated at the interface between moving and stationary sand forming a filtercake as shown below. The filter cake restricted flue gas flow progressively with time, a condition obviously not acceptable for continuous filtration, prompting further investigation into granular flow patterns. Tilting the panel off the vertical proved to be one solution to the granular flow problems which was then incorporated into the high temperature rig.

FIGURE 20: FILTER CAKE FORMATION



5.3.3 OFF VERTICAL PANEL ANGLE

It was realised on completion of the preliminary work that some modifications to the filter panel would ultimately be necessary to improve the filtration efficiency to that required for flue gas cleanup. Nevertheless, it was decided to complete the high temperature work with the same configuration to enable direct use of the information generated by the preliminary work in the higher temperature experiments. A pilot scale panel filter was thus constructed in mild steel using the same dimensions shown in Figure 11.

It was intended to experiment with the filter panel over a range of operating conditions particularly face velocity. The results from the study of granular flow as seen from Figure 12 showed that the optimum angle required was a function of the face velocity. Contemplating a range of velocities then ideally required accurate control of the panel off vertical angle. The dexterity required for this was considered impractical resulting in one angle being chosen. The panel angle employed was 6^{o} which referring back to Figure 12 indicates an attainable design face velocity range of 5 - 13 cms/sec where granular flow on the downstream face could be assured. However it was realised that a granular flow differential between the upstream and downstream faces would develop at face velocities other than 13 cm/sec.

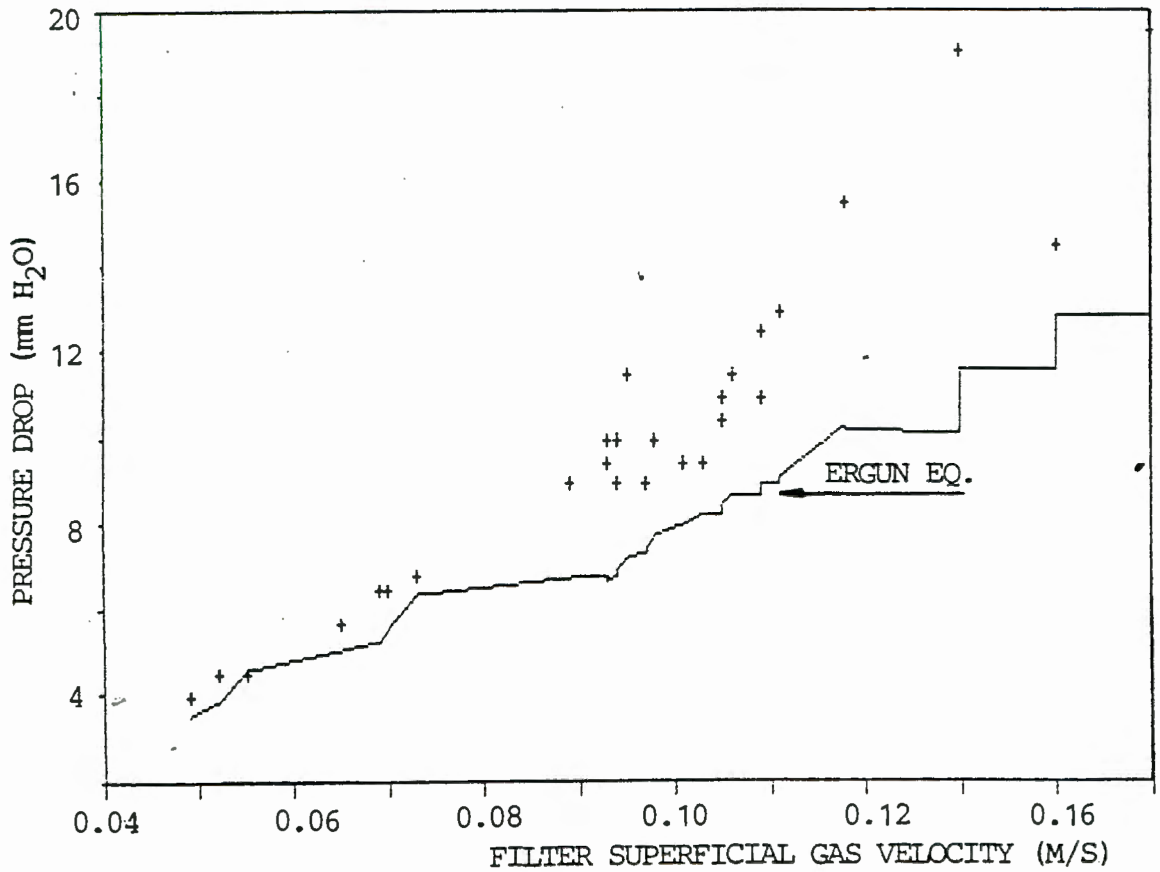
5.3.4 PRESSURE DROP RESULTS

The test conditions and corresponding operational results for the coal fly-ash trials from Appendix A are shown in the graphs that follow. The first records the operating pressure drop and the equivalent clean granule pressure drop, demonstrating that conditions approximating clean granule conditions were maintained in the data selected for evaluation.

The objectives of this project were to determine what collection efficiency could be achieved in a granular bed filter operating with a low particle loading in the filter bed and low operating pressure drop. The final design pressure drop may eventually need to be a compromise between an economic optimisation of fan power costs and filter bed regeneration costs, but, as a starting point, it is assumed that the bed will be regenerated at a rate sufficient to maintain clean granule pressure drop ie corresponding to a minimum pressure drop for a given face velocity and granule size. The clean granule pressure drop was calculated using the modified Ergun Equation 25.

The theoretical clean granule pressure drop superimposed on the pressure drop results was calculated using the Ergun equation allowing for temperature dependance of viscosity and density. Voidage used was that calculated for moving bed conditions in the preliminary work as described in the previous chapter.

FIGURE 21: PRESSURE DROP RESULTS

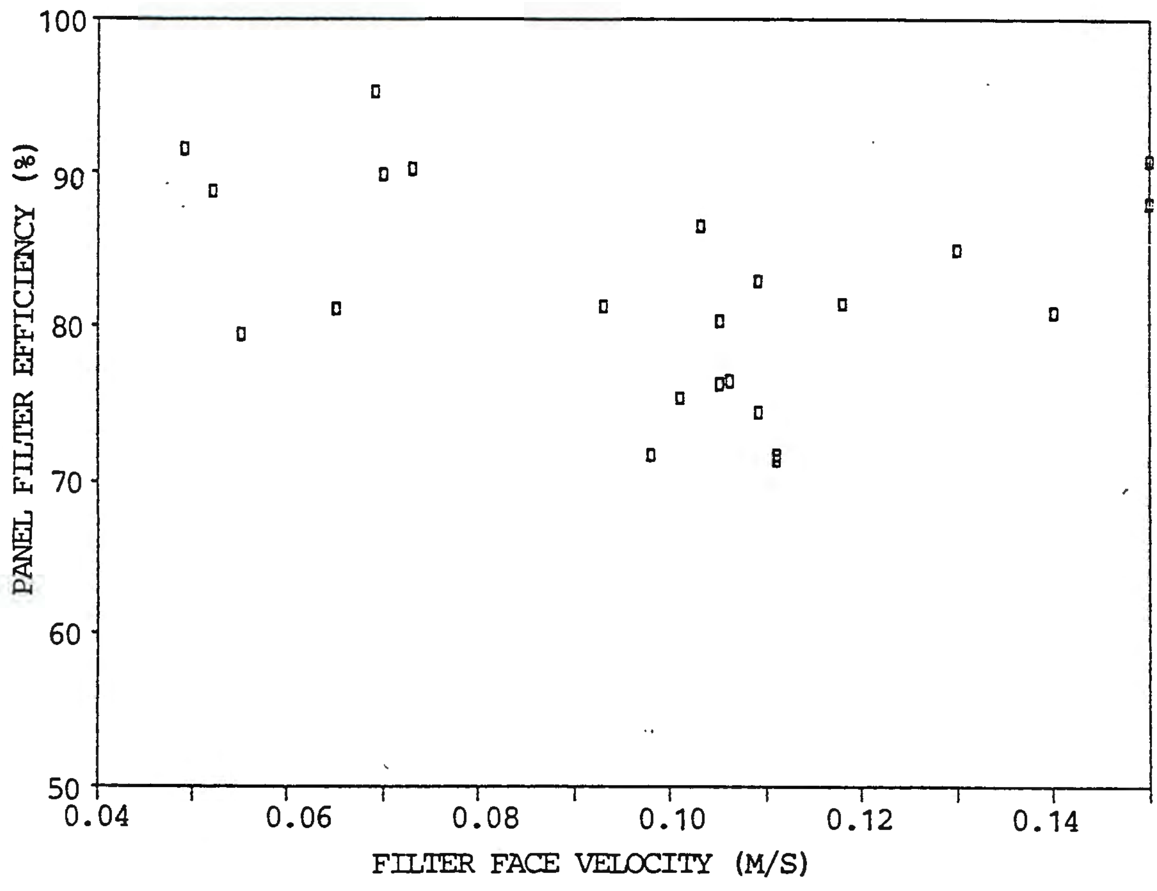


5.3.5 COLLECTION EFFICIENCY

The FBC was maintained at an operating temperature of 850 ^{o}C throughout the experiments by manual control of the coal feed rate and exturnal cooling air supply. Oxygen concentrations of the flue gases were continuously monitored to maintain 20% excess air in the flue gas to assist in temperature control and maintain some degree of consistency in the fly ash properties. Fluidising air was varied over a range of velocities namely 2 - 5 times the minimum fluidisation velocity, which could be considered typical of FBC operating conditions. Increasing the fluidisation velocity introduced higher temperatures and heavier dust loadings in the flue gases. It was the intention of this experiment to operate the filter at the FBC operating temperature, 850 ^{o}C, but heat losses prevented this achievement. Temperatures at the filter face reached a maximum of 414 °C and dust loadings varied over a range of 8 - 42 g/nM^3. Filtration efficiencies recorded over these conditions ranged between 70% - 95% for experiments where clean bed conditions were maintained. Granular recirculation rates used to maintain clean granule conditions resulted in dust loadings in the filter medium ranging between 1 : 27 to 1:130.

To present a broader perspective of the results a series of graphs are presented below. The numerical values on which these graphs are based are listed in the appendices.

