Hydrodynamics and Heat Transfer in a Gas-Fluidised Bed Bioreactor

by

Patricia Evans

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ADDENDUM

In section 4.2.2 (page 158) the thermal conductivity of Puffed Wheat has been wrongly quoted as 0.26 W m\(^{-1}\) K\(^{-1}\). This value has been used in the subsequent calculations of the particle-convective heat transfer coefficient, and in section 4.6.2 (page 186), to calculate the Biot number.

A more accurate value is 0.045 W m\(^{-1}\) K\(^{-1}\), this is a mean of values for cork board, sawdust, fibre insulating board and insulating wall board taken from Perry (1984). This reflects the porosity of the particles and the effect of the included air on the particle thermal properties.

a) Effect on Particle-Convective Heat Transfer Coefficient.

When this value is substituted into equations 4.11 through to 4.16 the results are as follows:

\[
\begin{align*}
  k^0_e &= 0.0354 \text{ W m}^{-1}\text{K}^{-1} & \text{equation 4.13} \\
  k &= 0.497 \text{ W m}^{-1}\text{K}^{-1} & \text{equation 4.12} \\
  h_p &= 190 \text{ W m}^{-2}\text{K}^{-1} & \text{equation 4.11} \\
  h_f &= 18.14 \text{ W m}^{-1}\text{K}^{-1} & \text{equation 4.15} \\
  h &= 16.56 \text{ W m}^{-2}\text{K}^{-1} & \text{equation 4.16}
\end{align*}
\]

The particle convective component is not significantly altered by the change in the particle thermal conductivity. This is consistent with the comments made in the main text, that the film component is the main contributor to the particle convective heat transfer coefficient.

b) Effect on Biot Number

The Biot number is calculated as follows:

\[
\text{Bi} = \frac{h_{ext} d}{k_p} \quad (4.25)
\]

therefore, using the values quoted in section 4.6.2, \(\text{Bi} = 11.4\)

This revised value for the Biot number is greater than the ratio of the particle and fluid heat capacities, therefore internal resistance to heat or mass transfer is dominant. This is consistent with the experimental results discussed later in chapter 4.

Reference:

Abstract
Solid-state fermentation has been superseded for most industrial applications by submerged culture; however, for some cultures it has a number of advantages, particularly when considering the downstream processing stages.

Gas-phase-continuous three-phase fluidisation has been used for solid-state fermentation applications at the pilot-plant scale, and provides a number of advantages over submerged culture. However, particles composed of a mixture of organisms and substrate are both adhesive and cohesive, and tend to agglomerate. In addition, the moisture content required for optimum fermentation frequently approaches or exceeds the wet quenching point for the fluidised bed. These problems also occur in drying applications where the material is cohesive or has a high initial moisture content.

In this work the restrictions on three-phase fluidisation were assessed with particular reference to the proposed use of a gas-fluidised bed as a fermenter. A modified bed design for three-phase gas-continuous fluidisation was designed and tested using non-fermenting systems. In this design the wet material was mobilised by introducing air tangentially into the bed via jets in the wall as well as through the distributor, to prevent the formation of agglomerates. The test bed was cylindrical and 145 mm in diameter. The behaviour of the bed, the air required to mobilise the bed contents and the interaction of the jets with the bed material were modelled analytically.

In addition, during fermentation, metabolic heat has to be removed from the culture. This occurs by evaporation from the particles, and by removal of sensible heat through the wall of the bed. Further experiments to determine the rates of heat transfer between the gas and the wall, and the particle drying rates were performed in a 225 mm diameter bed.
PREFACE
This work was done at the University of Surrey between 1986 and 1990. The work was jointly funded by the SERC and ICI PLC, as a Spinks project.
Hydrodynamics and Heat Transfer in a Gas-Fluidised Bed Bioreactor

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1.1 Solid Substrate Fermentation

The use of microorganisms, such as bacteria, fungi and other single celled organisms, goes back to the beginnings of history. Naturally occurring microorganisms have been traditionally used both to produce alcohol and in the preparation of bread and other foods. Before the existence of such small organisms was known, it was observed by their effects that the action of some organisms on foods, the process known as fermentation, could in some cases be beneficial, in contrast to the decay and spoilage caused by other, harmful organisms. Fermentation can improve keeping properties, add variety to a diet and make food more palatable; in addition some fermentation processes improve food values by releasing indigestible nutrients. Many of the traditional fermentation processes - bread, koji and beer for example - are based on solid substrates.

Solid substrate fermentation can be defined as any fermentation in which the substrate is not a free liquid (Aidoo et al, 1982). This is a relatively broad definition, which includes processes with very little free water, such as mushroom culture and silage production, as well as systems where the solids are suspended in liquid, such as the production of monoclonal antibodies by immobilised cells. The range of processes also includes traditional fermented foods, industrial processes for enzymes and other microbial products, microbial recovery of minerals and treatment of waste water. As the organisms and techniques used for fermentations with a low water content differ fundamentally from those where the solids are floating or submerged, it is convenient to consider these two groups separately.

1.1.1 Low Water Solid Substrate Fermentation Processes

Low water fermentation in general has been recently reviewed by Aidoo et al (1982) and Tengerdy (1985), while some of the technical problems have been discussed by Lonsane et al (1985). Indigenous fermented foods, which form a large subject by
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themselves, have been considered by Steinkraus (1983) and Hesseltine (1983).

The earliest controlled fermentation processes were bread-making, which is recorded in Egyptian tombs, cheese processes for preserving milk products, and koji, which is an oriental process which produces an enzyme preparation. The koji product is used as a starter material for a wide variety of fermented food processes, which often require further fermentation steps, one of the most important products being soy sauce. The earliest records of koji-type fermentation in China date back 2500 years (Aidoo et al 1982). Fermentation is also used for processing meat, fish and vegetables into a huge range of other foods, particularly in the Orient, but also around the world.

More recently, solid-substrate fermentation processes were adapted to produce chemical products. The first commercial use of enzymes, other than directly for food products, was vinegar (acetic acid) production, which was recognised as a microbiological process in the 19th century. In the vinegar generator, wine is allowed to trickle down a column packed with porous material in which a mixed population of micro-organisms has become established, and vinegar is rapidly produced. Also in the 19th century, Van Tieghem (1867) established the role of Aspergillus niger in the production of gallic acid, used in the tanning and printing industries. A process was used in which the mould was grown in heaps of tannin-containing substances which were occasionally stirred, and the gallic acid was washed out after about a month. Modern processes for both vinegar and gallic acid are based on submerged microbial culture systems. The enzyme industry expanded in the early 20th century, with production of a wider range of fungal and microbial enzymes. Then, in the 1940s, the discovery of antibiotics fuelled a further expansion of the industry. Many of these first products were produced using solid substrate processes, including penicillin.
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Historically it seems to be a pattern that the earliest processes used solid substrate methods, and these have then been superseded by submerged culture; this being the case with all modern antibiotic cultures for example. Food processes are often an exception and this may be for a number of reasons: the products are often solids, with a distinctive texture produced in the process; tradition may be important to a product's appeal, and it is usually not necessary to extract the product as a pure chemical from the fermentation process stream.

There are still some modern biological synthesis processes which preferentially use solid substrate methods. One example is the production of fungal toxins, of which the most well known are the aflatoxins. Hesseltine (1972), in developing a process to produce aflatoxins for toxicity tests, compared submerged culture with agitated solid substrate cultures and obtained significantly higher yields (a maximum of 1 g per kg substrate) in the latter case. It appears that the production of the desired product is dependent on physiological conditions which do not arise in submerged culture. Large scale production has been carried out by Silman et al (1979) in an aerated corn storage bin with a helical stirrer.

More recently, it has been discovered that fungal spores have a high biological activity, and they have been used as catalysts for the transformation of organic compounds, such as steroids, antibiotics, fatty acids and carbohydrates. The advantages of producing spores by solid state fermentation rather than submerged culture have been reported to include simplicity, higher yields and a pure and homogeneous spore product (Vézina and Singh 1975).

Solid substrate fermentation has also been used as a low cost process to treat waste materials, in order to recover the energy content or reduce the disposal problems. Waste reduction is
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becoming more important as sites for waste disposal, particularly landfill sites for domestic waste, are under increasing pressure. In this type of application, solid state processes, which are low in capital cost and running costs, are much cheaper than equivalent submerged culture methods and create less waste, often because there is much less liquid to dispose of at the end of the process. One common application is the upgrading of plant waste for use as animal feed: microorganisms break down indigestible materials such as cellulose to sugars and increase the protein content of the material, particularly when nitrogen is added in the form of fertilisers or urea. Animal wastes have also been processed to recover the nitrogen and carbohydrate content for feed. Biological methods for treating farm wastes are reviewed by Loehr et al (1977). Municipal waste can also be biologically treated by composting, which reduces the toxicity and odour of waste before it is dumped or put into landfill. A number of systems have been developed in different countries.

1.1.2 Organisms for Solid Substrate Fermentations

Tengerdy (1985), describes some of the many types of organisms which grow on solid substrates. A wide range of organisms can be cultured in moist solids but only filamentous fungi grow in the absence of free water; single cell organisms, such as yeasts and bacteria, always require a surface film of water to remain active. Lactic acid bacteria commonly occur in preservation processes in which they increase the acidity of the culture. Thermophilic bacteria are active in composting processes in which the high temperatures generated are essential to kill off many pathogens. Yeasts are often found in mixed cultures in food preparation, growing symbiotically with other organisms. Filamentous fungi are used in food preparation, in the koji process among others, and are important for waste treatment, as they produce many powerful enzymes which break down substrates such as cellulose and lignin. Rather than being dispersed in a film of surface water, they spread in a branching, root-like network of filaments called
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Hyphae, over the surface of the substrate, using the available liquid as a source of moisture and nutrients. As the fungus spreads the growing hyphae penetrate cracks and pores to seek out further nutrients. The most important industrial microorganisms are those filamentous fungi which grow naturally on fruits and grains. In many traditional food processes the culture is not grown aseptically, and the operators have little or no knowledge of microbiology. It is therefore not surprising that the types of fungi used are those that grow first in the succession of moulds growing on plant material in nature. These include Rhizopus, Mucor, Amylomyces, Aspergillus, Monascus and Neurospora. The main groups of yeasts found are Saccharomyces; Candida and Saccharomycopsis (Hesseltine 1972).