FIGURE 22: OVERALL COLLECTION EFFICIENCY

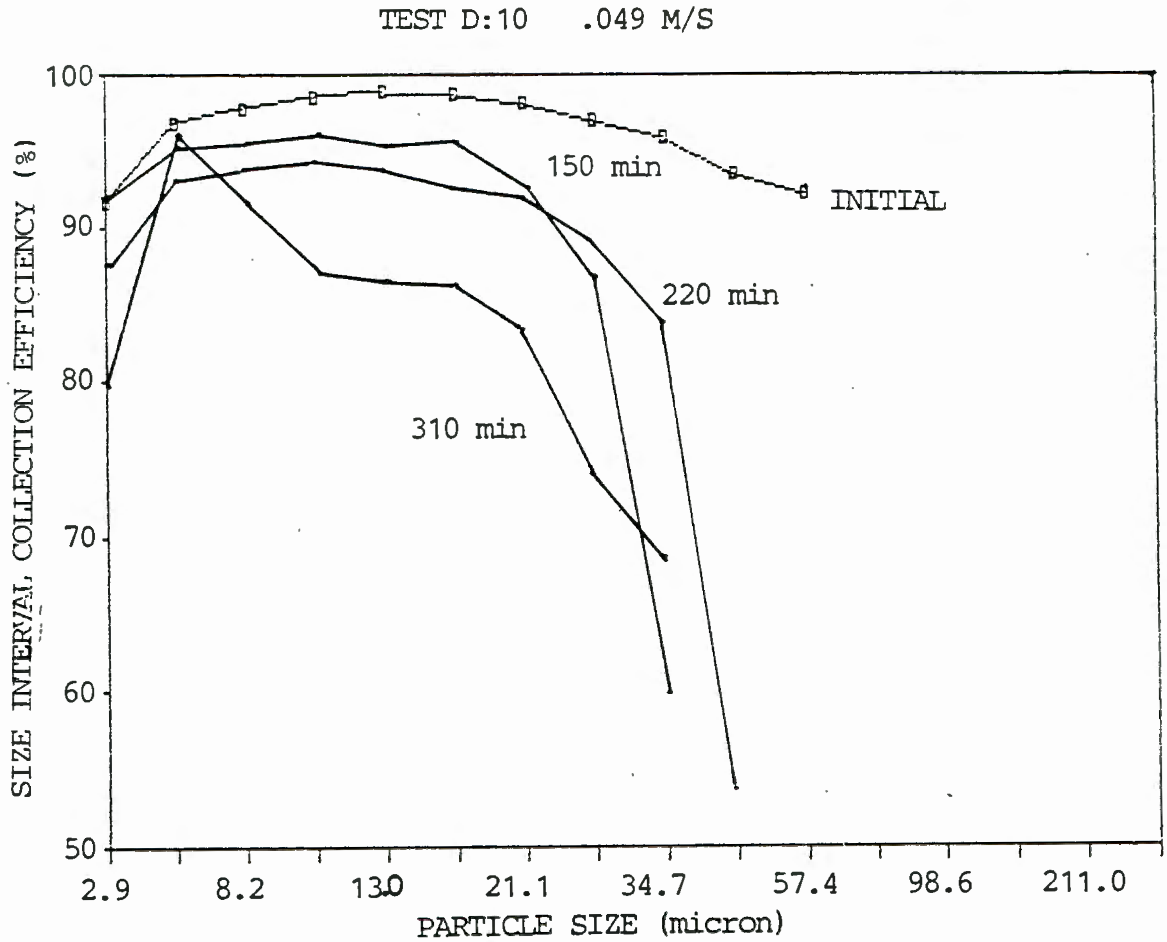


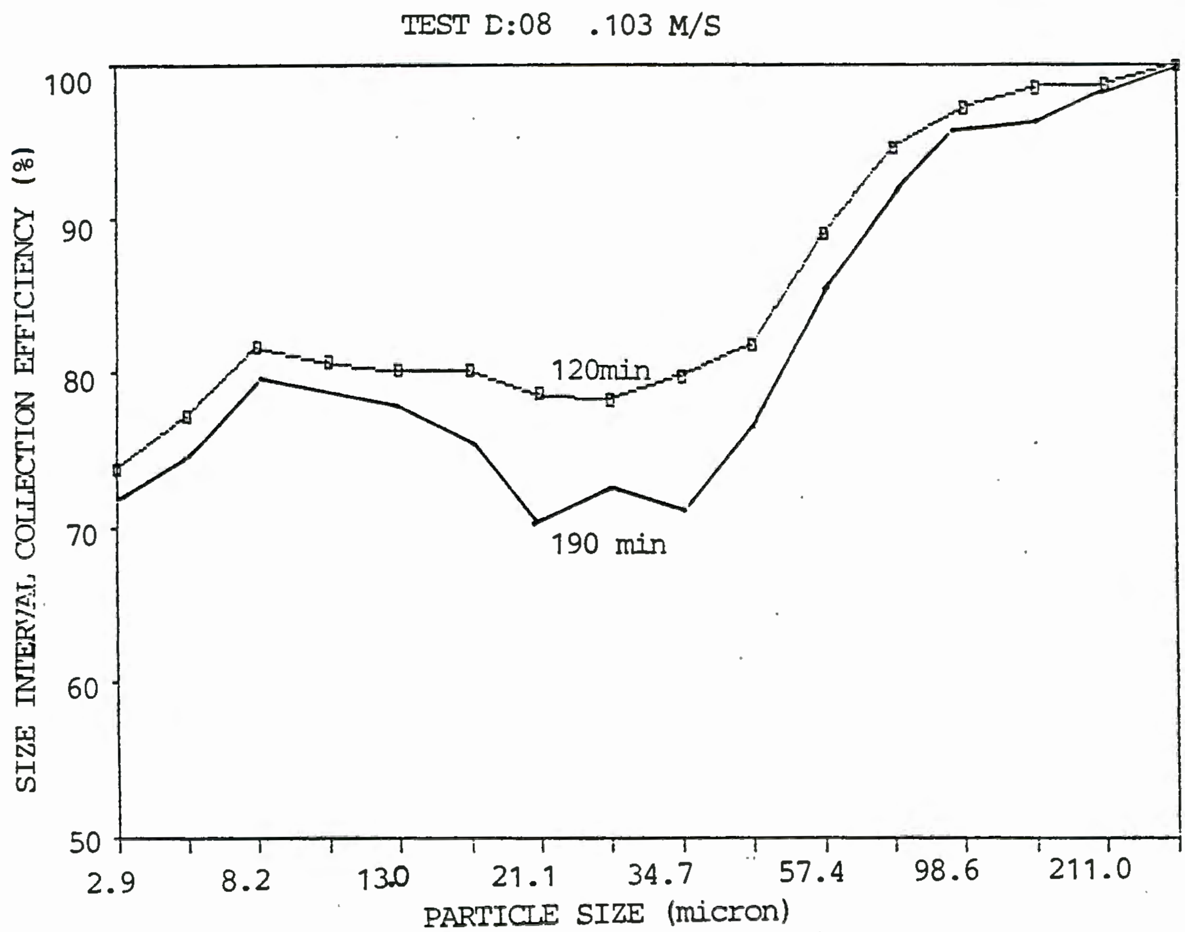
The results show that overall collection efficiency deteriorates with increased face velocity. The granular flow controller was unfortunately unable to regulate flow with much consistency as inclusion of particulate within the medium inhibited normal free flowing characteristics. As particulate was accumulated by the granular material progressing down the filter face, the increase in dust retained in the filter medium tended to reduce granular flowrates over the duration of the experiment. This was not considered a problem as long as the pressure drop remained representative of clean granule conditions. Pressure drop was maintained at this level but accurate control of the granular flow could not be achieved with the rotary valve used in this experiment. Consequently the results do not show the exact effects of granular flow or bed loading on collection efficiency. Temperature and inlet dust loading effects on collection efficiency could not be resolved because they were coupled with face velocity.

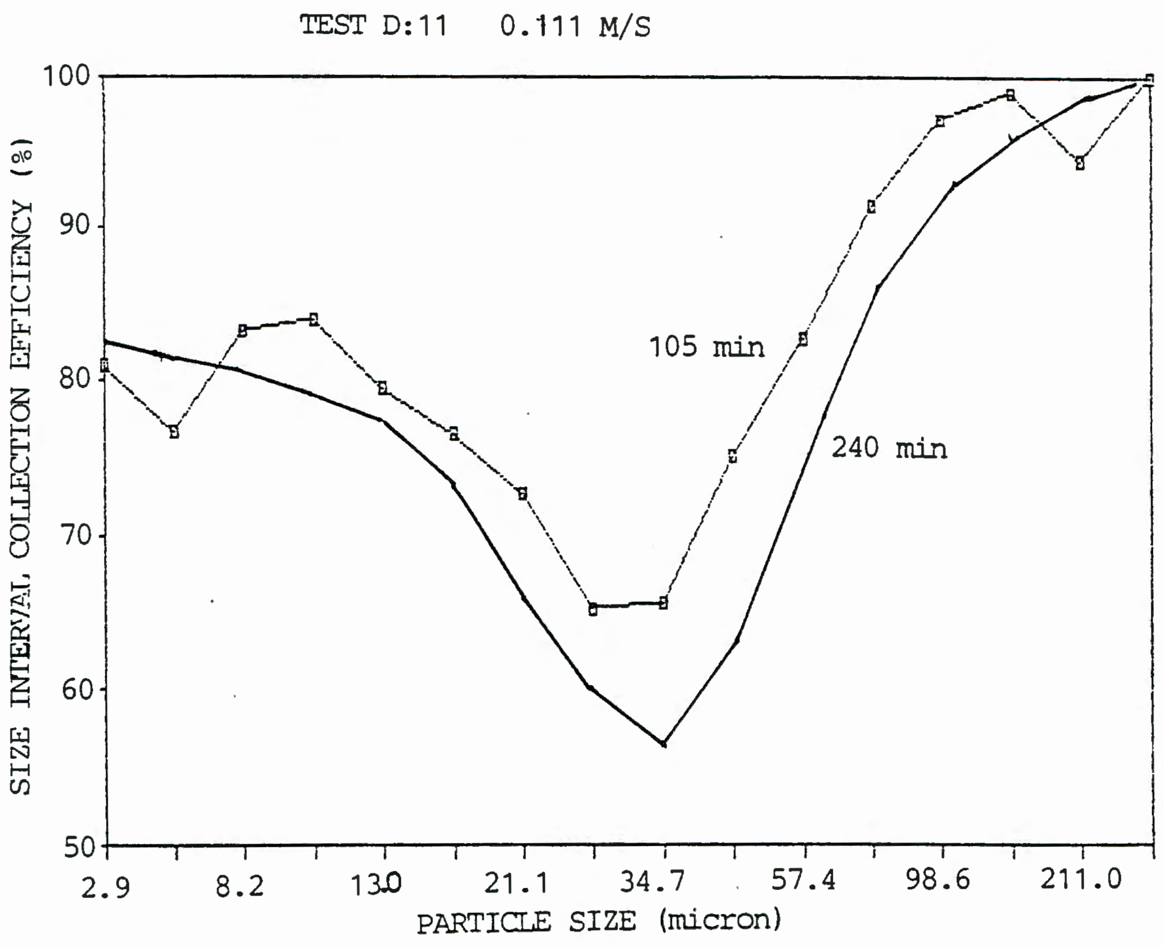
The progressive deterioration of collection efficiency with time noted in the preliminary experiments was also found to be prevalent in the high temperature experiments. Particulate entering the filter with the fresh granular medium could be ruled out in the preliminary experiments because the medium flow was only passed through the panel once during each experiment. The high temperature work was designed for longer test runs necessitating granular flow recycle. The pnuematic flow system returning the medium to the feed hopper doubled as a granule/dust separator the exact performance of which was not quantitatively established. It was thus possible that the efficiency deterioration was further exaggerated by poor granule cleaning. To investigate this possibility, a comparison was made between filter efficiency measured in the clean bed and those values documented once "cleaned" granules had been returned to the filter panel. No obvious deterioration was noted.

To assess whether any particle size was responsible for filter efficiency deterioration samples of the dust collected in the fabric filter were analysed. Conducting the same analysis of dust entering the filter panel permitted collection efficiencies of different particle size fractions to be calculated. Analysis of a number of samples from moving granular bed conditions showed consistent deterioration with time in the particle size range of 30 - 40 micron. The graphs presented below illustrate the results over a range of operating conditions.

FIGURE 23: FILTER EFFICIENCY DETERIORATION






TEST D:11 0.111 M/S
100
90
80
70
60
50
SIZE INTERVAL COLLECTION EFFICIENCY (%)
105 min
240 min
2.9
8.2
13.0
21.1
34.7
57.4
98.6
211.0
PARTICLE SIZE (micron)

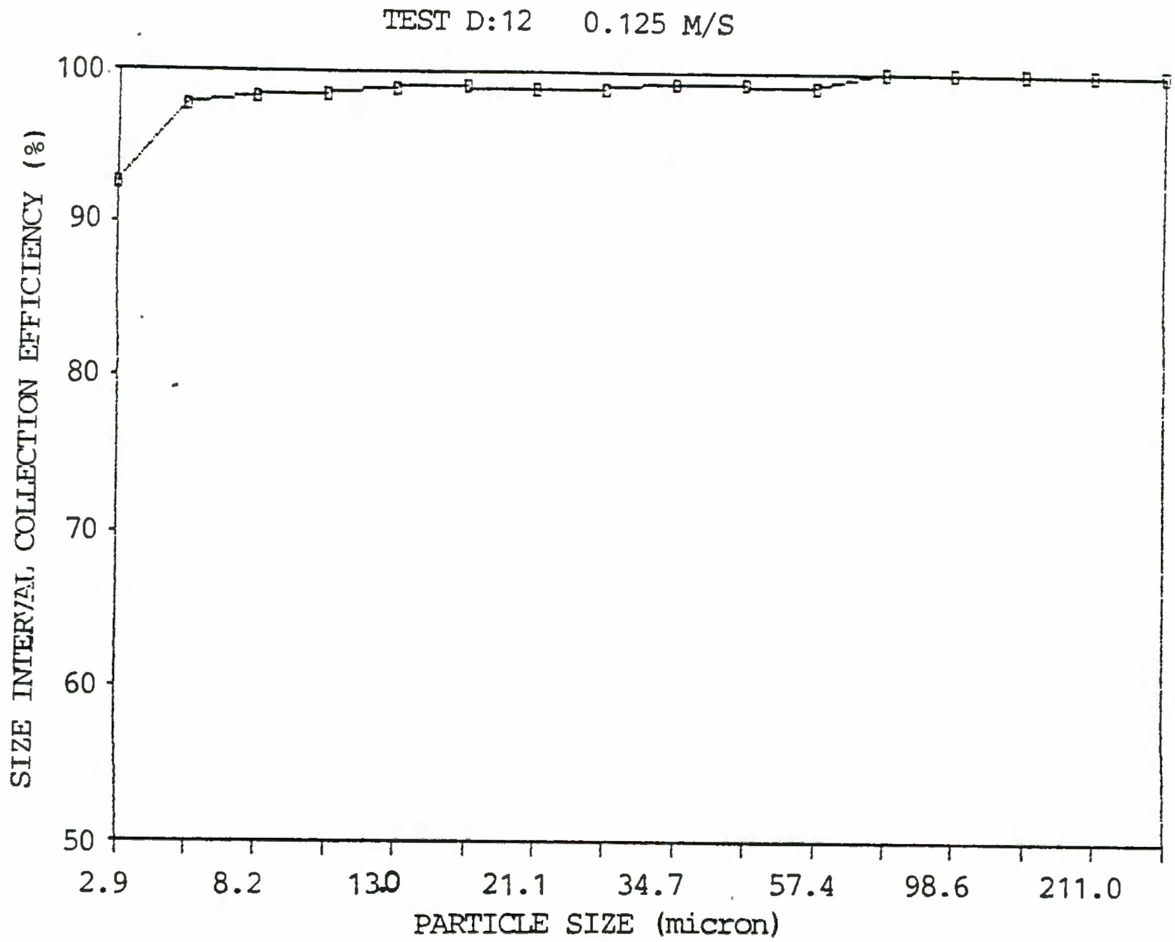
When the fabric filtration efficiency for sample capture was assessed, tests were also conducted to see whether a sample taken from the fabric was affected during removal from the cloth. This assessment was necessary to make certain that the results displayed in the graphs above were not a result of the sampling technique. The results of the test in which 100% overall collection efficiency was previously recorded showed no obvious trends that might explain the collection efficiencies as recorded in the graphs above. The sampling technique could then be assumed capable of producing a sample for particle size analysis with accuracy suitable for first order assessment. The ability to measure various particle size collection efficiencies with reasonable accuracy opened the way to comparing the recorded collection efficiency data with models published in the literature.

5.3.6 STATIONARY BED COLLECTION EFFICIENCY

Initial collection efficiencies were measured while keeping the sand in the filter panel stationary. These collection efficiencies should have been similar to the moving bed collection efficiencies if the granular movement did not affect collection or retention mechanisms. Collection efficiencies of between 98% - 98.7% were recorded even though the pressure drop indicated reasonably clean conditions. In general the moving granular bed therefore proved to have reduced collection efficiencies over the clean granule static bed condition.

Analysis of collection efficiency at different particle sizes did not demonstrate the very poor collection efficiency in the 30 - 40 micron size range as shown below.

FIGURE 24: FILTER EFFICIENCY FOR STATIC BED CONDITIONS



5.4 THEORETICAL INTERPRETATION

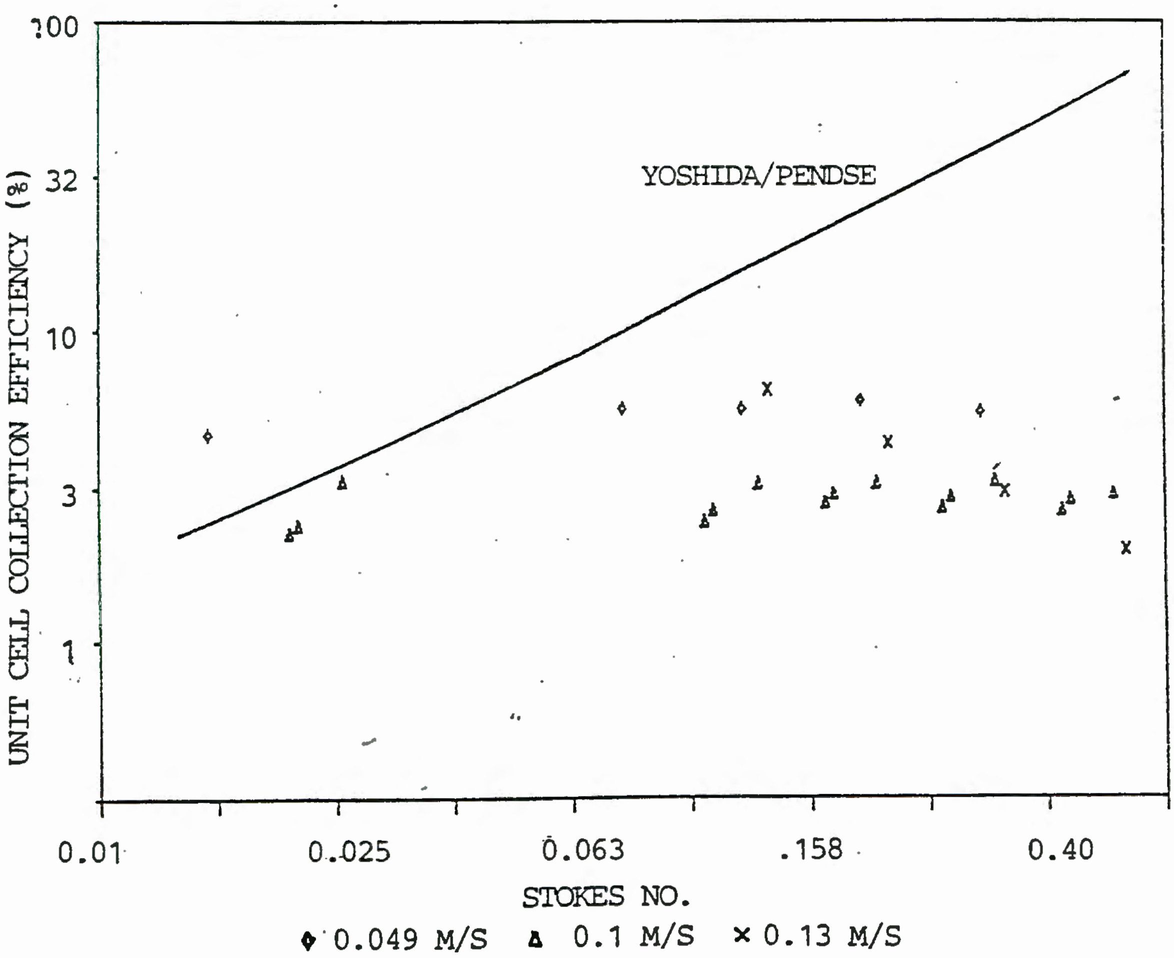
The particle size of the fly ash generated by the FBC was found to range from 260 micron down to sub micron. As the filter was to operate in a primary role it was necessary to consider the filtration efficiency across the whole size range. The results of particle capture and retention were demonstrated to vary significantly over the particle size range produced by the FBC. In general granular bed design, in combined cycle cleanup, operate in secondary and tertiary roles coupled to primary cyclone cleanup. Upstream cyclones are capable of removing essentially all particulate larger than 10 micron with the result that the bulk of research and development work on granular bed filtration is restricted to the less than 10 micron size fraction. Very little work has been published on aerosol filtration in granular beds for particles greater than 10 micron. As a result interpretation of the experimental data from this project is divided into two areas, less than 10 micron and greater than 10 micron.

5.4.1 THE 'SUB-10 MICRON' FRACTION

Literature on granular bed filtration pays particular attention to modelling of the initial clean bed condition as explained earlier in chapter 3. A number of semi-empirical and empirical correlations for filter performance are reported, usually incorporating the contributions of a number of the capture mechanisms (6,26,27,28,29,30). Capture is assumed if a particle touches the granule surface. In this project, because continual replenishment of filter material maintained a pressure drop representative of a clean granular bed, it was assumed reasonable to model the panel bed filter as a clean filter. Attention was focussed on the larger size particles ie greater than 3 micron, as these could be reliably captured for analysis using the fabric sample filter. The dominant mechanisms responsible for the larger particles capture could then safely be assumed to be inertial impaction and direct interception. Brownian motion and gravitational influences exert an influence only on very small particles (typically less than 1 micron diameter), (6), and were thus outside the current range of interest. Electrostatic effects were acknowleged to be significant over the size range of interest but they could not be interpreted because of the difficulty in measurement of electrical charges carried by fly ash particles.

The correlations cited in the literature are all based on a single unit collection efficiency where the unit collector varies according to the model assumptions as discussed in Chapter 3. Two flow model types are generally assumed in the development of granular filtration models, internal and external flow models. The collection efficiency results for different particle size ranges shown in Figures 23 and 24, need to be presented in unit cell collection efficiency format for comparison with published models. The two different models have different unit cells and thus have different ways of calculating unit collector efficiencies from experimental data. For the internal flow models, experimental data was reworked into unit cell collection efficencies using Equations 14 and 15 in Chapter 3. Graphical representation of these experimental unit cell collection efficiencies as a function of the dimensionless Stokes No are shown below.

FIGURE 25: UNIT CELL COLLECTION EFFICIENCIES (YOSHIDA,PENDSE)



The correlation included in Figure 24 represents the initial unit collector efficiency from both Yoshida and Tien (28), and Pendse and Tien (26). Both correlations are based on internal flow models with modifications from experimental results introduced to account for fluid inertial effects as described in Chapter 3. To recap, the general form of the expression for unit collector efficiency is

$$\eta = B\left[N_{st} + 0.48\left(4 - \frac{4N_I}{d_c^*} + \frac{N_I^2}{d_c^{*2}}\right)^{\frac{1}{2}} \cdot \frac{N_I^{1.041}}{d_c^*}\right] \qquad (10)$$

where B is defined as the ratio of unit collector efficiency to the calculated value at zero Reynolds number. The exact form of B is

$$B = 1 + 0.04N_{Re} \tag{11}$$

(Pendse and Tien)

$$B = 7 - 6\exp(-0.0065N_{Re}) \tag{12}$$

(Yoshida and Tien)

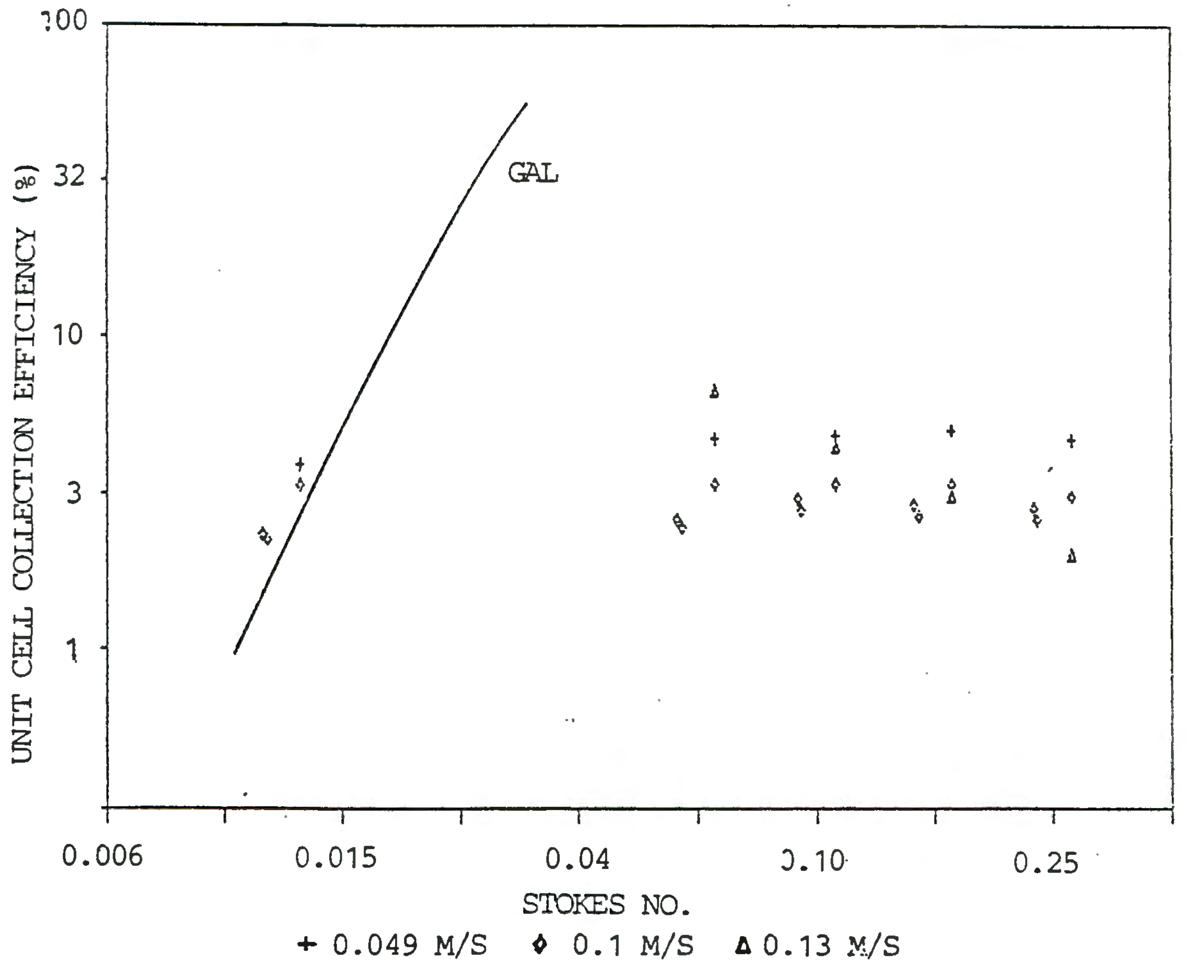
The range of validity of both the above correlations is not clear from the literature but it is obvious that they could not be extended beyond Stokes numbers of 1 as the predicted collection efficiency exceed 100% beyond Stokes numbers of 1.

Gal et al (30) have presented an external flow model developed from trajectory analysis of dust particles using a three dimensional flow field, based on a dense cubic lattice of filter grains. The final expression, also found empirically, takes the form

$$\eta = 2 \cdot N_{st}^{\prime\,3.9} / \left[4.3 \times 10^{-6} + N_{st}^{\prime\,3.9}\right] \tag{19}$$

where the value of the Stokes number has been modified to account for fluid inertial effects using Eq 20 from Chapter 3. The external model by virtue of its different unit collector geometry has a different procedure for calculating unit collection efficiencies from experimental data. The data from this experiment has been modified to represent the external unit cell collection efficiencies using Eq 17 from Chapter 3. The experimental results shown in relation to the correlation by Gal et al is presented below:

FIGURE 26: UNIT CELL COLLECTION EFFICIENCY (GAL ET AL)



Gal et al have defined a very narrow range for use of their correlation given by $0.01 < St' < 0.03$. (30) Extrapolating the correlation beyond these limits displays the same shortfalls as Yoshida's and Pendse's correlation, namely collection efficiencies in excess of 100% for increased Stokes numbers. Gal acknowledges that collection efficiencies are reduced by particle bouncing occuring at high Stokes Nos ($St > 0.02$) but no correction is cited to account for these effects.