1.1.3 Types of Fermenter
Many solid substrate fermenters are simple in construction, based on a relatively small unit size, and labour-intensive to operate. Lonsane et al (1985) review a variety of wooden boxes, covered pans, cells and trays, used in both laboratory scale and commercial production. Scale-up is difficult, as there is a limit to the depth of culture which can be grown if the conditions in the fermenting mass need to be monitored and controlled. For this reason geometric scale-up of the individual units has often been unsuccessful and industrial production is often achieved by loading many small trays onto racks or conveyors and processing them in environmentally-controlled tunnels or rooms.

The largest plants have been built for koji production: these plants are semi-automated and the fermenting solids are kept in trays up to 5 x 12 m in area, in humidity- and temperature-controlled rooms; mechanical stirrers are automatically triggered by sensors. However, the depth of culture is still limited to 30-40 cm because of the local build up of heat (Hesseltine 1983). Heat produced by the respiration processes raises the temperature above the optimum, reducing the efficiency of the process and
eventually killing the organisms. Other large commercial units include bins and drum fermenters: continuously stirred bins up to six metres in diameter have been used to produce mycotoxins, while small rotating drum fermenters holding up to 70kg have been used by various workers. However it is also hard to scale up drum fermenters, as in larger units the fungal mycelium tends to be broken up by the rotation, and hence the growth of the culture is slow. There is also a large unused volume in a drum fermenter. The low-cost, high-bulk processes use the most basic technology: for example, in waste treatment processes and some mushroom culture, the material is inoculated, piled in heaps and left to mature under a plastic sheet. No attempt to control the process is made apart from limiting the size of the heaps to allow some heat to dissipate from the surface.

1.1.4 Control of Fermentation Parameters

Only a limited number of fermentation parameters can be controlled in a solid substrate fermentation, and the control of individual nutrient components is particularly difficult. The controllable parameters (temperature, moisture and aeration), are often inter-dependent; for example, humidity and hence drying rates are a function of temperature. However in many cases moisture and temperature can be controlled independently, as is shown below.

Moisture

Lonsane et al (1985) have discussed the monitoring and control of solid substrate fermentations. One of the most critical parameters is the amount of water in the fermenting mass: both high and low extremes can be a problem. At low water levels release of nutrients is reduced and growth slows. When excess water is present the air spaces in the material fill up and the oxygen transfer rate is reduced, the rates of growth and product production can be affected. With fungal cultures there is a risk of bacterial contamination, as any bacteria present in the culture are normally inhibited by the lack of free moisture. Moisture loss
can be controlled by varying the humidity of the air, or by adding sterile water or a solution of nutrients.

Temperature
Temperature control is also important: metabolic heat is generated throughout the culture and has to be dissipated rapidly to keep the temperature in the optimal range, usually 25-32 °C for most fermentations. The amount of heat generation may be as high as 155,000 kJ/kg dry solids for compost (Lonsane et al 1985), or 2600 kJ/kg for koji type fermentation (Aidoo et al 1982). Methods for control include keeping the fermenter in a water bath or temperature-controlled room, circulating cooling water in pipes through the fermenting mass, and forcing excess air through the material to promote evaporative cooling.

pH
The pH of solid substrate fermentations is not usually monitored because there is no easy way to do this, although it is an important factor. Conveniently, many substrates have good buffering characteristics which reduce the fluctuations in pH, and the addition of pH-adjusted solutions to moisten the substrate and the use of urea rather than ammonium salts to provide nitrogen are measures which can be used to exploit this characteristic. Local pH variations as the organism grows over the surface of the solids are best controlled by agitation.

Aeration
Most fermentations are aerated by forcing sterile air through the fermenting mass. The optimum gas flow rate through the fermenter is linked to a variety of parameters including oxygen supply, moisture levels and the dissipation of carbon dioxide and heat. Carbon dioxide concentrations as high as 21% (by volume) have been reported in the bottom of a 6.35 cm bran bed (Lonsane et al 1985). The concentrations of carbon dioxide or oxygen can determine the rate of production of certain metabolites, so they
can be optimised to promote the desired product. As most solid substrate fermentations are highly aerobic, excluding composting and a minority of food processes, maintaining good oxygen transfer rates is essential. Problems with oxygen transfer rates may arise from excessive moisture filling the interparticle spaces, agglomeration, or dense microbial growth blocking the transfer of oxygen into the particles. Oxygen transfer is improved by using loosely packed coarse particles, using shallow layers of substrate, shaking or agitating the substrate or rotating the fermenter. If heat is being removed by evaporating water in the air stream, the optimum moisture levels can be maintained by the addition of water or nutrient solutions. The rate of drying is controlled by varying the humidity of the air.

Advantages of Solid Substrate Fermentation
Solid substrate fermentation has a number of advantages over submerged culture. Firstly, it reproduces the conditions in which the organisms grow in nature, allowing morphological differentiation appropriate to the production of spores and expression of certain metabolites. Secondly it is intensive, reactor volumes can be small, and the nutrients and products are concentrated. Less water has to be removed, and in some instances the product can be recovered directly from the solid mass by solvent extraction, therefore the cost of separating the product after fermentation may be less. Thirdly, it is very suitable for aerobic organisms as the diffusion path between the air stream and the organism is relatively short. Fourthly, and particularly for non-sterile processes, the capital costs are much less, as the fermenter vessels are smaller and do not have to support the weight of large quantities of water. Finally, operating costs may be lower: in some cases air compression costs are lower, and spores can be used to seed fermenters, so large volumes of seed inoculum in the form of suspended cells do not have to be generated and handled.
Disadvantages of Solid Substrate Fermentation

Not all commercially useful organisms can be grown in this way, for example, *Escherichia coli*, mammalian cells and plant cells require consistent conditions which are only produced in a submerged culture. The types of organism which can be cultivated on a solid substrate are limited to those organisms which can withstand the variable and relatively harsh process conditions, a classification which includes the filamentous fungi, some yeasts, some bacteria and the streptomycetes. The fermentation parameters tend to be difficult to control, especially pH, but also temperature and the moisture content. The rate of heat generation limits the working depth of some processes. Substrate utilisation is incomplete, as the branching of the hyphae bypasses a certain proportion of the surface area, and a large spore inoculum is needed to get good distribution of the organism throughout the substrate. Finally the substrate may need pre-treatment to release more of the nutrients, by such processes as pearling, dehulling, cracking, steaming or acid hydrolysis.

1.1.5 Submerged Solid Substrate Fermentation Processes

Nearly every organism or microbial product produced commercially is grown in a submerged culture system, i.e. in a tank in which the cells are freely suspended in the culture medium; nutrients are supplied as solutes in the liquid phase. As has been observed above, submerged culture is currently used for many processes which were originally developed as solid substrate fermentations. The system owes its success to the ease with which process parameters such as temperature and pH can be controlled, while oxygen and nutrients can be delivered effectively in solution to all the cells, producing a more consistent product. Fermentation equipment is based on processing liquids rather than solids, and similar vessels can be applied to different organisms and readily scaled-up for increased production. However in some aspects solid state culture continues to out-perform the equivalent submerged culture. This is seen in the degree of process intensification,
in terms of the higher cell densities per unit volume which are obtained in solid materials, while soluble products are often present in higher concentrations in the liquid phase, with a concomitant reduction in the cost of extracting a pure product from the culture medium. There is also scope to develop a culture system for the large-scale commercial exploitation of those groups of organism which do not grow readily in submerged culture, such as plant cells. In the search for a way to combine the benefits of solid-state culture with the controllability and consistency of submerged culture, systems in which the cells are grown on a suspension of solids in a liquid culture medium have been developed.

Such immobilised cells have been applied to a wide variety of biological systems, both to improve the productivity of standard submerged cultures and to overcome some of the problems which have prevented the commercial exploitation of certain specific organisms. The range of processes which have been developed and the engineering aspects of immobilised cell systems have been reviewed in depth by Webb, Black and Atkinson (1986).

Process systems which demonstrate the advantages of immobilisation include both simple systems such as waste water treatment, where the cells and suspended solids are held together in flocs by microbial polysaccharides, and sophisticated technologies such as the encapsulation of animal cells in hollow polymer spheres to protect them from shear stresses. Black et al (1984) have classified immobilisation processes into 'active immobilisation', where the cells are incorporated into the supports in a preparation step before being put into the reactor, and 'passive immobilisation', in which the supports and the inoculum of cells or spores are put into the same reactor and the cells then grow preferentially in or on the supports until the majority of cells are supported.
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1.1.6  Types of Support Particle

Many different types of supports have been proposed and tested and it is important to note that the support need not be a source of nutrients if these are adequately supplied in the liquid medium. In this case inert supports have a number of advantages, as they need not interfere with the process conditions when these have already been optimised to suit the specific needs of an organism; for example, they will not affect the control of the pH by buffering the solution. In addition they are less likely to fragment or dissolve away during the fermentation, and hence the particle size distribution does not vary a great deal over the course of the fermentation, which is important if the particles are maintained in a fluidised state. Inert particles also do not need to be replenished as often as consumable particles. As inert particles are not consumed during the process, the particle size and size distribution can be fixed to suit the process, so that liquid fluidised systems for example can operate with a narrow, fixed size distribution, which is chosen to suit the throughput required. The particle size can also be decreased to reduce intra-particle mass transfer resistances if these are limiting the productivity. Due to these advantages inert supports are widely used for immobilised cell processes. Types of support include natural materials such as celite (Gwebonyo and Wang 1983) and vermiculite (Bland et al 1982), which are both porous minerals noted for their large surface area, and carageenan gel which is derived from seaweed, (Behie et al 1987). Synthetic supports include cubes of polyvinyl and polyurethane foams (Kirkpatrick and Palmer 1987, Hall and Rao 1988), and plastic or steel meshes compressed into spheres (Black et al 1984).

In submerged culture any soluble products are produced as very dilute solutions which leads to high capital costs associated with the large volumes of liquid which have to be processed, and high product recovery costs when the product has to be recovered from solution. Insoluble products are also produced in low bulk
concentrations, even when highly concentrated in individual cells, because of the low cell density. Hence biological syntheses do not compete with chemical synthesis except for very complex molecules or very high value products. Any process intensification which increases the final concentration of the desired product in the reactor, either in solution or as an increased cell density if it remains associated with the cells, can give significant improvements in the economics of a process.

Several workers have increased the cell loading or product concentration by using immobilised cells rather than submerged culture. Cooper (1986) has described the use of support particles to intensify the treatment of waste-water. The throughput of water in the activated sludge process is normally limited by the settling velocity of spontaneously formed flocs of solids and micro-organisms. The addition of supports made from compressed stainless steel mesh or reticulated polyester foam provides nuclei for floc formation. The large stable flocs thus formed have a higher settling velocity and a narrower size distribution, which reduces the settling volume required and allows a higher flow rate of water through the unit. The use of steel meshes, in particular, significantly increases the relative density of the flocs compared to water.

1.1.7 Antibiotic Production

Immobilised cells have been used by a number of workers for antibiotic production, Behie et al (1987), Gwebonyo and Wang (1983) and Arcuri et al (1983) among others. Gwebonyo and Wang (1983) immobilised Penicillium chrysogenum on celite beads in a bubble column reactor and obtained a cell density almost twice as high as the cell density of cells in suspension under similar process conditions; the final penicillin concentration was eleven times higher in the immobilised culture. They ascribe this improvement to the lower viscosity in the immobilised culture, and a concomitant increase in the rate of oxygen transfer to the
cells, as at high cell concentrations the submerged culture is very viscous and non-newtonian in behaviour, which reduces the mass-transfer coefficients. They found that the oxygen mass transfer rate in the immobilised culture was close to the oxygen demand of the cells, giving higher rates of growth and metabolite production, while in the submerged culture the cells were partially starved. As the viscosity is lowered by confining the culture to beads, the energy required to aerate the culture is lowered compared to submerged culture, which is also an advantage.

These results are encouraging but it is more difficult to predict the impact of microbead systems on commercial penicillin production, as in the vigorously agitated systems which are used commercially, the celite beads are likely to be disrupted by the shear stresses around the agitator, or degraded by attrition.

Behie et al (1987) discuss the use of immobilised penicillium moulds for the continuous production of two antibiotics, penicillin-G and patulin. The work had two aims: to increase the intensity of the process by producing more product per unit reactor volume, and to develop a continuous process which would be competitive with the batch and fed batch processes used commercially. In their process they used two support matrices, carageenan gel and celite beads, in a three phase fluidised bed. The nutrient medium formed the continuous phase and was pumped up through the reactor. The celite beads supported a higher cell density than carageenan, but in neither case was the cell growth limited by low oxygen mass transfer coefficients, as the liquid viscosity remained low. The cells were successfully grown continuously for more than ten days, and a control system was developed to maintain optimum conditions in the reactor. In particular it was observed that the production of penicillin-G by *Penicillium chrysogenum* is linked to cell growth, while the production of patulin by *Penicillium urticae* is at a maximum when the cells are suffering from nitrogen starvation. Thus the
control system could be set up in a different way for different organisms; for example, by regularly removing some cells and providing a high level of nutrients, the majority of the *P. chrysogenum* cells can be kept in the exponential growth phase. Therefore a continuous system can be developed to produce high levels of antibiotics over a long period.

The ability to maintain cells for long periods with constant growth conditions has important implications in a number of antibiotic syntheses. Bushell (1988) has shown that in streptomycetes the secondary metabolites, which include the desired antibiotic product, are not produced simultaneously but as a succession, in response to changes in nutrient availability. As many antibiotics are secondary metabolites, it follows that in batch culture the desired product may only be produced transiently, and even in fed batch the period of production may be short and less than that which would be achieved in continuous culture, if the appropriate conditions could be maintained.

Continuous culture has been successfully used by Arcuri et al (1983) to produce thienamycin, a broad spectrum β-lactam antibiotic, which is extremely susceptible to degradation by medium components and some of the other metabolites produced during fermentation. The work is based on a study by Baker (1982) who showed that immobilised cells will continue to produce thienamycin for as long as 180 days. In this system *Streptomyces cattleya* is immobilised on celite beads which are perfused with nutrient medium. By using a perfusion system in which the immobilised cells remain in the reactor and the medium is recycled, the medium can be periodically replaced to maintain the supply of essential nutrients as they are used up and remove any harmful metabolites, and hence the production of thienamycin can be extended.

Bland et al (1982) have used the bacterium *Zymomonas mobilis* to
produce ethanol in a continuous process. The bacterium was immobilised on vermiculite to increase the density of biomass in the reactor and to hold the bacterium in the reactor at high dilution rates. It was shown that a mixture of flocs and immobilised cells formed in the reactor, and that high rates of ethanol production, up to 105 g l⁻¹ hr⁻¹, could be obtained, without requiring filters or centrifuging to return cells to the reactor. This can be compared to ethanol productivities of 120 g l⁻¹ hr⁻¹ and 82 g l⁻¹ hr⁻¹ for submerged culture of Z. mobilis with recycle by membrane filtration and vacuum fermentation of Saccharomyces cerevisiae with cell recycle respectively.

1.1.8 Novel Products from Immobilised Cells

Hall and Rao (1988) and Kirkpatrick and Palmer (1987) have used immobilisation to produce fungal enzymes. The white rot fungus, Phanerochaete chrysosporium, was immobilised in cubes of polyurethane foam and cultured semi-continuously by periodically replacing the medium and adding veratryl alcohol which induces ligninase production. The results show that the production of ligninase was higher from the immobilised cells than for control cultures with no foam supports, while the variation between flasks was small, giving more consistent results. It was also shown that the culture of immobilised cells was stable and could be stored at 4°C and then reactivated to give similar enzyme yields.

Shi and Hall (1988) have also successfully immobilised Anabaena azollae, a symbiotic cyanobacterium which is usually found inside the leaves of the ferns of the genus Azolla. When it is grown in the pores of polyurethane foam particles Anabaena establishes the same morphology as it has in the leaves of Azolla and the immobilised cells can be used to generate hydrogen gas or ammonia.

1.1.9 Conclusions

Solid state fermentation could be used to improve a number of
fermentations currently performed in submerged culture, if the process parameters, particularly in low water fermentations, could be more closely controlled in a large scale process.
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1.2 Fluidisation

One of the methods which has been developed for maintaining constant conditions in an aerated bed of particles is fluidisation. When an increasing flow of air is percolated up through a packed bed of particulate solids, there is a point at which the drag force exerted by the air becomes equal to the weight of the particles. At this point the particles are no longer supported by the inter-particle stress, and the bulk properties of the bed undergo a step change; the bed is then said to be fluidised. In a fluidised bed the material flows and mixes like a fluid, and the efficient mixing, combined with the heat capacity of the solid particles also leads to high heat transfer coefficients both in the bed and between the bed and the walls of the container. This in turn leads to well mixed isothermal conditions throughout the bed. The air flow required to fluidise a bed of particles can be predicted, if the weight per unit area is known, by using the Ergun equation, which relates the pressure drop across a packed bed to the flow rate through it. A bed of particles with negligible inter-particle and particle-wall forces will fluidise when the pressure drop across it is equal to the weight per unit area.

Ergun Equation

\[
\frac{\Delta P}{L} = 150 \frac{(1 - \varepsilon_m)^2 \mu_f U_f}{\varepsilon_m^3 (\phi_s d_p)^2} + 1.75 \frac{1 - \varepsilon_m}{\varepsilon_m^3} \frac{\rho_f U_f^2}{\phi_s d_p^2}
\]  

Minimum Fluidisation Condition

\[
\frac{\Delta P}{L} = (1 - \varepsilon_{mf})(\rho_s - \rho_f) g
\]  

where \( \Delta P/L \) is the pressure drop per unit height, \( \varepsilon_m \) is the bed voidage, \( \varepsilon_{mf} \) is the voidage at minimum fluidisation, \( \phi_s \) is the sphericity factor for the particles, \( d_p \) the particle diameter,
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\[ \rho_p \text{ the particle density, } \rho_f \text{ the fluid density, } \mu_f \text{ is the fluid viscosity, } U_f \text{ the fluid superficial velocity and } g \text{ the acceleration due to gravity.} \]

From the very earliest work on fluidisation, variations in the fluidised behaviour of different sized powders were noticed, particularly if the particles were very small (i.e. less than 50 µm). The most widely accepted classification of different types of fluidisation behaviour is that proposed and described by Geldart (1986), who separates them into four broad groups, listed here in order of increasing particle size.

Group C
This group includes all particles which are in any way cohesive. These are difficult to fluidise "normally", as the air tends to pass through the bed through cracks or channels, so that the actual pressure drop at the air flow predicted to produce fluidisation is less than the weight per unit area of the bed, and may be as low as half this value. These materials exhibit few of the desirable properties of a fluidised bed.

Group A
Most commercial fluidised bed catalytic reactors use group A particles and therefore they have been the subject of a large body of research. These particles are known as "aeratable" because the bed expands smoothly and uniformly as the gas flow is increased above the minimum fluidisation velocity, \( U_{mf} \), until the minimum bubbling velocity, \( U_{mb} \), is reached, above which point any increase in the flow of air travels through the bed in the form of stable voids or bubbles. When the air supply is cut off the bed collapses slowly, at a rate comparable with the air velocity through the dense phase during fluidisation.

Group B
This group is typified by sand, and contains solids in the mean
particle diameter \((d_p)\) and density \((\rho_p)\) ranges:

\[
60 \, \mu m < d_p < 500 \, \mu m \quad \text{when} \quad \rho_p = 4000 \, \text{kg m}^{-3}
\]

\[
250 \, \mu m < d_p < 1000 \, \mu m \quad \text{when} \quad \rho_p = 1000 \, \text{kg m}^{-3}
\]

These materials differ from group A in that bubbles start to form at or only slightly above the minimum fluidisation velocity. Bubbles travel upwards at a higher velocity than the gas in the dense phase. Bed expansion is small and the bed collapses very quickly when the air supply is cut off.

**Group D**

Large and/or dense particles belong to this group. The gas velocity in the dense phase is high and gas and particle mixing are relatively poor. The bubbles travel more slowly than gas in the dense phase. As the particles are large, relatively cohesive particles can be fluidised because there are fewer inter-particle contacts. The Reynolds number for these particles at minimum fluidisation, that is: \(\rho_p U_m d_p / \mu_f\), may be high (i.e. greater than 1000), so that the flow is governed by inertia.