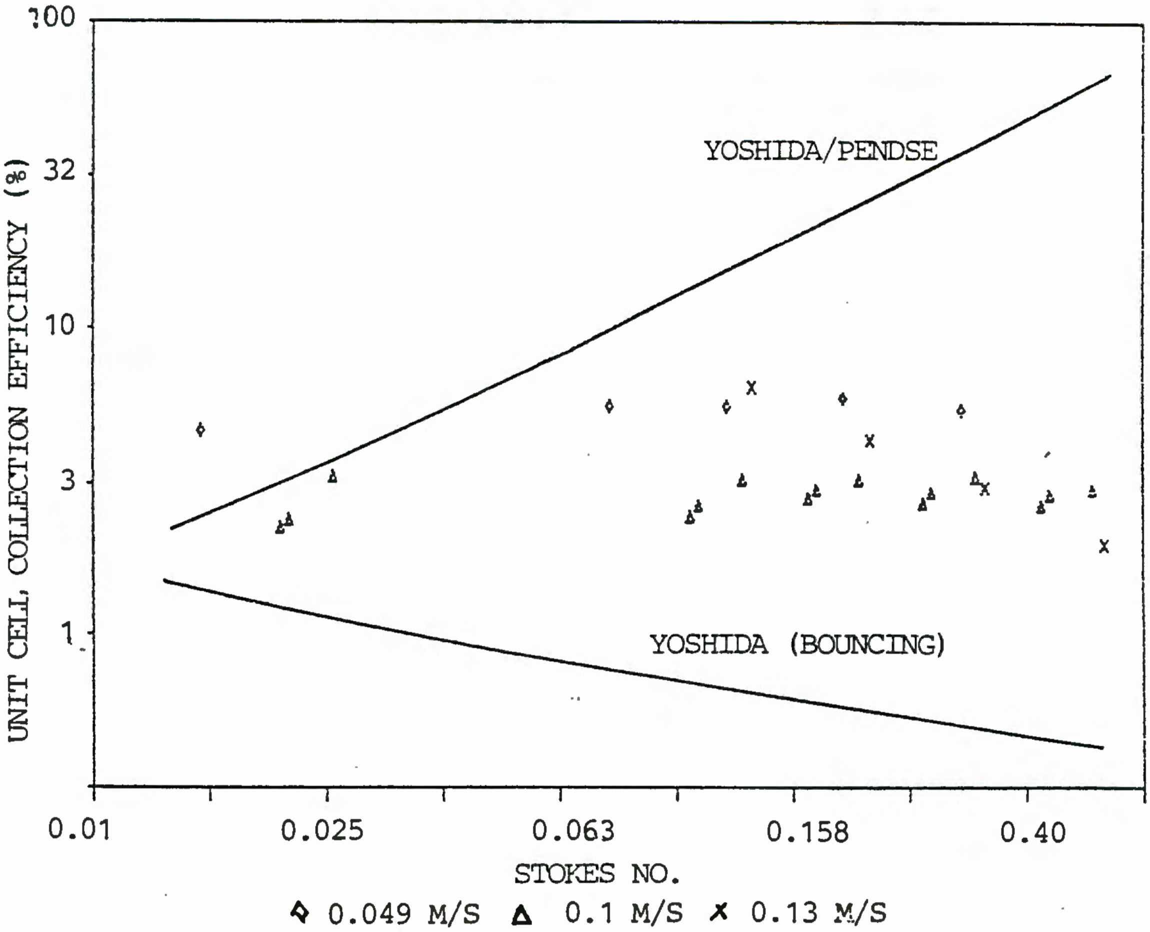
Comparison of the experimental data from this project with both correlations above is complicated by poor resolution of the data in the smallest size fractions. It is however reasonable to conclude that deviation in collection efficiencies from both the correlations occurs at Stokes Nos of 0.05. Yoshida and Tien (28) have proposed a modification to their original correlation for the range of Stokes Nos greater than 0.01, to account for, as they call it, "particle bouncing", or non retention of particles at high Stokes numbers. This is perceived to be a function of particle kinetic energy and is therefore related to

particle size as well as particle velocity. The empirical modification is expressed in terms of an adhesion probability, γ ,which is defined as the ratio of measured collection efficiency, to that obtainable if all particles coming in contact with filter grains are retained. The exact form of γ is given by
equation (26).

$$\gamma = 0.00318\ N_{st}^{-1.248} \tag{26}$$

The effect of the modification to Yoshida's correlation is depicted below.

FIGURE 27: PARTICLE BOUNCING EFFECTS (YOSHIDA)



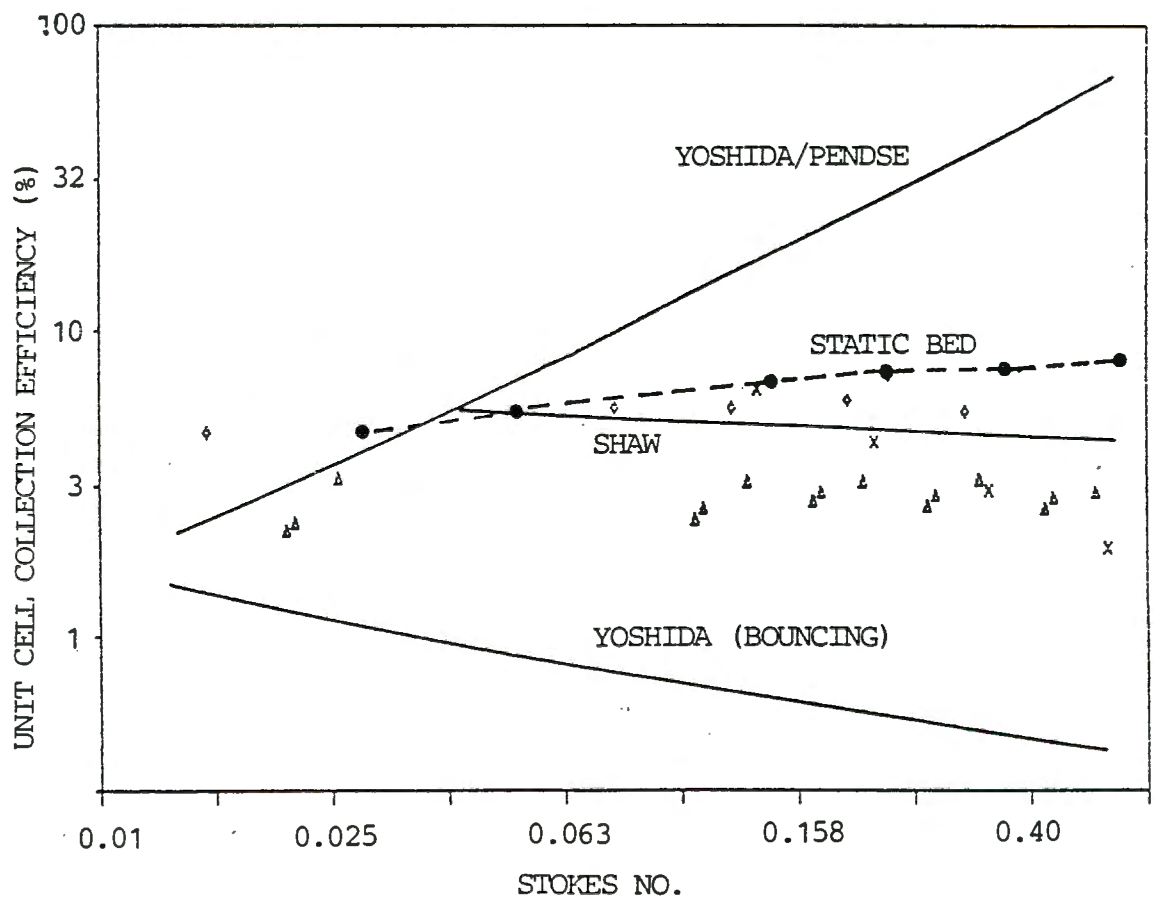
Yoshida does not define an upper limit to the validity of the modified correlation either but it shows consistent trends with the data produced in this experiment up to Stokes Nos of 1. Other influences such as physical straining are introduced above Stokes Nos of 1 improving the unit cell collection efficiency. No correlations based on Stokes Nos in this region have been found for aerosol capture.

In the case of a moving granular bed with intergranular movement, and the resulting disturbances created, it is more appropriate to link particle bouncing and movement related reentrainment together as a reentrainment probability. Defining a " reentrainmnet probability" , Θ , in an analogous fashion to Yoshida regression analysis of the experimental data produces:

$$\theta = 0.0221\ N_{st}^{-1.018} \qquad (27)$$

A comparison of the adhesion probability as predicted by Yoshida and the reentrainment probability from this work is shown below:

FIGURE 28: REENTRAINMENT PROBABILITY



◊ 0.049 M/S Δ 0.1 M/S × 0.13 M/S

The Yoshida correlation does not predict results for these experiments but the same trends are evident. The experimental data displayed higher retention probabilities than those predicted by Yoshida's correlation. From Figure 28 it can be seen that granular movement results in reduced retention over static bed conditions over the same range of Stokes Nos. The explanation for the difference between the experimental data and Yoshida's work was not resolved but could be related to differences in particle size and shape. However the similarity in trends displayed indicate that particle bouncing certainly contributes to the deterioration of collection efficiencies with increasing particle size. As both the static and moving bed experiments were completed with relatively clean granules, particle capture mechanisms should be similar. The similarity between initial collection efficiency for moving and static bed operation indicates deterioration of particle retention rather than reduction in particle capture.

5.4.2 THE COARSE PARTICLE SIZE FRACTION

As this filter is designed to keep the granular medium moving it is important to assess why the granular movement reduces the collection efficiency, especially in the larger size fraction where effects were shown to be most significant. The results show that the effect of granular movement is most significant in the 30 - 40 micron size range. In addition, granular movement increases the largest particle size penetrating the bed. Inertial effects of the larger particle sizes guarentee particle to granule contact and filtration efficiency thus becomes a function of retention rather than of capture. Retention of larger size particles is discussed almost exclusively in aqueous filtration studies but as the effects are mostly due to physical mechanisms the findings should apply to aerosol filtration. Discussion that follows is taken from aqueous filtration studies.

Herzig et al (34) discuss retention sites for particlate within a granular bed, identifying four types:

1) surface sites - particle retained on granule surface.

2) crevice sites - particle wedged between convex surfaces of two adjacent grains.

3) constriction sites - particle cannot penetrate pore smaller than its own diameter.

4) cavern sites - particle retained in a sheltered area eg small pocket formed by several grains.

Surface effects for larger particles (d_p < 10 micron) are reported (34) to be of the same order of magnitude as volume effects up to 30 micron whereafter volume effects dominate filtration mechanisms. This is consistent with the experimental results from this project which display improvement in collection efficiency due to volume effects (sieving) beyond the 30 - 40 micron particle size range. Maroudas et al (35) report collection efficiencies in relation to collector grain diameter and reports low particle retention for particle size < 0.065 d_g. The grain size used in this experiment was 740 micron which would represent a low particle retention size of 48 micron which is consistent with the data for this experiment. Herzig reports crevice sites to become important above a particle size of 0.05 d_g which could be used to explain the improvement of the collection efficiency seen in the experimental data at approximately 35 micron. Collection efficiency should reach 100% beyond 0.154 d_g as this is reported to be the minimum constriction diameter. This condition is evident in the static bed experiment but not in the moving bed case. The increase in minimum constriction diameter for the moving bed case is not unusual as it is consistent with the higher voidage recorded in moving granular beds in the preliminary work.

Gradual decrease in collection efficiency is reported to occur for larger particle sizes by Maroudas et al (35) to a point where bed saturation occurs and further no particle retention is possible. Indications of this condition are evident in the progressive deterioration of the 30 - 40 micron collection efficiency with time. This condition would be the result of too low a granular circulation rate, a problem easy to solve but the condition reported by Maroudas is accompanied by high pressure drop, not evident in these experiments.

The most likely cause for the deterioration of the collection efficiency is the intergranular movement accompanying granular recirculation in this specific design. The louvre and baffle arrangement used to enhance the filter surface regeneration impart a rolling action to the granules as they move down the filter face. The rolling action would disturb all the crevice, constriction and cavern sites retaining generally larger particulate and allow reentrainment of the larger particle fractions. Because of the small percentage of grain surface in contact with adjacent granules, dislodging of adhered particles through intergranular friction should not affect surface sites retaining the smaller particles significantly. However larger particles penetrating the bed could conceivably carry finer particles affixed to their surfaces thereby reducing collection efficiency as seen in the results for size fractions smaller than 10 micron. The smaller particles would de-agglomerate and could thus be detected by the analyser used in this experiment because the analysis was conducted on a water borne sample with a sonic agitator.

When considering the MGPF as the primary separator and the broad spectrum of particle sizes needed to be collected (especially the size fraction greater than 30 micron) the action of sieving and straining has been shown to be very important. Further work to improve efficiencies should therefore concentrate on bulk movement of the granular medium to minimise and if possible eliminate inter granular movement. Alterations to the panel bed design necessary to meet SA emmission regulations are discussed in the following chapter.

An alternative solution to improved collection efficiencies is inclusion of an upstream cyclone to remove the larger size fractions responsible for the poor collection efficiencies. Further work with low particle loadings in the order of 5 g/Nm_3 and particle size range less than 10 micron should be conducted.

CHAPTER 6

INDUSTRIAL SCALE FILTER DESIGN

6.1 SCALE UP OF MGPF EXPERIMENTAL RESULTS

The filtration experiments were designed to minimise scale up considerations in extrapolation of the filter design to an industrial size unit. The experimental rig design incorporated full scale louvre spacing in a panel of dimensions 2 by .5 metres as this was envisaged to be a practical size for an industrial module filter element. Unfortunately the face area presented by this size panel was found to be too large to cover the range of face velocities during experimentation. As a result portions of the panel were blanked off effectively increasing face velocities by reducing the area for flow. The blanking plates were positioned to expose only the bottom quarter of the 2 metre length. The filter length was considered to play an important role in the overall collection efficiency and it would have been a distinct advantage to use a full scale length in the experiments. No practical solution was found to avoid shortening the panel, so the results from the experiments have to be seen as an initial estimate of the panel filtration efficiency. At this stage, the design of a full scale unit is necessary only for cost estimation purposes and it is assumed that equivalent filter efficiencies at least could be achieved over the length of a 2 metre panel.