Geldart (1986) has shown that particles can be roughly classified into these groups on the basis of particle size and the density difference between the particles and fluidising medium, as in Fig 1.1. The boundaries between groups were originally based on observations of the fluidisation behaviour of a large range of materials, classified according to density and size; subsequent work has shown that the positions of the boundaries between groups are not only functions of density and size, but can also be affected by temperature and pressure variations and in particular by inter-particle forces. Such forces may arise from a range of effects, including liquid bridges, sintering, and magnetism.
1.2.1 Effect of Interparticle Forces on Fluidisation Behaviour

Seville (1987) has reviewed the recent advances in the field of fluidisation of cohesive materials, and discusses the role of inter-particle forces in determining the observed fluidisation behaviour. These forces have been identified as the major influence on the transition between different types of fluidisation, and therefore quantitative criteria which predict the transitions between Geldart's groups are often functions of the relative magnitude of the forces acting in the bed.

Molerus (1982) developed a criterion for the A/B and B/C group boundaries based on the ratio of the immersed weight of a single particle and the adhesion force $F_H$ at a single inter-particle contact:

$$K = \frac{\frac{1}{6} \pi d_p^3 \left( \rho_p - \rho_f \right) g}{F_H}$$

(1.3)

where $K$ is a different constant for each boundary. In order to find the transition values of $K$, Molerus calculated the magnitude of the inter-particle forces for particles known to lie on group boundaries and derived values of $K$ as follows:

- $K < 10^{-2}$ Group C
- $10^{-2} < K < 0.16$ Group A
- $K > 0.16$ Group B

More recent research has measured values of the inter-particle forces directly and made improved estimates of the values of $K$ at the boundaries, but the basis of the analysis, classifying particles according to the forces acting on them, is still supported by these researchers.
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Molerus' criterion is a good fit to Geldart's classification for hard relatively non-cohesive materials, as for these particles the relative magnitude of the drag forces and inter-particle forces (such as Van der Waals' and electrostatic effects), is a function of size and density. Hence, the inter-particle forces are small for particles in groups D and B, comparable to the drag forces for group A, and larger than the drag forces for group C type materials. In systems where other inter-particle forces are significant, Molerus' criterion can be used to predict which group properties the particles are likely to exhibit.

For the group B/D boundary, the changes in bed behaviour are more continuous; they have been described in terms of bubble size and velocity, and both types of behaviour can be observed in one bed if the bubbles grow as they travel upwards. The transition has been defined empirically in terms of the bubble velocity by Geldart (1986), so that a powder is in group D if:

\[ u_b < \frac{U_{mf}}{\epsilon_{mf}} \]  

(1.4)

where \( u_b \) is the bubble velocity and \( U_{mf} \) and \( \epsilon_{mf} \) are the minimum fluidisation velocity and the voidage at minimum fluidisation respectively.

1.2.2 The Effect of Liquid Layers on Fluidisation Characteristics

One source of significant inter-particle forces is the presence of liquid layers on the surface of the particles. Seville (1987) has reviewed the effect of liquid layers on fluidisation characteristics. Inter-particle forces due to thin liquid layers arise in a variety of systems, such as drying and cooling processes, coating, agglomeration and the cultivation of micro-organisms. The addition of liquid leads to increases in both the bed voidage at minimum fluidisation, \( \epsilon_{mf} \), and the minimum fluidisation velocity, \( U_{mf} \). This is explained by observing that
the higher inter-particle forces stabilise a more open bed structure, and the increase in voidage leads to a reduction in the resistance to fluid flow through the bed, so that a higher velocity is required to provide the pressure drop to support the bed weight.

The effect of varying the quantity of liquid present on the type of fluidisation observed has been studied by Seville and Clift (1984). They conclude that as the amount of liquid is increased, the inter-particle forces also increase, while the effect on the bed weight was negligible in their study, so that for a type B material, the fluidisation characteristics can be made to move through type A to type C, at which point the bed becomes liable to defluidise. This is compatible with Molerus' approach described above.

The effect of the liquid layer is affected by the presence or addition of fines to the bed which substantially reduces the observed changes in fluidisation behaviour. One example of this is the addition of microporous silica, which is often used as a flow conditioner to improve the fluidisation of cohesive particles. It is likely that the fines are held on the surface of the particles in the surface film, and protrude above it. The interparticle contacts are then between fines particles, with a small radius of curvature, which results in a much smaller area of contact and weaker liquid bridges, and hence lower interparticle forces.

1.2.3 Three-Phase Fluidisation
The presence of a surface film of liquid in a gas-fluidised bed of solid particles puts it into the same classification as three-phase fluidised beds. Three-phase fluidisation describes fluidised systems with more than one fluid phase, usually particles, liquid and gas or more rarely, particles and two immiscible liquids. Epstein (1981) reviewed the work in this
field and classified the different types of three-phase fluidised bed which had been studied into seven different categories. However, it is apparent from this review that fluidised beds in which the gas forms the continuous phase and the liquid is supported on the surface of the particles have received substantially less attention than systems in which the liquid is continuous and the gas is dispersed as bubbles. A recent book by Fan (1989) on gas-liquid-solid fluidisation also reveals a lack of research into the fundamental behaviour of gas-continuous three-phase fluidised beds. In this area the system most often considered is the turbulent bed contactor (TBC) in which the liquid flows downward as a surface film through a bed of inert particles fluidised in air; this has also been called counter-current fluidisation. Liquid dispersion and some aspects of mass transfer, particularly volumetric mass transfer coefficients and interfacial areas, have been correlated, but other fundamental properties, such as solids mixing and heat transfer are not reported on. The TBC has been tested for cooling tower applications but no other heat transfer behaviour has been studied.

1.3 Fluidised Bed Fermenters
One of the ways which has been suggested for improving the performance and controllability of solid substrate fermentations is to perform them in a gas-fluidised bed. The problems of heat transfer, mixing, and aeration can be addressed, and the addition of moisture and additional nutrients equally to all parts of the material is possible. The difficulties arise due to the limited understanding of three-phase fluidisation and fluidisation of cohesive particles.

The first gas-fluidised-bed fermenter was developed in Germany by Moebus et al (1978) and used for the production of ethanol and single-celled protein from Saccharomyces cerevisiae. Since then a number of other workers have studied the system, both for yeast
Moebus was interested in exploiting several of the unique features of the gas-fluidised bed. Using a bed of pelletised baker's yeast, fluidised with moist air and fed by spraying on a solution of glucose and nutrients, he was able to grow yeast with a low moisture content; high growth rates were achieved due to the high rate of mass transfer of oxygen between the air stream and the yeast pellets. The product, dried yeast for food applications, could be produced with lower separation and drying costs compared
to submerged culture processes, with the added advantage that the drying process could be carried out in the reactor vessel or a similar fluidised bed without any additional process steps.

The initial work by Moebus was followed up by Mishra et al (1982), who reported that the productivity of the system, expressed in terms of the specific growth rate of the yeast, was an order of magnitude lower than in submerged culture. However from the published data it appears that the bed material was drying out over the course of these experiments due to poor moisture control, a factor which could easily account for low productivity if part of the culture was becoming dormant.

While the system has attractive features for the production of low-value bulk products such as dried yeast or ethanol, it has also been considered for high value products such as enzymes. Both Moebus and Teuber (1984) and Bauer (1986) have investigated the production of glutathione by *Saccharomyces cerevisiae* from a substrate containing cysteine and other amino acids. The synthesis consists of the two reactions shown below.

\[
\text{GC-synthetase} \\
\text{L-glutamate} + \text{L-cysteine} \rightarrow \gamma-\text{glutamyl-cysteine}
\]

\[
\text{glutathione-synthetase} \\
\gamma-\text{glutamyl-cysteine} + \text{glycine} \rightarrow \text{glutathione}
\]

The yield of glutathione with respect to cysteine, which is the most expensive feed component, was found to be up to 48% by Moebus and up to 40% by Bauer, compared to less than 10% in submerged culture. Neither Moebus nor Bauer gives a thorough explanation for this improvement in yield, though Bauer describes the conversion of glutathione to glutathione disulphide, an unwanted byproduct which, unlike glutathione, can cross the cell membrane.
In the light of this a likely explanation is that in submerged culture the large amount of extracellular water dilutes the extracellular glutathione disulphide and it is effectively lost from the cell. Therefore the reaction does not reach its equilibrium point and glutathione continues to be degraded. The much lower water content of the solid state fermentation allows the extracellular concentration of glutathione disulphide to rise, opposing the further diffusion of glutathione disulphide across the cell membrane. The intra-cellular concentration of glutathione disulphide will also rise and this inhibits the reduction reaction and limits the conversion of glutathione and its loss from the cell.

A number of research groups in Japan have extended the work to other organisms, in particular, the fungi used in the production of soy sauce: Aspergillus sojae and Aspergillus oryzae. Akao and Okamoto (1983) have developed a process in which enzymes are produced from A. sojae grown on powdered wheat bran. They achieved a cell yield 2-3 times higher, and an enzyme activity 5-15 times higher than that achieved in an unfluidised solid-state culture, with cells making up 50% of the product dry weight. The process has been developed to pilot scale using a 3300 litre fermenter, and the cost of producing the enzyme has been favourably compared with static solid-state culture and with submerged culture.

1.3.1 Properties of Fluidised Bed Fermenters
The fluidisation characteristics of the support particles have been considered by Mishra et al (1982) and by Zimmerman et al (1984), who measured some of the basic fluidisation parameters in their systems. Mishra measured the minimum fluidisation velocity
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for pelletised yeast and compared this to a value calculated from equations 1.1 and 1.2.

Using this approach good agreement was found between the predicted and the measured values. The optimum operating conditions were chosen on the basis of these non-growth fluidisation experiments, but it was still found necessary to use a stirrer located at the bottom of the bed to break up agglomerates and any heavy, wet particles falling to the bottom of the bed.

Zimmerman et al (1984) studied the effect of varying the moisture content of the particles on the fluidisation behaviour. They observed that for cohesive particles, the minimum fluidisation velocity increases as the moisture content of the particles is increased, and associated this effect with changes in the particle dimensions, particle density, and the voidage of the bed. They were able to correlate the changes in the particle properties with the moisture content and used the resulting values for particle size and density and bed voidage in the Ergun equation to predict the minimum fluidisation velocity. When compared with measured values, this method predicts the minimum fluidisation velocity quite well for moisture contents up to 50% of the particle weight. They also studied the effect of micro-porous silica, a flow conditioner, on the fluidisation behaviour of the pellets. With the flow conditioner added, the minimum fluidisation velocity was constant up to 50% moisture, so that the experimental results did not agree with the correlation. To investigate the difference in the behaviour with the addition of the flow conditioner, the flow behaviour of the particles was studied using a rotational shear cell or Peschl shear tester. The measured parameter was flowability, $f_{f_c}$, where flowability is defined as the quotient of the major consolidation stress and the unconfined yield strength, and the cohesiveness of a material can be assessed on a scale:
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\[ ff_c < 2 \] very cohesive and non-flowing
\[ 2 \leq ff_c < 4 \] cohesive
\[ 4 \leq ff_c < 10 \] easy-flowing
\[ 10 \leq ff_c \] free-flowing

This classification is related to the fluidisation behaviour in that the unconfined yield strength determines the stability of any channels in the bed, which give rise to the poor fluidisation properties of cohesive materials. In this instance it was shown that the increase in \( U_{mf} \) observed as the moisture content was increased corresponded to a decrease in the value of \( ff_c \), from \( ff_c > 4 \) to \( ff_c < 4 \) for both untreated and conditioned pellets. Therefore this parameter appears to give a good indication of how the moisture content affects the ease with which a particular material can be fluidised.

However despite the apparent success of this method for particle characterisation, high biological activity of the yeast is only achieved with a particle moisture content of 60 - 70% w/w, (Zimmerman et al 1984), a level at which even the addition of conditioners cannot keep \( ff_c \) consistently above 4. Therefore any variation in the moisture levels will result in a change in \( U_{mf} \), which will affect the air flow required to keep the bed fluidised. Hence, if the bed is to be operated at the optimum conditions for the biological process, it will need, in addition to flow conditioners, a very tight control on the addition of liquid and the supply of air. A number of sensors for monitoring the conditions inside a fluidised bed have been developed. Zimmerman et al (1984) suggested using an opto-electronic measuring device, based on a helium-neon laser, to monitor the fluidisation conditions in the bed. Different signals are associated with bubbling fluidisation, homogeneous fluidisation and a fixed bed, and these could be used to change the set-point on the air-flow.
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controller, and/or the flow of liquid into the bed. Akao and Okamoto (1983) used a different approach, using an electrode in the bed to measure the electrostatic capacity between the electrode and the wall. The capacity can be correlated with the moisture content of the particles, and therefore the system can be used to control the addition of water to the bed, to maintain the optimum moisture content in the particles. The two systems differ in that the first is designed to keep the bed fluidised and the second is designed to maintain the optimum operating conditions; the result of the latter approach is that there is still a need to put a stirrer in the base of the bed, to break up any regions of defluidised particles and to mix any wet particles back into the bed. These results also show why the decision by Mishra et al (1982) to choose the operating conditions based on the fluidisation behaviour resulted in poor productivity from the culture, as the moisture content was almost certainly below the optimum.

This leads us to consider the control of the other process variables in the bed, such as moisture content, temperature, nutrients and oxygen, as well as the removal of waste products. As has been shown above, the fluidisation characteristics are closely linked to the moisture content of the bed, and the optimum moisture level is a compromise between the conflicting demands of good fluidisation characteristics and sufficient water activity for efficient fermentation. Control of the moisture levels is essential as the fluidised bed is a very efficient dryer (Reay 1985), though the drying rate can be reduced by humidifying the fluidising air before it enters the fermenter. However saturating the inlet air with water is a poor control method as the saturation point of air is closely linked to temperature. Thus any temperature fluctuations in the feed line or the bed itself will lead to condensation or drying, adding or removing water from the bed. Therefore it is easier to design for a controlled rate of drying and add additional water to the bed as an atomised spray.
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onto the surface, into the airstream or into the bulk of the bed. Hong et al (1988) attempted to run a fluidised bed system with no addition of water or nutrients, i.e. as a batch process rather than a fed batch. To achieve this they included amylases in the substrate charged to the reactor, to release glucose from starch gradually over the course of the fermentation. However they do not discuss the fluidisation characteristics or particle size of this bed material, which was a mixture of mashed potato, amylases, synthetic growth medium and baker's yeast, with about 80% moisture content on a wet basis. The bed was operated with a saturated air supply. It is clear that this system is not a gas-fluidised bed but an example of the type of three phase system classified as a slurry reactor or a three-phase bubble fluidised bed (Epstein 1981), in which the mixture of solids and liquid is continuous and the gas forms a dispersed phase.

The regular addition of a liquid phase into the reactor also allows the addition of soluble nutrients which may be desirable for optimum fermenter operation. For example, if yeast is grown aerobically in the presence of a high concentration of glucose, the glucose is only partially oxidised, producing ethanol; this is known as the Crabtree effect (Bailey and Ollis 1986). In a fluidised bed the volatile ethanol would be lost, and the yield of cell mass with respect to glucose would be low. However if the glucose is added slowly during the course of the fermentation, the concentration is never high enough to initiate the Crabtree effect, and it is oxidised to carbon dioxide and water, giving a higher yield.

1.3.2 Temperature Control

Moebus et al. (1978) have shown that the evaporation of water can be used to regulate the temperature of the fermentation at the laboratory scale. For effective operation the fermenter needs to be kept within one degree of the operating temperature, which for
most fermentations lies between 20 and 37°C; above 40°C the organisms begin to die. Heat is generated by the metabolism of the fermenting organisms, and must be continually removed. Calculations presented by Moebus et al (1978) show that for an aerobic yeast fermentation, it is possible to supply glucose as a solution such that the heat generated by the metabolism of the glucose is balanced by the latent heat removed by the water when it evaporates. This is the case for laboratory-scale fermenters, but if the system is scaled up, the removal of heat by evaporation is limited by the saturation concentration of water vapour in the gas stream. The saturation concentration of water vapour is a function of temperature, and is therefore fixed for most fermentations, while the flow rate of gas is linked to the fluidisation characteristics required. Therefore the capacity of the air stream for water vapour is defined by the process conditions. As the production of heat occurs throughout the bed and is proportional to the volume of the biomass and hence to the bed volume, and the volumetric flow rate of the gas in a fluidised bed is proportional to the base area, the capacity for water removal does not increase as rapidly as the heat production rate when a bed is scaled up with a fixed aspect ratio. Only in the case where scale-up is achieved entirely by increasing the base area (without changing bed depth) does the capacity for cooling increase in proportion to the heat production rate. Therefore there is a theoretical limit to the height of bed which can be operated without an additional heat removal system. Additional removal of heat as sensible heat in the gas stream is limited by the low heat capacity of air, even when the use of chilled air is considered. A simple calculation will show that for a unit mass of 1 kg of dry air at 30°C the heat removed by the evaporation of enough water to saturate the air would be roughly equivalent to the heat required to heat the air to 91°C:
latent heat of evaporation at 30°C = 2430.7 kJ / kg
saturation of air at 30°C = 0.025 kg water / kg air
heat removed by evaporation = 61 kJ
heat capacity of air at 30°C = 1.0057 kJ / kg.K

Temperature rise produced if 61 kJ of heat are absorbed, or equivalent subcooling of fluidising air = 61 K.

Therefore we can calculate the maximum height of bed which can be operated with evaporative cooling as the main heat removal mechanism, if the specific heat production rate and the fluidisation characteristics of the support particles are known. For vessels of high aspect ratio, which is typical of industrial fermenters as it reduces the land costs and the volume of air which needs to be compressed, additional cooling in the form of jackets or cooling coils must be added to remove excess heat. This type of design requires an understanding of the heat transfer rates between the bed and the heat transfer surfaces.

1.3.3 Control of Other Process Parameters

The supply of oxygen to the fermenter is provided by the fluidising air, which can be enriched with oxygen if necessary. Similarly gaseous waste products, notably carbon dioxide and water, are carried out of the reactor in the flow of gas. Volatile products such as ethanol can be recovered from the exhaust gases by condensation (Moebus et al 1978), though this will give a dilute aqueous solution. Non-volatile products and waste products will stay on the particles and accumulate in the reactor; these can be removed either continuously by removing some of the particles and replacing them with fresh ones, or at the end of a batch process. In comparison with submerged culture the products are in a much more concentrated form, as there is much less water in the system, and this should make the recovery and purification of the final product much cheaper, by reducing the dewatering steps necessary.
1.4 Conclusions

The literature indicates that there may be advantages to operating fermentation processes in gas-phase-continuous three-phase fluidised beds. However, experiments have shown that there is insufficient data to design an effective system, particularly at scales suitable for industrial production. Development of a successful fluidised bed fermenter design will depend on both an understanding of the required process conditions and a process control system which can maintain these consistently. In particular, scale-up of the process will require provision for the removal of metabolic heat above that quantity which can be carried away by the air-stream. This will probably depend on a cooling duty supplied by internal cooling plates or coils, or through the wall of the bed. As has been stated above, heat transfer in three-phase fluidised beds is a field about which little is known.

1.5 This Work

Chapter 2 describes a study of a range of bed designs, mainly taken from the literature, which were tested for their suitability for three-phase gas-continuous fluidisation. From this work the essential design features of a fermenter as described above were determined and a number of prototype designs were built and tested.

In Chapter 3 the hydrodynamics of this design are discussed. The mobilisation point, which has some similarity to the minimum fluidisation point but also important differences, has been modelled analytically, and the results tested against experiment. Chapter 4 covers a study of the heat transfer between the walls of the bed and the gas stream. Drying rates in the bed are also discussed. In Chapter 5 the conclusions from the work are summarised.
Figure 1.1 Fluidisation behaviour as a function of particle size and density (from Geldart 1986)
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2.1 Introduction
It has been established in Chapter 1 that a fluidised bed can be used for solid substrate fermentations, but that the control of the process is difficult, due to the number of parameters involved. In this chapter the required operating conditions will be discussed more thoroughly. The design of a fluidised bed fermenter poses a number of problems for the process engineer because it is a three phase process; all three phases have to be controlled to maintain constant operating conditions in the reactor. In the process of designing a suitable system, some of the various bed designs which are available have been tested, using wet porous particles to model the fermenter contents.