Collection efficiencies measured over the range of face velocities tested are presented graphically in Figure 22. A summary of the experimental results have been extracted from the data and shown below. Included are PF boiler and combined cycle requirements.

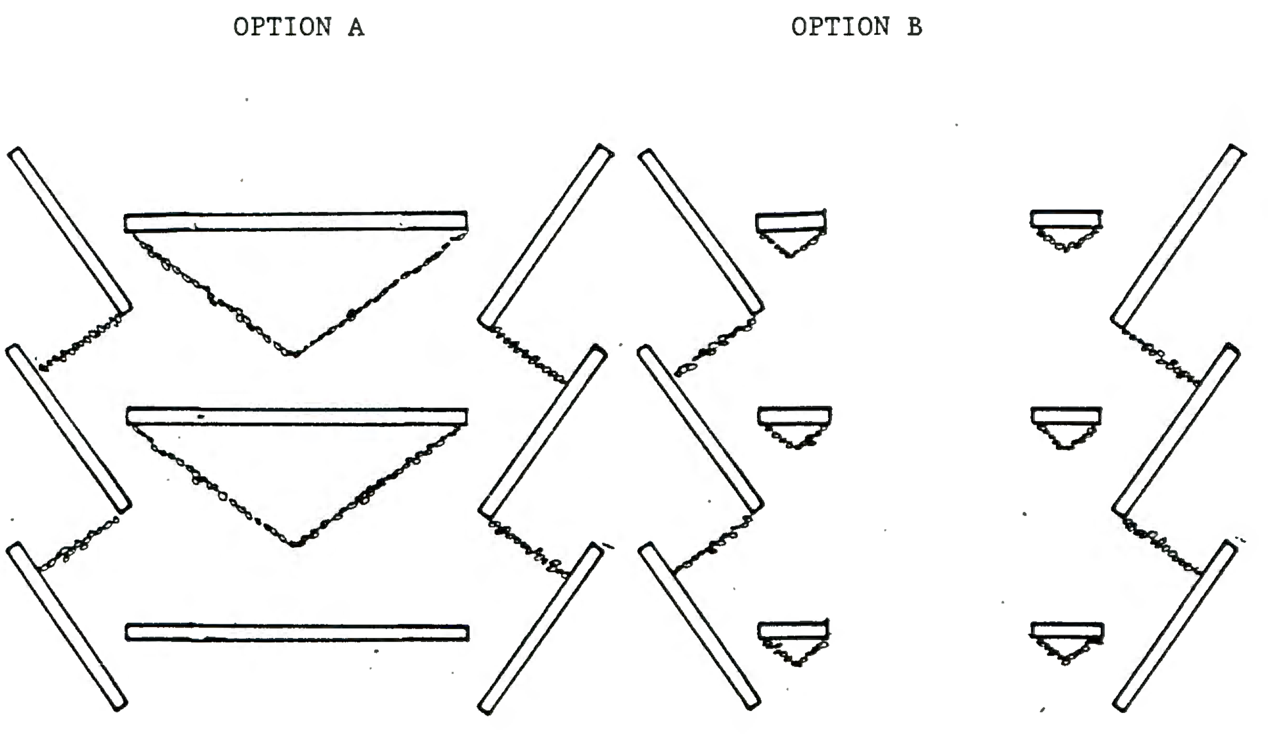
TABLE 5: SELECTED FILTER EFFICIENCY REQUIREMENTS

	TEMP	INLET LOADING	OUTLET LOADING	EFFICIENCY
	C	g/Nm3	g/Nm3	%
MGPF	320	32	3.8	89
PF combustion	132	33.5	0.05	99.85
Combined Cycle	850	28.4	0.3	99
Combined Cycle	1000	3.7 *	0.004	99.89

* Inlet loading after cyclonic cleanup

The operation of the thin granular panel, as tested, is shown to be insufficient to meet either the PF or CC requirements. As a result, in the design of an industrial scale filter unit, a longer path length through the filter medium was required entailing a thicker filter panel if the same size granules were to be used. In view of the low pressure drop measured during experimentation a thicker filter bed was seen to be easily affordable from pressure drop considerations. Increasing the thickness however presented some difficulties. Two possible options for increasing the panel thickness were considered capable of maintaining the vertical alignment. These two are shown below.

FIGURE 29: THICKER PANEL DESIGN OPTIONS



Option A however would not increase the flow path length because of the short path created via the annulus beneath the baffle. Option B, although its configuration differs somewhat to the original experimental design, was considered the solution to a thicker filter bed. The design has not been tested, but for the purposes of a cost estimate this configuration is considered feasible.

The average fluid path length through the granular bed was 41 mm. This figure was backed up by pressure drop measurements and thus assumed to be reasonably accurate. (See Chapter 5). Based on this path length at 14 cm/s face velocity, improving the filter efficiency from 89% to 99.85% (required for PF cleanup) necessitates a fluid path length of roughly 120 mm. The calculations assumed the unit cell collection efficiency to be unaffected by an increase in fluid path length and assuming that the filter was operated in an almost clean state, this assumption would be reasonable. The fluid path length required would be calculated using the following:

$$Z = l \frac{\ln(1-E)}{\ln(1-\eta)} \tag{27}$$

6.2 MECHANICAL DESIGN CONSIDERATIONS

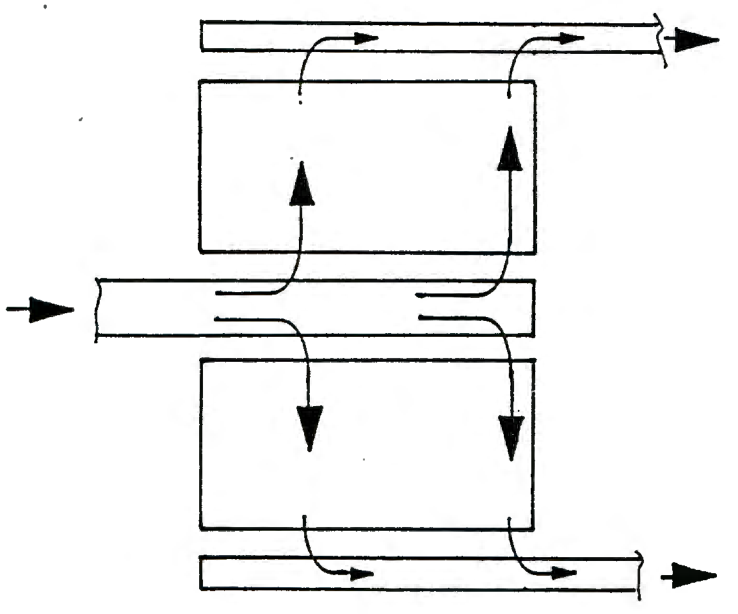
6.2.1 FILTER EFFICIENCY

Two industrial scale filter sizes, 10 and 600 MWe, were considered in the design calculations. Taking note of the presumptive limits included in the SA emission regulations (in Chapter 2) it will be noted that the air pollution control regulations are lenient in comparison with American and European regulations. The regulations also become less restrictive at smaller scale. For the purposes of this project the filter design was adjusted to meet the more restrictive limits applicable to the larger scale users. This allows one module design to be developed which is suitable for all the size ranges. The large installations would be made up of a collection of modules operating in parallel.

6.2.2 PHYSICAL SIZE CONSTRAINTS

As an initial estimation the floor area available for a filter unit would be of the same order of magnitude as an equivalent baghouse filter. von Bulow (20) reports large scale baghouse units to measure roughly 60 by 30 meters of which two units are needed per 600 MWe boiler. The same floor plan was seen to be attractive if retrofitting of existing baghouse facilities is considered and because the similarity in size would simplify future large scale power plant design. The housing and ducting would be laid out as shown below.

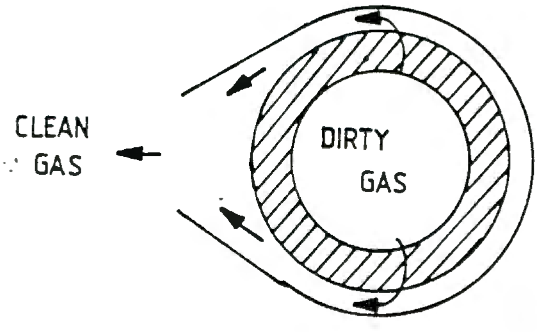
FIGURE 30: FILTER HOUSING AND DUCTING LAYOUT

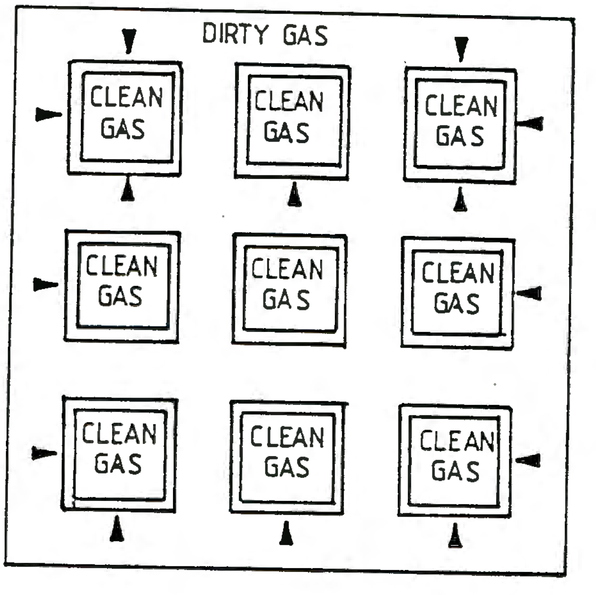


The layout of the panels within the floor space was limited by the restriction of the panel height to 2 meters. Lee et al (36) proposed a commercial design from his experimental work comprising a collection of tall rectangular panels making up square elements within a module as shown below. The Granulate-Tube-Filter and Electroscrubber use one cylindrical panel per module also shown below.

FIGURE 31: PLAN VIEW OF INTERNAL PANEL LAYOUT

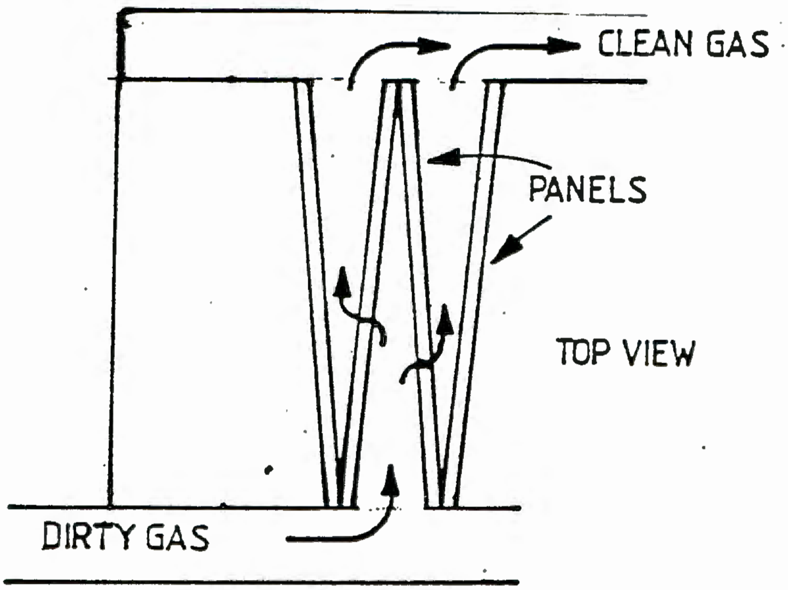
ELECTROSCRUBBER (TM) SQUIRES PANEL FILTER





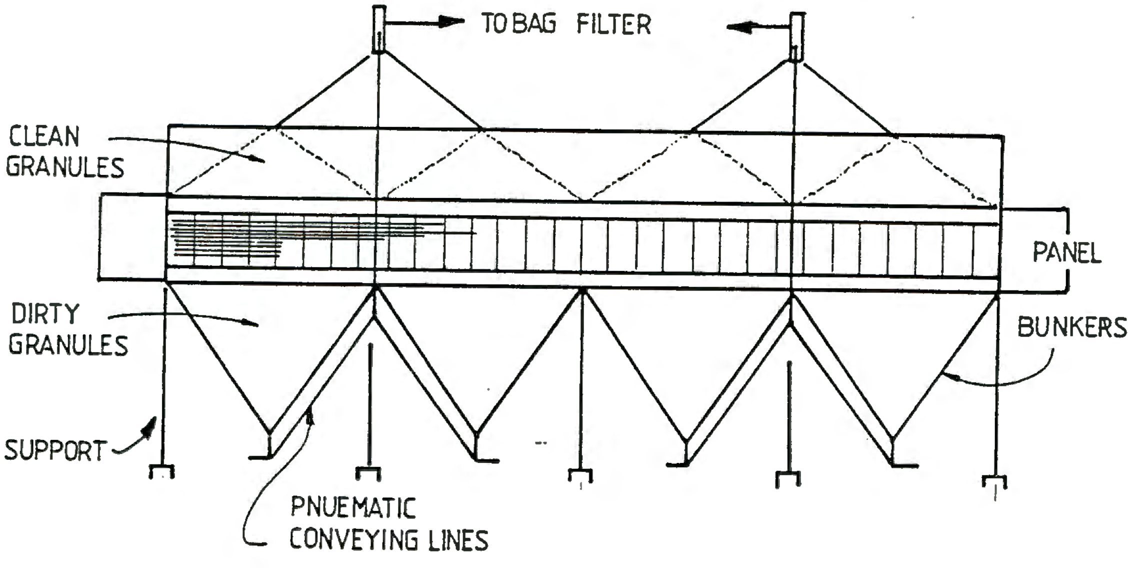
Both these designs have the advantage of being able to use much longer panel lengths. Limiting the panel height to 2 meters precludes these two options as the associated floor space requirements would be too large. The final panel layout design is as shown below.

FIGURE 32: PLAN VIEW OF PROPOSED INTERNAL PANEL LAYOUT



A cross-section of the filter housing is illustrated below. The cross-section illustrates the basis for solid flow control. The storage area immediately above the panel distributes the filter medium to all the panel sections. Flow through the panels is regulated by the solid withdrawal rate from the base of each bunker and is controlled by pneumatic transport of filter medium away from the base of the bunker outlet doing away with any control valves. The central airlift pipe ducts the pneumatic air flow carrying the dirty granules to a de-engagement section above the unit. Transport of the granules is reported to be sufficient to dislodge the accumulated particulate from the granule surface. (11) The same cleaning operation is assumed to be sufficient in this design with final cleanup carried out by bag filtration of the transport air in an auxiliary baghouse filter.

FIGURE 33: CROSS-SECTIONAL VIEW OF PROPOSED LARGE SCALE DESIGN



Collectively the filter medium flow from five panels is controlled by a set of four bunkers located beneath the panels. The design of these bunkers has to ensure mass flow of the granular media to maintain equivalent media flow rate down each panel length. One pneumatic air blower would be dedicated to two bunkers to provide the motive force for recirculation of the dirty granular material to the top of the unit. The pneumatic air flow would be ducted to a de-engagement vessel by one vertical 254 mm (10″) line. Each blower and pneumatic recirculation duct would have a dedicated de-entrainment vessel from which the clean sand would return to the upper storage section by gravity flow.

The preliminary design was developed to enable a realistic cost estimate to be calculated. A final economic assessment is presented in the following chapter.

CHAPTER 7

ECONOMIC ASSESSMENT OF THE MOVING GRANULAR PANEL FILTER (MGPF)

As part of a first order assessment, an estimate of the economics of the design is important to determine whether to pursue investigations any further. Although the filter did not demonstrate the ability to meet the filtration requirement in its present form, cost estimates for an improved design can be drawn from the results. At this stage the combined cycle filter requirements are not well understood so filtration costs for conventional power plants only are considered. Assessment is based on meeting S A emission regulations so wherever possible cost data is based on local conditions. The technical assessment of the MGPF was restricted to an analysis of FBC flyash filtration, although it is anticipated that the filter design would find application in other combustion techniques. Different combustion techniques produce varying flyash quantities and characteristics as well as volumes of flue gas. As an initial estimate however, the cost of a MGPF will be inferred from the FBC data collected in this investigation.

7.1 COAL BASED FLUE GAS CLEANUP TECHNIQUES

Coal combustion is used extensively in South Africa in steam raising installations at both small industrial and large utility scale. Coupled to combustion, is the inevitable production of pollutants both solid and gaseous which are released to the atmosphere. Quantities released are subject to control by regulations enforced by a recognised authority in an effort to preserve atmospheric air quality. The choice of cleanup technology is related to firstly the duty required, secondly, the effect of the fly ash characteristics on filter efficiency and thirdly, the cost of each filter option capable of doing the job. With several different combustion techniques being employed, fly ash characteristics and quantities vary. In addition, emission regulations vary from country to country impacting on the flue gas cleanup efficiency requirements. The efficiency of different filter systems varies considerably, and, in some cases, preclude the use of a given design for a particular duty.

Of the variety of filter designs available for gaseous filtration, electrostatic precipitation (ESP), bag house filtration (BH) and cyclone cleanup are the primary techniques used on existing coal fired installations. ESP and BH filtration are generally used where high filtration efficiencies are required. The choice between the two usually depends on the cost of each system but in some cases certain physical conditions may preclude the use of one or the other.

ESP is particularly suited to removal of particles from gas streams that can be easily ionised. (37) The particles are subjected to a high voltage discharge setup between the discharge electrode and a grounded collection surface and then migrate and adhere to the collection surface. Intermittent rapping of the collection surface dislodges the collected particulate into bunkers below. The accumulation of particulate on the collector surface has a significant effect on the electrostatic field and capture efficiency. Fly ash characteristics thus play an important role in the capability of ESP.

The cost of the basic ESP is a function of collection plate area which itself is dependant on the collection efficiency required. Larger plate areas are required for higher collection efficiencies. Capital costs of ESP are also more attractive in large scale units but have been shown to be more expensive in small scale applications. (42)

Baghouse filters are best suited to removal of particulate from dry gas streams or from gases where the water remains in the gaseous state. The bags are made of a range of fabric materials which have inherently high filtration efficiencies with the result that costs are relatively independent of the filtration efficiency up to a point. Capital costs are not affected by scale to the same extent as ESP and tend to vary linearly with increases in volumetric flow. (42) Costs are however increased with higher operating temperatures as more expensive cloth is required to with stand the higher temperatures. (38) Replacement of fabric used in the filter bags is a major drawback because this is usually necessary every 3-6 years (8) and needs to be taken into account when comparing the economics of the two filter options. BH filters are also limited to an operating temperature of less than 300 degrees Celsius.

The cyclone is simply a conical vessel within which a circular flow is induced. Particle inertia forces the particles to the vessel walls from where they are collected. The particle capture technique is suited to removal of heavier particles, typically larger than 20-30 micron. (40) Of the three options, cyclonic cleanup is the cheapest technique but has relatively low filtration efficiency which is related to inlet gas velocity and cyclone dimensions.

The economic viability of each filter type does not only depend on the capital cost of the unit. Power and operational costs are also significant when considered over the expected life of the unit. ESP units are reported to have reasonably low operating and maintenance costs (37) and with the low pressure drop across the unit the fan power costs are also low. The electrostatic field although operating with high voltages draws very little current and does not affect power costs much. As already mentioned BH maintenance costs are high because of the need to replace the filter bags periodically. Fabric filtration also has characteristically high pressure drop which incurs high fan power costs. Cyclones generally have low maintenance costs but they suffer higher pressure drops than ESP. These factors have significant bearing on the choice of filter system and must be considered in any cost comparison.

7.2 SOUTH AFRICAN COAL FIRED INSTALLATIONS

In SA, ESP is used on large coal fired power plants because it is the most economical option capable of meeting the current emission regulations. The economic advantages of ESP over BH, at this scale, are highlighted by a recent cost comparison by ESCOM, (20) the major electrical power producer in SA. The results show capital costs of ESP to be 65% less than those for a baghouse unit (on a 600 MWe unit scale) capable of reducing the effluent concentration to 50 mg/actual m^3. However it was speculated by von Bulow (20) that in the case of emission regulation of only 15 mg/actual m^3 particulate, BH would be favoured economically over ESP. To substantiate this one only has to look at the costs of the same size unit in the USA where the emission regulations require dust effluent levels as low as 20 mg/actual m^3. Shannon (16) reports ESP to be between 13-33% more expensive than the equivalent baghouse at the 600 MWe scale, highlighting the effect of emission regulations on filter costs. We can therefore, only expect to see ESP being replaced by baghouse filtration in SA, if emission regulations are revised along the lines of the those in Europe and the USA.

The assessment of MGPF costs has been made for both large and small scale applications to determine the effect of equipment size. Two alternative granular panel filter designs have also been costed and included in the report. These two filter systems are not specifically designed for fly ash removal but technically should be suitable.

The assessment is split into two case studies based on plant electrical rating. Case 1 covers the large scale based on a plant rating of 600 MWe. This rating is typical of the size used by ESCOM in new coal-fired power plants. Case 2 is based on a 10 MWe unit, typical of small scale industrial boilers (especially FBC systems eg Babcock -(Fluidburn)) units. Estimates of plant capital costs are calculated for the two different filter systems. The final economic evaluation includes auxiliary power costs and operation/maintenance costs as these have significant bearing on the final filtration costs.

7.3 BASIS FOR COST ESTIMATION OF EACH FILTER TYPE

7.3.1 MOVING GRANULAR PANEL BED

A full scale filter design was necessary to calculate a realistic cost estimate. The scale up from experimental size, and the development of the mechanical design details for both the large and small scale are covered in the previous chapter. The scale up required a substantially increased filter area to accommodate the large volumes of flue gas and a thicker filter panel to improve the efficiency. The panel bed design had to be altered slightly to allow for a thicker panel. The preliminary design of the full scale units is sufficient only to estimate quantities of materials required. This level of design detail enabled sheet and structural steel requirements to be estimated. The size and quantity of auxiliary equipment required was also estimated in conjunction with experimental data. Exact figures are listed in Appendix C and D attached.

The calculation of fabrication and erection costs was carried out along lines suggested by a local boiler manufacturer. Typical material wastage and additional stiffening metal requirements as well as realistic material and labour costs were derived from consultation with the same boiler manufacturer. The cost data sheets in the appendices include the figures used to arrive at the total installed capital cost. Additional information regarding filter installation costs was taken from a series of articles by Vatavuk (37,38,39,40) relating to this type of estimate.

The costing procedure is as follows :

1) calculate the weight of sheet and structural steel

2) make allowance for metal wastage associated with cutting up of standard lengths supplied by the steel industry.

3) make allowances for stiffening of large areas of sheet metal used in the filter housing and bunkers.

4) with these allowances, calculate a gross metal weight required.

5) calculate metal purchase price

6) calculate labour time and rates based on metal weight of structure

7) sum labour costs and metal price to estimate purchase price

8) calculate the selling price by multiplying the purchase price by a final cost factor which includes contractor overheads and profit margin.

9) where sub contracting is envisaged the final cost factor is replaced by a project management factor.

The figures shown are in 1986 terms.

7.3.2 BAGHOUSE/ESP

CASE 1: 600 MWe

Real cost estimates for both bag filters and ESP units suitable for a S A, 600 MWe power plant installation are available. These estimates were carried out in 1983 by ESCOM, who were considering gas cleanup options for a new power plant. Values have been escalated to reflect 1986 costs. Escalation factors were calculated using the SA consumer price indexes quoted in International Financial Statistics. (41)

CASE 2: 10 MWe

No cost data was available for an ESP unit of such small scale. Price scale-up correlations (37) were used to estimate the ESP costs at the 10 MWe scale using the 600 MWe costs as a basis. It is realised that the large difference in the two sizes will introduce some error in the scaling down required to estimate the price of the small ESP. However, the scale ratio of 60:1 falls within the range tolerated by Peters and Timmerhaus. (42)

Costs of the small scale baghouse filter were estimated from a small unit installed on a FBC unit housed at the SA National Institute for Coal Research. Scale up correlations based on cloth area (37) have been used. The costs are directly proportional to cloth area for a given face velocity and hence also proportional to volume flow, so the costs could be scaled up directly. Escalation of the costs to 1986 was also necessary as original prices were quoted in 1984 Rands.

7.3.3 OTHER GRANULAR PANEL FILTERS OPTIONS

Two alternative cross flow granular filters designs were included in the assessment. These two filters are currently being marketed in the USA and Europe under the trade names of Electro Scrubber TM and Granulate-Tube-Filter TM. Cost data for the Electro Scrubber was obtained from an installation operating in South Africa at ISCOR's sinter plant in Vanderbijl Park (11). The unit was originally designed by the Combustion Power Company (CPC) in the USA. Costs of the Granulate-Tube-Filter were supplied by Lurgi in the form of a budget quote. All costs were escalated to 1986 rands.

It should be noted that the two filter designs here have different operating principles. The Electro Scrubber TM is a moving granular bed using large gravel filter medium augmented by electrostatic field. The Granulate-Tube-Filter TM is a fixed granular bed using very fine granular material requiring intermittent reverse flow cleaning much like a filter bag system.

7.4 EQUIPMENT COSTS

A summary of the estimated costs for both the small and large scale - filter requirements are listed in Tables 6 and 7. The results are applicable to the filter requirements for pulverised fuel combustion and FBC systems. The costs quoted are broken down into basic capital cost, auxiliary and fan power costs and operation and maintenance costs at 1986 present value based on a plant life of 30 years and nett discount factor of 6%.