2.1.1 Water Content
In the gas-fluidised-bed fermenter the bed of solid particles is fluidised by the flow of air and the liquid phase is supported on and within the particles; the micro-organisms may grow both within the pores and on the surface of the particles. The amount of free water present is particularly critical; microorganisms cannot grow without moisture as they need to take up nutrients as solutes from an aqueous environment. The amount of liquid which needs to be present is related to the resistance of the organism to dessication. Some organisms such as bacteria only grow when they are in a film of moisture; they are killed by dehydration and only survive dry conditions by producing spores. Yeast also needs a film of water for growth, but it can be dehydrated and remain viable but dormant, whereas filamentous fungi do not need to be submerged in water and will continue to grow as long as sufficient soluble nutrients are available to part of the fungal mycelium. The productivity of any organism is maximised by keeping as many of the microbes as possible viable and active. In a fluidised bed, if water is lost by evaporation into the airstream, this must be replaced to maintain the optimum conditions. Therefore the particles are usually subject to continuous or intermittent addition of water or nutrient solution. The requirements of the micro-organism will define the
minimum desirable liquid loading for each fermenting system.

However, as has been discussed above in Chapter 1, fluidisation behaviour is very sensitive to the amount of water present. The addition of water to a gas-fluidised bed can give rise to large inter-particle forces, due to capillary action and viscous shear. In a fermentation process these are increased by the presence of polysaccharides and other biological solutes in the liquid film, which increase the viscosity of the liquid. The increase in interparticle forces changes the fluidisation behaviour from type B through type A towards type C (using Geldart's classification), and this leads to the loss of some of the characteristics of good fluidisation such as good mixing and high heat and mass transfer coefficients.

The combination of capillary and viscous forces holding particles together and the drying conditions in the bed also produces a marked tendency for the particles to agglomerate or attach to the walls of the bed. As the minimum fluidisation velocity of the system is a function of particle size, at a constant air flow rate any increase in the particle size due to agglomeration will eventually lead to defluidisation. In a gas-fluidised bed this is known as quenching. Agglomeration may proceed gradually, leading to failure after some time, or it may be rapid, leading to catastrophic failure of the bed. Although a wet quenching "point" can only be defined for the simplest model systems, the probability of a failure due to quenching is linked to the liquid loading of the bed. Therefore for a particular system, the possibility of wet quenching sets an upper limit on the liquid loading that the fluidised bed can support.

From the arguments above it can be seen that there is a "window" of conditions, defined in terms of the liquid loading, within which fluidised bed fermentation is possible. At ICI*, work on a process

*Personal Communication, J. C. Middleton, 1986
2.1.2 Particles
In addition to the water content, in a fermenting system the nature of the particles is also changing due to biological processes. Substrates are consumed and biomass, carbon dioxide and water are produced, so the absolute amount of biomass may change over the course of a fermentation, increasing at first and then decreasing as the fermentation matures. The mechanical properties of particles made by granulating the substrate (or the biomass in the case of yeast fermentations) may be substantially altered by the fermentation process. If, for example, they become softer or more brittle they will tend to erode, and the presence of fines in the bed may aggravate the problems of cohesion and agglomeration described above. Any changes in the mean particle size will make the bed more difficult to control, and the presence of a surface layer of soluble fines and water may lead to higher adhesive forces between the particles. In the case of non-porous supports the evolution of metabolic gas within the particles may be sufficient to disrupt them.

2.1.3 Previously used Fluidised Beds
In trying to improve the quality of fluidisation various approaches have been tried but the majority of workers have placed some form of mechanical stirrer in the bottom of the bed to allow them to operate with the preferred levels of free water. The stirrer breaks up agglomerates and scrapes off material which has stuck to the wall or base. At ICI (ibid) problems with agglomeration were only overcome in the laboratory by abandoning the fluidised bed and using a helical ribbon stirrer in a packed bed of particles, but this approach is not seen as a solution for an industrial scale process.

2.1.4 Stirrers
A mechanical stirrer is unsatisfactory for two reasons. Firstly, it
introduces a need for seals around rotating shafts in the wall of a sterile system. There is therefore an increased risk of contamination, the capital cost is higher and extra sterilisation steps are necessary in the operation of the reactor. Secondly, it is one of the advantages of the fluidised bed that it is very well mixed, and therefore the main purpose of a stirrer in submerged culture, to provide mixing and particularly to increase the rate of oxygen transfer into solution, can be fulfilled by the flow of fluidising air alone. Therefore it would seem possible and preferable to design a system where the flow of air is used to regulate conditions in the bottom of the bed, rather than a mechanical stirrer.

In submerged culture the advantage of eliminating the drive shaft has been seen in airlift fermenters where the agitation is provided by the flow of air; the shaft and seals are dispensed with and this removes a major source of contamination. Airlift fermenters are only limited in application because of the low mass transfer coefficients obtained.

2.2 Initial Phase of Current Work:
In trying to develop an improved design for a fluidised bed fermenter it was decided to consider both the support particles and the bed design itself.

2.2.1 Choice of Particles
The first aspect of the system which was considered was the material used to make the support particles. It was decided that the predictability and control of the bed would be improved by using inert supports for the substrate, as this reduces the effect of both substrate consumption and fluctuations in the biomass hold-up on the particle properties.

Biological compatibility is important for fermenter applications, as the culture will be more stable if the micro-organisms bind strongly
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to the support. Therefore materials should be chosen which provide the right combination of surface texture and surface chemistry to promote adhesion.

By using porous materials, a large surface area can be provided for microbial growth which increases the intensity of the process (expressed as yield per unit volume of reactor). High porosity also increases the heat and mass transfer rates between the surface and biomass within the particle, and permits inoculation of the seed culture throughout the particle.

The particle size tends to remain more uniform if the support provides a matrix stronger than the supported biomass and substrate: biomass growing outside the matrix is more likely to be eroded away by interparticle collisions, while any agglomerates which form will tend to break along the inter-particle planes if these form lines of weakness. Stronger particles are also more robust in the recovery and extraction stages; in the extreme case particles can be recycled to the reactor after removal of the product, if they are sufficiently robust.

In choosing a support material it is also important to consider the density of the material, as in a gas-fluidised-bed the pressure drop across the bed equals the bed weight, and hence the compression costs are a function of the bulk density of the bed.

The important characteristics for the particles used in a gas fluidised bed application were therefore considered to be: biological compatibility, low bulk density, high intrinsic strength, high porosity and high specific surface area.

To ensure biological compatibility various media which have already been used as biological support particles were considered, such as: alginate gel beads; microcarriers made of alginate gel; celite; steel and plastic meshes; porous foams and wheat bran. Some of
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these have been used only for suspension culture or liquid fluidised beds and are not suitable for non-submerged conditions, such as alginate gel which becomes brittle when dehydrated. Steel meshes are relatively heavy, while wheatbran is not biologically inert and is also relatively weak and subject to attrition. Therefore for our application porous polyester foam was chosen, as it fulfills the selection criteria. For this work foam was supplied by Declon Plastics as both cubes (5mm each side) and cylinders (5mm in length and 5mm in diameter). These particles were then used to assess a selection of modified fluidised-bed designs, to try to identify a system which could be used for cohesive particles without the use of a mechanical stirrer.

Examples of the various particle types used in this study are shown in Fig. 2.1. These include the two types of foam supplied by Declon; the same foam coated with substrate before and after autoclaving; and "Puffed Wheat" breakfast cereal. This selection of particle types was used for the preliminary experiments and also for the quantitative experiments on hydrodynamics and heat transfer which are discussed in chapters 3 and 4.

2.2.2 Initial Tests of Bed Designs

There are a number of designs for modified fluidised and spouted beds in the literature which are specifically intended for use with large or difficult particles. Many of these have air inlet layouts designed to impose a regular circulation pattern on the bed contents. Prototypes of a number of these designs were available in the University, and others were built to test their suitability for use with the support particles under both dry conditions and with water present.

The polyester cylinders were used for the initial experiments, in

4( 83/84 Manton Rd, Earlstrees Industrial Estate, Corby, Northants, NN17 2JL )
three forms: the dry cylinders on their own, coated with a dry mixture of nutrients, and with water added. The results of the tests of these designs are presented below. For each of the six types tested the bed design is shown and the behaviour of the bed is described qualitatively in the text. The beds were assessed on the following points: whether the cubes could be fluidised or "mobilised"; how stable the "mobilisation" was over time; whether the bed was well mixed or only partially mixed with areas of stagnation; and finally the beds were compared quantitatively on the basis of the amount of air needed both to initiate some motion of the bed contents and to establish a "mobilised" state. Results for each bed design and comparisons between the designs are plotted as bar charts, to compare the amounts of air needed. These predictions are given as a basis for comparison between the bed designs, being based on qualitative observations only.

2.2.3 Calculated Values of Minimum Fluidisation Velocity and Minimum Spouting Velocity for Test Particles

For the purposes of comparison the minimum fluidisation velocity of the test particles can be calculated from the correlations published in the literature. One of the most commonly used correlations is that of Wen and Yu (Geldart 1986), although this correlation is known not to be reliable for particles well above 1mm in diameter. The equation is derived from equations 1.1 and 1.2 in Chapter 1, which have been simplified by using a correlation for some of the variables:

$$U_{\text{mf}} = \frac{\mu}{\rho_f \, d_v} \left( 1135.7 + 0.0408 \, Ar \right)^{1/2} - 33.7$$  \hspace{1cm} (2.1)

$$Ar = \frac{\rho_f \, d_v^3 \left( \rho_p - \rho_f \right) \, g}{\mu_f^2}$$

Similarly, a correlation is available for the minimum spouting velocity in a spouted bed. One of the most widely applied equations is the Mathur-Gishler equation (see Mathur and Epstein 1974).
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This has been shown by experiment to give results within 10% of experimental results.

\[ U_{ms} = \left( \frac{d_p}{D_c} \right) \left( \frac{D_1}{D_c} \right)^{1/3} \left( 2 \ g \ H \ (\rho_p - \rho_f) \right)^{1/2} \]  \tag{2.2}

In this equation \( d_p \) can be taken as the smaller dimension for ellipsoidal particles such as wheat, or as \( d_e \) the diameter of the equivalent-volume sphere for irregular particles.

Theoretical results for two of the experimental particles are shown in table 2.1.