The cost breakdown of each of the filter options is shown graphically thereafter in bar graph format:

TABLE 6: SUMMARY OF COSTS FOR THE 600 MWe CASE

		BAGHOUSE	ELECTRO SCRUBBER	GRANULATE-. TUBE	ESP	MOVING GRANULAR BED
SIZE	M	60-60-30	60-10-30	26-38-24		60-60-15
FACE AREA	M2	60000	2400	1600	200000	9000
FACE VEL	M/sec	0.021	0.52	0.78	1.37	0.14
PRESS DROP	mmH20	150	200	145	57	55
BASIC COST	MILLION	11.25	12	54	5.25	10.9
FAN UPRATING COST	MILLION	1.5	2.25	1.5	0	0
ESCALATED COST	MILLION	18.6	20.8	63.4	7.7	10.9
AUXILIARY POWER COSTS	MILLION	5.7	43.3	5.7	5.7	24.3
FAN POWER COSTS	MILLION	36.5	48.7	35.3	13.9	13.3
OPERATION/ MAINTENANCE COSTS	MILLION	34	LOW	LOW	LOW	LOW

TABLE 7: SUMMARY OF COSTS FOR THE 10 MWe CASE

		BAGHOUSE	ELECTRO SCRUBBER	GRANULATE- TUBE	ESP	MOVING GRANULAR BED
SIZE	M	5-5-12	4-3-25	3-6-8		15-4-15
FACE AREA	M2	1000	40	27	3500	150
FACE VEL	M/sec	0.021	0.52	0.78	1.37	0.14
PRESS DROP	mmH20	150	200	145	57	55
BASIC COST	THOUSANDS	134	200	900	9	185
ESCALATED COST	THOUSANDS	197	290	1000	767	185
AUXILIARY POWER COSTS	THOUSANDS	95	720	95	95	405
FAN POWER COSTS	THOUSANDS	608	812	588	230	220
OPERATION/ MAINTENANCE COSTS	THOUSANDS	567	LOW	LOW	LOW	LOW

FIGURE 34: COST SUMMARY 600MWe

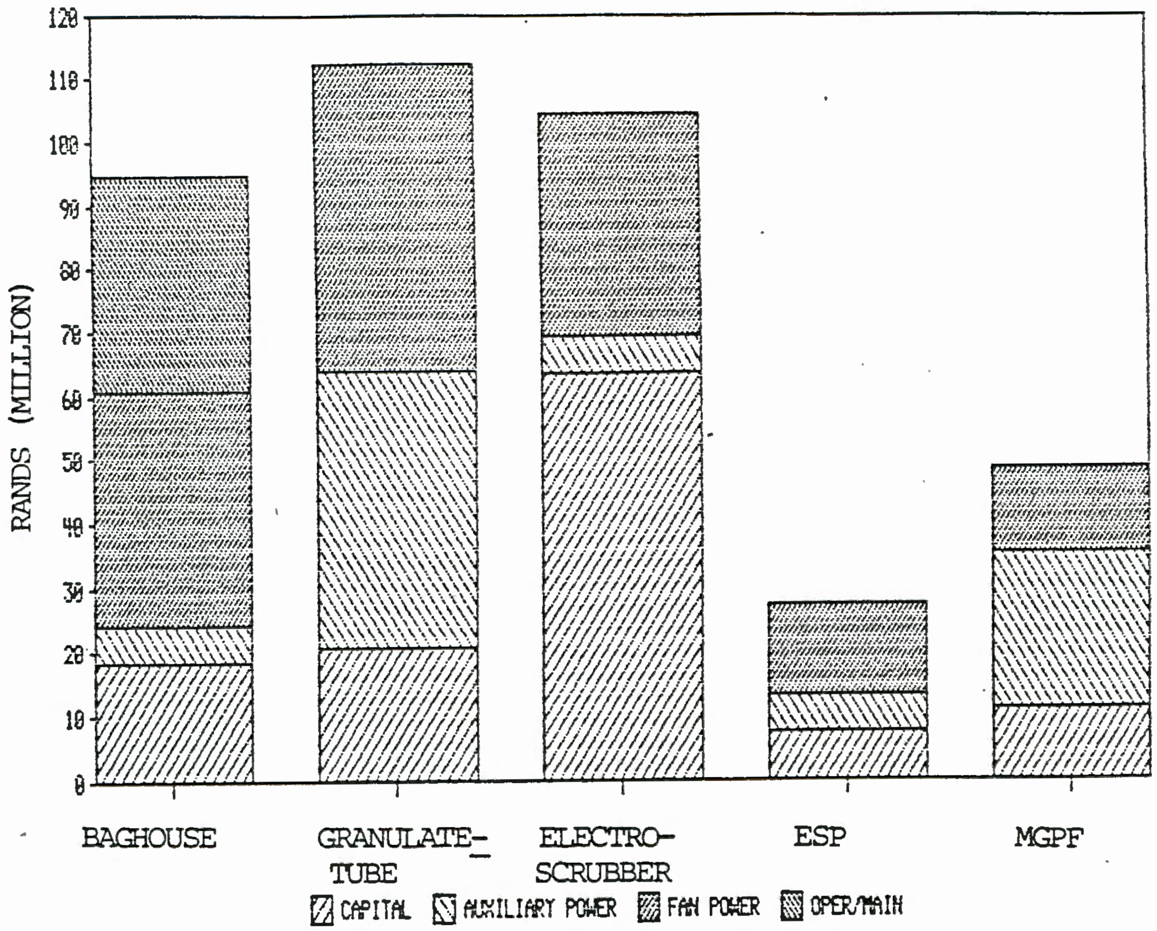
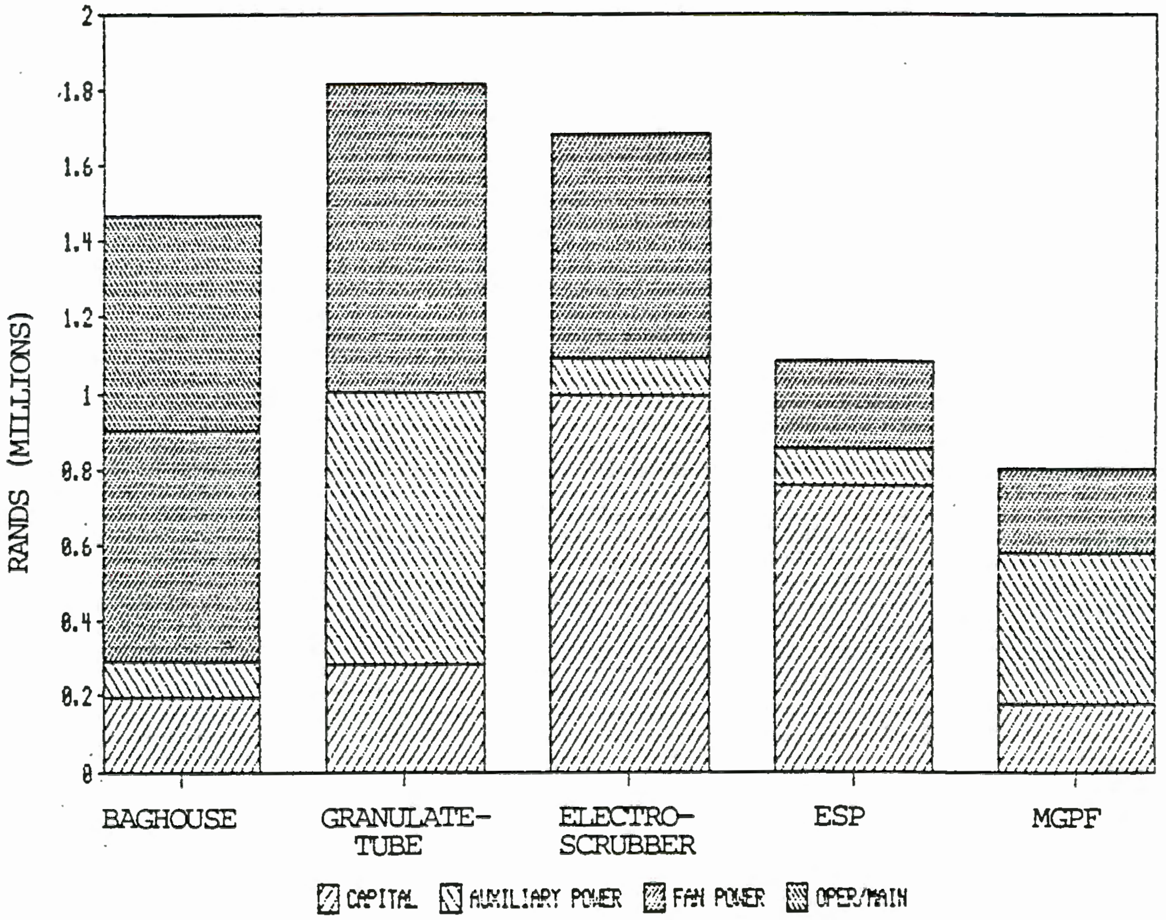


FIGURE 35: COST SUMMARY 10MWe



Operating and maintenance cost data were available only for the baghouse filter. Specific costs for the other options were not available but references were made by Lurgi and CPC to the effect that operation and maintenance costs on both the Electroscrubber TM and the Granulate-Tube-Filter TM were low. Baghouse filter maintenance involves cost of replacing filter bags every 3-6 years, whereas moving granular filters need only replace sections in the pneumatic transport piping eroded by gravel attrition. Elsewhere gravel flow rates are too low to cause any erosion.

7.5 ECONOMIC EVALUATION

The economic assessment of the filter options was found to differ for the two scales involved. ESP was found to be the most economic system with the MGPF lying between the ESP and Baghouse filter costs for the large scale unit (600 MW). The MGPF however, was found to be the most economic option for the small scale unit (10 MWe).

Only ESP costs are found to be affected by filter size. All other filter systems are based on modular format and so cost scale up is essentially linear with size. Costs associated with ESP do not scale up as simply. This explains why costs of ESP are more favourable in the large scale unit and less favourable at the small scale. This trend is consistent with costs quoted generally for dust collector systems.(42) Correlations show baghouse being cheaper than ESP up to a capacity of 140 m^3/sec equivalent to 70 MWe, after which ESP becomes the less expensive option of the two.

Capital costs of the filter units did not dominate the overall economic analysis. Significant contribution was found to come from the fan power costs associated with the pressure drop across the filter unit. In the moving granular filter option a further significant cost was the power required to convey the granular material back to the top of the filter. The conveying power costs are dependant on the dust loading to the filter and the solid/dust ratio required for operation. Typically the Electroscrubber(TM) uses a ratio of 130 kg of filter medium for every 1kg of dust. (11) The MGPF was successfully operated at 30:1 but costs are based on a design figure of 80:1 to allow for any long term effects. The higher the solid/dust ratio the higher the circulation rate of the filter medium required.

A reduction in the inlet dust loading, to reduce recirculation flow requirements, could be achieved by installing a cyclone upstream of the filter. Cost analysis shows that any savings in recirculation flow power requirements would be offset by the additional fan power costs associated with the increased pressure drop resulting from the cyclone. However, the optimum configuration was not investigated but should be included in any further studies.

Auxiliary power costs for the remaining filter options were not as significant as the moving granular filters. The Granulate-Tube-Filter is not a moving granular bed so auxiliary power costs were not as high.

The very high capital costs of the Granulate-Tube-Filter are not completely understood but could be related to the complications of intermittent cleaning cycles with large automated isolation valves in the mainstream.

Although, operational and maintenance costs were, for most filter options, incomplete their effect would only really be seen in the baghouse costs. Granular filters and ESP are typically low maintenance units but the baghouse filter needs bag replacement every 3-6 years. Nonhebel (43) quotes operation and maintenance costs of baghouse filters to be between 5-9 times that of an equivalent ESP. Thus, in the comparison of ESP, baghouse and the MGPF, inclusion of these operational and maintenance costs would only make the baghouse option less favourable in both the large and small scale units.

The cost estimation of the MGPF, as it included knowledge of only major capital items, is considered to be accurate to +/- 30%. This degree of uncertainty should always be included in such cost comparisons. The estimation of the costs of the MGPF design indicate that it could be the most economic flue gas filter option for small scale coal fired installations. Considering the fact that little or no optimisation of the MGPF design has been attempted it is possible that the granular bed filter could compete economically with ESP at the large scale as well. Reduction of auxiliary power costs by optimisation of the solid recirculation rates or the inclusion of an upstream cyclone is almost certainly possible and could bring overall costs down to within the ESP price range.

However, in realistic terms the MGPF would first have to be proven at the small scale size before any consideration to its replacing the ESP or baghouse units would be given. The track record of the ESP and baghouse filters are well known making it very difficult for a new concept to be accepted.

CHAPTER 8

CONCLUSIONS

This work was undertaken in an effort to assess the suitability of a specific granular panel bed filter for flue gas cleanup at high temperature. The panel filter configuration is demonstrated to be capable of processing hot flue gas streams with attractively low pressure drop. Although the continuous filter operation offers some advantages, granular movement was found to have several disadvantages. The following conclusions can be drawn from the project:

1) With low particle loading in the filter medium, filtration efficiency of the MGPF was found to be poor and lower than filtration efficiencies in static beds for the same operating conditions. The reduction in collection efficiency was found to be more pronounced for particles greater than 10 microns in size. The exact mechanism responsible for the reduced efficiency is not well understood but believed to be the result of a combination of voidage increase, particle bouncing and particle re-entrainment. Pressure drop analysis showed voidage to increase during granular movement but this, on its own, was not considered sufficient to explain the poor efficiencies achieved. In the range of high Stokes numbers seen in this experiment, particle bouncing off the collector surface and re-entrainment effects contribute to the significant reduction in collection efficiency. A correlation based on the experimental results for FBC flyash shows similar trends to those depicted in the literature for static beds, identifying the same fall off in efficiency for particles of high kinetic energy.

2) Granular flow patterns in this panel design are particularly unsuited to filtering of particles greater than 10 micron in diameter. The flow pattern enhanced intergranular movement by

introducing a rolling action to individual granules, thereby disturbing crevice sites retaining generally larger size particles ie larger than 10 - 30 micron. Re-entrainment due to intergranular movement was considered the most significant contribution to the reduction of filter efficiency. Straining mechanisms were also affected by intergranular movement, allowing larger particle sizes to penetrate the filter bed. However the flow conditions within the granular bed were designed to keep the filter surface renewing continuously so the rolling action was unavoidable in this particular design. The MGPF would therefore not function well as a primary separation device.

3)Granular flow patterns were found to be significantly altered when the gas stream was forced through the panel. The flow on the back face of the panel was found to stop, as gas velocities were increased. Over the gas velocity range suited to this design, granular flow on the rear face could be maintained by angling the panel off the vertical. The inclination required for optimal flow conditions was a function of the gas stream velocity through the panel. Sensitivity of the granular flow patterns to the gas stream is seen as a major draw back to this design.

4)Face velocity was found to be limited by fluidisation of the granular medium in the downstream louvres. Fluidisation of the granular medium in the downstream louvres was found to be influenced by louvre angle. It was thus possible to define the failure point of the filter in terms of minimum fluidisation and louvre angle.

5)The collection efficiency of the MGPF was found to deteriorate progressively from initial clean bed conditions. The initial collection efficiency can therefore not be used as a conservative design tool for the MGPF.

6)Improvements to the filter's capability to that required of an industrial filter have been considered. Essentially a thicker panel has been recommended using the same louvre configuration to retain a renewing filter face. Improvements to the design include a mass flow section within the filter bed where intergranular movement would be minimised. It is believed that further work with this design should show better results if mass flow conditions can be achieved in the mid section.

7)An economic assessment of this filter concept show it to be comparable with available flue gas cleanup techniques used on existing coal fired installations.

The first order assessment of this design show the MGPF to be inadequate for primary flue gas cleanup. However, in view of the results achieved so far and the potential for improving collection efficiency with the recommended modifications further investigation into this filter design concept is warranted.

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APPENDIX A : MOVING GRANULAR TEST RESULTS

TEST	FLUID VEL	FACE VEL	IN TEMP	IN DUST	SETTLED	DELTA P	EFFLUENT	TOTAL EFF	PANEL EFF	SAND FLOW
	m/sec	m/sec	C	g/Nm	g/Nm	mmH O	g/Nm	%	%	g/min
D:05:1	0.73	0.093	235			10	4.54			500
D:05:2	0.73	0.095	246			11.5	4.47			500
D:05:3	0.86	0.118	279	30.61	10.11	15.5	3.76	87.7	81.6	400
D:07:2	0.75	0.094	216			9	3.96			700
D:07:3	0.71	0.089	223			9	4.32			800
D:07:4	0.75	0.094	239			10	5.3			800
D:07:5	0.76	0.098	262			10	5.54	80.9	71.8	800
D:07:6	0.76	0.101	279			9.5	4.79	83.5	75.6	800
D:07:7	0.74	0.105	305		9.7	10.5	4.57	84.3	76.4	800
D:07:8	0.75	0.109	322	28.99	11.63	12.5	4.4	84.8	74.6	500
D:07:9	0.75	0.111	337			13	4.89	83.1	71.8	400
D:07:10	0.75	0.111	337			13	4.94	83	71.5	400
D:08:1	0.74	0.096	253	12.2	6.29	11	0.17	98.6	97.1	STATIC
D:08:2	0.75	0.097	258			9	2.95			500
D:08:3	0.76	0.103	291	36.7	9.8	9.5	3.62	90.1	86.5	600
D:08:4	0.76	0.109	325	30.4	9.2	11	3.64	88	83.1	600
D:08:5			328		8.81					
D:09:1	0.97	0.13	298	41.75	10.54	19	4.63	88.9	85.1	400
D:10:2	0.45	0.049	183	21.3	13.4	4	0.67		91.5	400
D:10:3	0.46	0.052	204	24	13.3	4.5	1.2	95	88.8	300
D:10:4	0.46	0.055	238			4.5	2.18	90.9	79.6	200
D:11:1	0.74	0.093	244	35.6	12.7	9.5	4.24	88.1	81.4	800
D:11:2	0.75	0.105	310	32.4	8.76	11	4.6	85.8	80.6	1000
D:11:3	0.75	0.106	318	29.2	9.86	11.5	4.52	84.5	76.7	1000
D:12:1	0.92	0.125	265	19.8	3.7	17	0.26	98.7	98.4	STATIC
D:14:2	0.47	0.065	234	12.4	8.81	5.7	0.68	94.5	81.2	400
D:14:3	0.54	0.069	258	53.4	13.8	6.5	1.89	96.5	95.2	400
D:14:4	0.54	0.07	279	36.9	13.8	6.5	2.34	93.6	89.8	500
D:14:5	0.54	0.073	291			6.8	2.26	93.9	90.2	500
D:15:2	0.94	0.13	283		8.85					
D:15:3	0.96	0.14	339		7.38					
D:15:4	0.96	0.14	339	27	6.3	14.5	3.93	85.4	81	900
D:15:5	0.96	0.15	376	39.9	8.43	19.5	3.79	90.5	88	900
D:15:6	0.96	0.15	376			19.5	2.84	92.9	90.9	900

APPENDIX B: VERTICAL PANEL FILTER TEST RESULTS

TEST	FLUID VEL m/sec	FACE VEL m/sec	IN TEMP C	IN CONC g/Nm	OUT CONC g/Nm	DELTA P mmH O	EFFICIENCY %	SAND FLOW g/min
1:02	1.00	0.04	304	90.3	1.80	22.5	98.0	STATIC
1:06	0.39	0.01	183	14.3	0.21	9.7	98.6	400
3:02	0.75	0.05	236	12.6	0.33	18.1	97.4	900
4:07	1.00	0.08	414	45.2	1.57	24.3	96.5	600
4:08	0.75	0.10	286	19.4	0.05	19.2	99.7	500

PRIMARY ITEMS

NAME	SIZE kg	WASTE FACTOR -	STIFFENING FACTOR -	MASS REQUIRED kg	PRICE R/kg	MATERIAL COST Rands	LABOUR TIME hours/ton	TIME REQUIRED hours	COST Rands	PURCHASE PRICE Rands	SELLING FACTOR -	SELLING PRICE Rands
LOUVRES	8559.50	1.07	1.00	9158.67	0.71	6502.65	100.00	855.95	14379.96	20882.61	1.43	29832.30
LOUVRE SUPPORT	490.50	1.03	1.00	505.22	0.71	358.70	100.00	49.05	824.04	1182.74	1.43	1689.63
LOUVRE FRAMES	2280.00	1.10	1.00	2508.00	0.78	1956.24	50.00	114.00	1915.20	3871.44	1.43	5530.63
FILTER HOUSING	3274.00	1.05	1.35	4583.60	0.71	3254.36	100.00	327.40	5500.32	8754.68	1.43	12506.68
BUNKERS	1541.00	1.20	1.35	2388.55	0.71	1695.87	100.00	154.10	2588.88	4284.75	1.43	6121.07
SUPPORT STUCTURE	1024.00	1.10	1.00	1126.40	0.78	878.59	50.00	51.20	860.16	1738.75	1.43	2483.93
SUB TOTAL	17169.00			20270.43		14646.41		1551.70	26068.56	40714.97		58164.25

AUXILIARY

NAME	QUANTITY	SIZE	UNITS	UNIT COST Rands	COST Rands	COST Rands	PURCHASE PRICE Rands	SELLING FACTOR	SELLING PRICE Rands
BLOWERS	1	67.00	kWatt	12000.00	12000.00		12000.00	1.15	13800.00
PIPING	6.00	152 mm	meters	49.00	294.00	294.00	588.00	1.15	676.20
	46.00	254 mm	meters	136.00	6256.00	6256.00	12512.00	1.15	14388.80
	3.00	356 mm	meters	210.00	630.00	630.00	1260.00	1.15	1449.00
	1.00	508 mm	meters	350.00	350.00	350.00	700.00	1.15	805.00
BAGHOUSE FILTER	1.00	2.30	M3/sec	14500.00	14500.00	1060.00	15560.00	1.15	17894.00
SUB TOTAL					34030.00	8590.00	42620.00		49013.00
INSTRUMENTATION	allow 5% of equipment purchase price						4166.75	1.15	4791.76
TOTAL	(purchase price)						87501.72		111969.01

INSTALLATION

NAME		PURCHASE PRICE Rands	SELLING FACTOR	SELLING PRICE Rands
ERECTION	allow 20% of equipment purchase price	17500.34	1.43	25025.49
INSULATION	151.00 M2 rquired @ R100/M2	15100.00	1.15	17365.00
ELECTRICALS	allow 8% of equipment purchase price	7000.14	1.15	8050.16
CIVILS	allow 6% of equipment purchase price	5250.10	1.15	6037.62
SAND	380.00 tons required @ 30/ton	11400.00	1.15	13110.00
CONTINGENCY	allow 3% of equipment purchase price	2625.05	1.15	3018.81
SUB TOTAL		58875.64		72607.08
TOTAL PROJECT COST				184576.09

PRIMARY ITEMS

NAME	SIZE kg	WASTE FACTOR -	STIFFENING FACTOR -	MASS REQUIRED kg	PRICE R/kg	MATERIAL COST Rand	LABOUR TIME hours/ton	TIME REQUIRED hours	COST Rands	PURCHASE PRICE	FINAL COST FACTOR -	SELLING PRICE Rands
LOUVRES	513572.00	1.07	1.00	549522.04	0.71	390160.65	100.00	51357.20	862800.96	1252961.61	1.43	1789945.15
LOUVRE SUPPORT	29428.00	1.03	1.00	30310.84	0.71	21520.70	100.00	2942.80	49439.04	70959.74	1.43	101371.05
LOUVRE FRAMES	136800.00	1.10	1.00	150480.00	0.78	117374.40	50.00	6840.00	114912.00	232286.40	1.43	331837.71
FILTER HOUSING	196445.00	1.05	1.35	275023.00	0.71	195266.33	100.00	19644.50	330027.60	525293.93	1.43	750419.90
BUNKERS	92481.00	1.20	1.35	143345.55	0.71	101775.34	100.00	9248.10	155368.08	257143.42	1.43	367347.74
SUPPORT STUCTURE	61440.00	1.10	1.00	67584.00	0.78	52715.52	50.00	3072.00	51609.60	104325.12	1.43	149035.89
SUB TOTAL	1030166.00			1216265.43		878812.94		93104.60	1564157.28	2442970.22		3489957.45

AUXILIARY

NAME	QUANTITY	SIZE	UNITS	UNIT COST Rands	COST Rands				
BLOWERS	60	67.00	kWatt	12000.00	720000.00		720000.00	1.15	828000.00
PIPING	720.00	152 mm	meters	49.00	35280.00	35280.00	70560.00	1.15	81144.00
	2280.00	254 mm	meters	136.00	310080.00	310080.00	620160.00	1.15	713184.00
	180.00	356 mm	meters	210.00	37800.00	37800.00	75600.00	1.15	86940.00
	60.00	508 mm	meters	350.00	21000.00	21000.00	42000.00	1.15	48300.00
BAGHOUSE FILTER	2	70.00	M3/sec	422000.00	844000.00	62000.00	906000.00	1.15	1041900.00
SUB TOTAL					1968160.00	466160.00	2434320.00		2799468.00
INSTRUMENTATION	allow 5% of equipment purchase price						243864.51	1.15	280444.19
TOTAL	(purchase price)						5121154.73		6569869.64

INSTALLATION

ERECTION	allow 20% of equipment purchase price	1024230.95	1.43	1464650.25
INSULATION	9058.00 M2 rquired @ R100/M2	905800.00	1.15	1041670.00
ELECTRICALS	allow 8% of equipment purchase price	409692.38	1.15	471146.23
CIVILS	allow 6% of equipment purchase price	307269.28	1.15	353359.68
SAND	22838.00 tons required @ 30/ton	685140.00	1.15	787911.00
CONTINGENCY	allow 3% of equipment purchase price	153634.64	1.15	176679.84
SUB TOTAL		3485767.25		4295417.00
TOTAL PROJECT COST				10865286.64

APPENDIX E

INDUSTRIAL SAND AND ENGINEERING COMPANY LTD.

P.O. BOX 19 PHILIPPI CAPE

SAND SPECIFICATIONS

TYPE: 16/30 FILTER

DATE 26 SEPTEMBER 197

EFFECTIVE SIZE: 0,62 mm COEFFICIENT OF UNIFORMITY 1,47

LOOSE BULK DENSITY: 1624 Kg/m³

TYPICAL GRADING ANALYSIS

U.S. MESH	APER (mm)	% RETAINED
4	4,76	
5	4,00	
6	3,36	0
7	2,83	0,1
8	2,38	0,2
10	2,00	0,4
12	1,68	0,4
14	1,41	0,9
16	1,19	5,0
18	1,00	16,0
20	0,84	27,8
25	0,71	24,9
30	0,59	16,5
35	0,50	7,1
40	0,42	0,6
45	0,35	0,1
50	0,297	0
60	0,25	
70	0,21	
80	0,177	
100	0,149	
120	0,125	
140	0,105	
200	0,074	
-200		
OTAL		100,0

TYPICAL CHEMICAL ANALYSIS

	%
SiO_2	99,56
Al_2O_3	0,24
Fe_2O_3	0,019
TiO_2	0,025
ZrO_2	0,006
CaO	0,003
MgO	TRACES
L.O.I.	0,15

SABS - CKS 305 - 1971

GRAIN FINENESS NO: 16

CLAY CONTENT:

UNIFORMITY OF GRAIN SIZE:

TOTAL FINES:

SINTERING POINT:

MARKS:

2 6 JUN 1989

GINATOR: D.W. NIXON

CERTIFIED CORRECT : C.B.PA

LYST: T. DILJEE

AUTHORISED :