<table>
<thead>
<tr>
<th>( \mu_f )</th>
<th>1.8 ( \times ) 10(^{-5} ) N s m(^{-2} )</th>
<th>( D_c )</th>
<th>145 mm</th>
</tr>
</thead>
<tbody>
<tr>
<td>( \rho_f )</td>
<td>1.29 kg m(^{-3} )</td>
<td>( H )</td>
<td>200 mm</td>
</tr>
<tr>
<td>( D_1 )</td>
<td>25 mm</td>
<td>( g )</td>
<td>9.81</td>
</tr>
</tbody>
</table>

Table 2.1 Theoretical Results for The Experimental Particles

| \( d_v \) | 5.72 mm | 7.9 mm |
| \( \rho_p \) | 100 kg m\(^{-3} \) | 60 kg m\(^{-3} \) |
| \( \alpha_r \) | 7.23 \( \times \) 10\(^5 \) | 4.95 \( \times \) 10\(^5 \) |
| \( U_{mf} \) | 0.43 m s\(^{-1} \) | 0.46 m s\(^{-1} \) |
| \( U_{ms} \) | 0.38 m s\(^{-1} \) | 0.41 m s\(^{-1} \) |

At this scale the results are similar for minimum fluidisation and minimum spouting velocities. However, as is apparent from the equations, the minimum fluidisation velocity is independent of the bed dimensions, while the minimum spouting velocity is a function of the bed diameter \( D_c \), and bed depth \( H \).

2.2.4 Results From Preliminary Experiments

Standard Fluidised Bed

The dry particles were not expected to perform well in the standard fluidised bed but it was included in the study for comparison. The bed used (Fig. 2.2) was a perspex cylinder with an aluminium perforated plate distributor, 145 mm in diameter. The distributor had 3mm holes on a 5mm triangular pitch. The bed was loaded with
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two litres of the dry foam cylinders, a weight of approximately 20 g with a porosity of 99%.

According to Geldart's classification by density and size, (Geldart, 1986) the particles fall into group B, but as the interparticle forces are not negligible, Molerus' criterion moves them towards group A.

In the case of the polyester foam, the porosity of the particles is as high as 99%, and as they are reticulate, i.e. the intra-particle voids are interconnected, the air passes through the bed both between and through the particles. As the air flow is increased the pressure drop across the bed increases according to the Ergun equation, as discussed in chapter 1, but the particles are not forced apart by the flow of air, due to the effect of mechanical interlocking. Therefore the static friction between the particles is not reduced by expansion of the bed and separation of the particles.

The foam has a rough surface, which gives the particle-particle contacts a high coefficient of friction, while the the coefficient of friction for the particle-wall contacts is much lower as the wall is smooth. Therefore, as the air flow rate is increased and the pressure drop across the bed becomes equal to or greater than the weight of the bed, the initial failure is at the wall and the whole bed is pushed upwards as a plug. If the plug is broken mechanically, or by applying a pulsating gas flow, and fluidised behaviour is established, the bed then tends to collapse around stable channels through which the largest proportion of the gas passes. At higher gas flowrates the particles are carried out of the bed with the gas, as their terminal velocity, a function of weight, is only slightly higher than their fluidisation velocity. With heavier particles, in this case foam cylinders coated with dry solid nutrients, some of the same effects are observed; for example, the inter-particle forces are sufficiently high to interfere with
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the initiation of fluidisation. At higher flowrates the bed can be fluidised, as local inequalities in the flowrate through the plug exert sufficient force to break the interparticle bonds and establish a "slugging" fluidised bed. But there is still a tendency for the bed to collapse around a few large channels.

With the addition of water, which introduces inter-particle forces due to liquid bridges between the particles, fluidised behaviour cannot be established at all, and the particles form large stagnant clumps against the walls and on the base of the bed. The wetted bed only begins to fluidise when the bulk of the liquid has evaporated into the air stream; i.e. it has dried out sufficiently for all the remaining water to be accommodated within the pores in the particles.

For these conditions, under which the particles do not demonstrate a clearly defined minimum fluidisation point consistent with the theory, \( \Delta P = \frac{W}{A_B} \) there is nonetheless a point at which free movement and mixing of the particles occurs. For the purpose of this work this point is described as the mobilisation point.

These results are summarised in figure 2.3 and show that the air flow to establish mobilisation is almost the same as the airflow required to maintain it. This is indicative of the high flow required to initiate any motion at all. The initial resistance, due to static friction and adhesion, is higher than the dynamic friction once the bed is mobilised, therefore maintenance of mobilisation in the bed requires no additional air. These results show that the unmodified standard fluidised bed is not suitable for three-phase fluidisation.

A second distributor plate was also tested in this bed, which had 1 mm holes on a similar 5 mm triangular pitch. This produced a marked reduction in the superficial velocity required to initiate motion and a reduction in the air required to maintain mobilisation.
of the bed contents. For this bed the initial resistance is overcome at a flowrate lower than that required to maintain mobilisation. It was also possible to mobilise wet particles in this version. The results are shown in figure 2.4. An initial examination of the differences between the two beds leads to the observation that the momentum of the jets at the distributor will be higher in the second bed and this may overcome the inter-particle forces sufficiently to establish mobilisation.

Spouted Bed
The spouted bed design is shown in figure 2.5. The spouted bed was developed by Mathur and Gishler (1955) for operations on large particles, such as grain, which could not be processed effectively in fluidised beds. In the basic design the air is introduced through a single concentric orifice at the bottom of a conical section. The particles are entrained into the air flow at the base of the bed and carried up in a central spout. At the top of the bed the particles fall back onto the bed surface around the spout. The particles return to the bottom of the bed in the annular region, which has the properties of a moving packed bed.

For these tests a spouted bed with a cylindrical perspex top section 150 mm in diameter and a conical section with a 30° half angle was used. The inlet was 25 mm in diameter but it could be varied by inserting a flat plate in the flange. The bed was loaded with two litres of dry particles, as before.

With the 25 mm orifice the bed spouted well, forming a steadily circulating bed, though there was a small fluctuation in the downwards velocity at the wall. With the inlet reduced, which introduced a small concentric flat area in the base of the bed, the downwards flow in the annulus was arrested and the particles stacked up against the walls. The rate of particle entrainment was reduced until the greater proportion of the gas passed through the spout without promoting any circulation of the particles. With an
increased gas flowrate the particles were carried out of the bed.

Therefore the spouted bed is suitable for handling the large, light particles being tested. However, if there is a small area of the base on which the particles can come to rest, stagnant areas occur and rapidly grow, allowing more of the air to bypass the bed through the central void. If the particles form a stable mass, circulation of the particles is much reduced as they are not entrained by the gas at the base of the spout.

The results are shown in figure 2.6. Tests with the coated and wetted particles were similar to those with dry cylinders, showing an increase in the minimum mobilising velocity with increasing particle density, and a consistent increase of 15 to 30% of the initial mobilisation velocity required to maintain stable circulation in the bed.

Tumbling Bed
The tumbling bed is shown in figure 2.7. This design is one of a number of configurations which have been proposed for handling awkwardly shaped particles. Sul'g and co-workers (1974) developed the bed in the Soviet Union, for dehydrating solutions and drying dusty particles. The base is curved and the air is supplied tangentially through a slit at the bottom of the bed so the flow of air sweeps the base and entrains some of the particles. These are carried up in a fountain on the opposite side to the inlet and fall back onto the surface. As the air also percolates up through the bulk of the bed this produces an aerated bed with a steady pattern of circulation, in which any particles which sink to the bottom are recirculated by the stream of air across the base.

The bed used in this study was square in cross-section, 150 mm on each side, with a 150 mm radius curved section forming one wall and the base of the lower section. The air inlet was 5 mm deep and 150 mm long. The bed was loaded with one litre of foam cylinders as
Chapter 2: Preliminary Tests on Bed Designs

described above.

This design was ineffective for mobilising the particles, as it was not possible to establish a regular circulation. As the air flowrate was increased the bed did not expand and the particles did not separate, due to the interlocking between them. Therefore, when the air flow was sufficient to produce a pressure drop across the bed equal to the weight of the particles, the whole bed lifted as a plug. This rapidly broke up and the particles came to rest around a number of channels which allowed the air to flow vertically upwards from the inlet thus bypassing the bulk of the bed completely. Further increases in the air flowrate increased the size of the channels and did not mobilise the bed material. Attempts to mobilise wet or coated particles were also unsuccessful. Therefore, no results can be quoted for this bed design.

The tumbling bed does not appear to be suitable for our use. In the USSR the design was used for mineral drying, in which the particles are much larger, allowing larger flowrates to be used. The tendency of the air to bypass the bed and take the shortest route to the surface allows any stagnant area in the base of the bed to grow; this tendency might be reduced at higher gas flowrates, when the air would have more momentum to sweep across the base.

Whirling Bed
The whirling bed is shown in figure 2.8. This type of bed was developed by Rios and co-workers (1980) for use with large particles. In action it is similar to the tumbling bed, in that it induces rotation in the bed around a horizontal axis. In this case, however, the air enters vertically through a perforated plate distributor on one side of the base. Particles are entrained at the distributor where the air velocity is high, and carried up to the top of the bed, where they fall back either onto the surface or onto the sloping insert, which directs them back into the air stream at the base.
For these tests a cylindrical perspex bed with an aluminium perforated plate distributor (3 mm holes on a 5 mm triangular pitch) was used. An insert was made which covered half the distributor and formed a slope at 45° to the horizontal (as recommended by Rios et al.) for the particles to fall back onto. The behaviour of the whirling bed when the airflow was increased was different from that observed in the tumbling bed. With the bed filled with two litres of dry cylinders which weighed approximately 20g and gave a depth of approximately 150 mm, the bed did not expand as the airflow was initially increased. However, the contents could be made to rotate about a horizontal axis at higher gas flowrates, though there was a tendency for the bed to separate into two sections: the lower section rotated rapidly, while the upper section formed a stationary plug supported in the top of the bed by the gas flowrate.

With the coated particles, which are heavier, a regular circulation was established more easily, with no visible stagnant areas. In this case the weight of the particles was sufficient to prevent bridging across the bed to form a plug. The wetted particles were also mobilised successfully: however, when the bed was emptied, a stagnant area of wet particles was found on the distributor at the base of the slope, where they had come to rest rather than mixing into the bulk of the bed. The results are shown in figure 2.9. The pattern is similar to that observed in the spouted bed, with a significant difference between the air required for initiating motion and that required to produce a stable circulation in the bed.

The tendency for a stagnant area to form on a small horizontal surface is similar to the situation in the spouted bed observed above. This indicates that in order for this design to be suitable for use as a fermenter, the whole base must be aerated efficiently.
Annular Spouted Bed
This design is shown in figure 2.10 and was tested on the basis that the main problem with the spouted bed is that the slowly descending particles in the annulus can stack up against the walls of the bed. By directing the flow of air upwards at the walls and allowing the particles to descend in the centre of the bed there is less likelihood of the particles coming to rest and forming stagnant zones.

The test bed was 145 mm in diameter, with a cylindrical perspex top section. The distributor was a perforated aluminium plate with 3 mm holes on a 5 mm triangular pitch. The conical insert was placed concentrically on the distributor, and had a base diameter of 100 mm, so that the air entered through an annulus 22.5 mm in width. The cone subtended an angle of 60° at the apex.

In this bed both the dry foam and the coated foam particles could be mobilised, with vigorous motion at the base of the bed (where the air velocity is highest) and a bubbling fluidised bed above this region. However with the coated particles the gap at the base of the bed was sufficiently narrow for the particles to form arches and become jammed, which allowed the formation of stagnant areas. The wet particles were less difficult to mobilise, and did not form stagnant zones; however they tended to clump together and mixed unevenly until the greater part of the water had been removed by evaporation. The results are summarised in figure 2.11.

This design fulfills the requirement for vigorous agitation and circulation of the particles in the base of the bed. However the narrowness of the gap above the annulus allows bridging, as was noted for dry particles in the tumbling bed. The probability of bridging is a function of both the particle size and density and the dimensions of the bed; therefore in larger units or with smaller particles this would not necessarily occur. However, the use of the extra unaerated sloping surfaces in the base of the bed may also
cause problems, because the particles can develop stagnant areas from the point where the slope reaches the base, as seen in the tumbling bed.

**Legros Bed**

A novel bed design which was developed at the University of Surrey was also tested for this application (Legros 1986). The design is shown in figure 2.12. The Legros bed was developed to process fibrous materials such as tobacco, which are fragile and because they interlock readily are difficult to handle in a conventional fluidised bed. The design takes the form of a modified square section spouted bed. The distributor is an inverted right pyramid, with an air inlet at the apex and a number of inlets in the side walls. The air supply to the apex, known as the accelerating flow, and the supply to the side walls, known as the mobilising flow, can be controlled separately. The mobilising flow enters the bed as a number of free jets, perpendicular to the bed wall; these jets entrain air from the volume around them, and because the base of the bed is enclosed, most of this entrained air is drawn down from the surface of the bed producing a downflow of air at the wall. The accelerating flow lifts the fibres and aerates them, and produces an upflow of air in the centre of the bed, entraining air from around the apex. The overall result is a smooth circulation in the bed, between an upward flow in the centre and a downward flow at the wall, which circulates the fibres without tangling or breaking them.

The bed used for these tests was a 150 mm square perspex top section on a perspex base with four walls set at 45° to the horizontal. Each side of the base had nine 2 mm air inlets in a three by three array, and the inlet at the bottom of the bed was 12 mm in diameter.

In the work reported here this design performed well with the dry particles which were easily aerated and circulated, while no stagnant areas formed in the bed. However there was a significant difference in the flow pattern in the bed compared to that observed
Chapter 2: Preliminary Tests on Bed Designs

by Legros et al with fibrous material. Legros found that a mobilised bed of fibres will collapse if the downflow at the wall is obstructed. This can be tested by using a mask, for example a square of rigid material the same size as the bed cross section with a concentric square hole in it. When this is lowered into the bed, parallel to the base, the circulation fails as it approaches the bed surface. This shows that the flow of air entrained from the surface into the wall jets is a significant proportion of the air required to mobilise the fibres. However, in the case of the light foam particles, the bed does not collapse when the mask is applied. So it appears that sufficient air to lift the particles is supplied through the distributor and the circulation of the bed is due to the upward force from the accelerating flow being relatively higher than the upward force at the wall due to the mobilising flow. Particles which fall outward from the central spout return to the base of the bed by sliding down the walls under their own weight. The coated particles were not available when this bed was tested, and therefore there are no comparative results.

These results were encouraging, as this design aerates and mixes the bed without the introduction of narrow gaps or extra surfaces in the base of the bed. Hence it overcomes some of the problems observed in the alternative designs.

2.2.5 Summary of Results from the Initial Experiments

The performance of five of the bed designs is compared for each particle type in figures 2.13 to 2.15. The whirling bed is omitted because no useful results were obtained, and data for wet particles were not obtained with the Legros bed. In general the beds are comparable on the basis of the air required for mobilisation. Two beds showed a significant difference in behaviour: the fluidised bed with the more open distributor required more air to move or mobilise the heavier, coated particles, and failed to mobilise the wetted particles; the spouted bed required more air to mobilise the dry and coated particles.
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On the basis of the selection criteria, the first fluidised bed is unsuitable because the wetted particles could not be mobilised. The conical and tumbling beds were rejected due to the uneven mobilisation, and the tendency to form stagnant areas either by bridging, or between the distributor and the inclined plane. From the remaining beds, which include the Legros design, the spouted bed and the fluidised bed with a low free area distributor, a number of common features can be identified. In each case jets are acting on the bed, either from the base or the walls, which may help to break adhesion between the particles and prevent bridging and stagnation, and the bed is open, with no internals to provide areas for stagnation. In the case of the spouted bed and the Legros bed the particles are well mixed from top to bottom, which is of particular importance for fermentation applications; therefore these were the preferred designs after the initial tests.

2.2.6 Tests of Selected Beds with Wet Materials
From the initial results it was decided to test the spouted bed and the Legros bed, under more typical three phase conditions. The first tests, in which water was added to the uncoated particles, were unsuitable as a model as the water ran down through the particles and collected in the base of the bed. This is a poor representation of the conditions in the fermenter where the water is held by capillary forces within the biomass and substrate supported on the particles. To model this the particles were filled with alginate gel by precipitating it in the pores, to represent the mixture of biomass, substrate and water in a fermenting system. Wet substrate coated particles cannot be used in the test rigs for two reasons: firstly the rigs provide an ideal environment for microbial growth so that any test material is rapidly contaminated by bacteria or yeast; secondly, the initial rigs had no humidity control on the air supply, so any unsaturated particles would be dried out very rapidly.

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2.2.7 Experimental Method for Alginate Cubes

Calcium alginate gel is formed by precipitation when a solution of sodium alginate in water is mixed with a solution of calcium salts. The method used here was adapted from a method used by the biotechnology group in the Chemical Engineering department at Surrey for the preparation of gel microbeads for submerged culture. To precipitate the gel in the pores of the foam, the 5 mm cubes were soaked in a 2% w/v solution of sodium alginate, and then rapidly transferred to a 0.05% solution of calcium chloride. On contact, a gel skin forms at the particle surface, trapping the alginate solution inside the particle. The particles are then left in the calcium solution overnight, to allow calcium ions to exchange with the sodium ions by diffusion, so that the gel is precipitated throughout the whole particle volume.

The particles were kept moist between runs in a solution of calcium chloride, to maintain the stability of the gel. They were transferred directly from the solution to the bed, including any free liquid on their surfaces.

These particles are relatively heavy: the particle density is about 700 kg/m³ compared to about 200 kg/m³ for the dry coated particles. The substrate-coated particles are expected to be more rigid and less elastic than the test material, while their water content is not expected to be higher than about 50-90% of the dry weight of the particles, considerably less than that of the gel-filled particles which are saturated with water. Therefore alginate-filled particles represent a "worst case" and typical substrate coated particles in a fermenter should be easier to mobilise. The particles which are used should provide data on the limiting case for heavy, wet particles. It should also be noted here that the capillary forces acting in the model are due to the presence of a weak salt solution only; in the fermenter these forces may be amplified by the presence of polysaccharides. However it is considered that the model system was adequate to test the bed designs.
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In the spouted bed the alginate-coated particles could be circulated, but there were stagnant areas where the particles tended to stick to the walls. Around the spout there was sufficient agitation to keep most of the particles in motion, but at the base of the bed, in the conical section, some particles could come to rest at the walls. Capillary forces then held the particles in place and they were not carried down into the spout. The air flow did not dislodge them as the air will flow preferentially around a constriction and the gap between the particle and the wall or between two particles is reduced when there is a film of water between them. Therefore any small area of stagnation grew rapidly as the particles stacked up, sticking both to the wall and to each other, until it extended into the region where the spout provided sufficient agitation to shear off the particles and recirculate them.

The Legros bed was more effective than the spouted bed, but the same phenomenon of stagnant areas in the angled base of the bed was observed, though they were less stable and were regularly mixed back into the bed. The moving areas of stagnation led to changes in the pressure drop across the small air inlets in the wall, altering the air distribution to the jets and leading to unsteady, pulsating flow through these inlets, which in turn led to both uneven mixing and circulation of the particles, and channelling, whereby a proportion of the air bypassed the bed.

2.2.8 Overall Summary of Problems and Possible Solutions
A spouted bed with auxiliary gas flows to enhance the entrainment of particles into the spout (the Legros design) has some advantages over a conventional spouted bed when used for a three-phase system. However, the particular effectiveness of the design seen when processing fibres depends on the large-scale entrainment of gas into the jets, which promotes a regular recirculation of the particles, and this was not observed in our tests. In addition the shallowness
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of the bed and the proximity of the wall inlets to the surface reduce the gas residence time and reduce the efficiency of the gas-solid contacting in the bed. Scale-up in this rather complex design may also prove to be difficult; tests by Legros (1986) showed that the behaviour of larger beds could not be easily predicted and that the usable bed depth is limited to the depth of the conical section.

In the standard spouted bed, regions of stagnation occur where the shear forces (exerted by the flow of gas) are less than the force required to overcome the friction or adhesion, either between particles or between the particles and the wall. In a fermentation application, the microorganisms will be active even in static areas and these will become more strongly glued together as biomass and large biological molecules accumulate between the particles, so the formation of any static areas must be prevented. The problems of agglomeration and adhesion are most acute in the bottom of the bed where particles stack up against the walls and base. The drying effect of the air is strongest near the inlet in the base where the air is not saturated, and therefore agglomerates tend to form here; any large agglomerates will be insufficiently fluidised and will rest on the base forming nuclei for stagnant areas. The downflow of particles in the annulus as a moving packed bed encourages the particles to stick to one another, and particles are then compacted when they reach the base of the bed, aggravating this tendency. The stirrers used in the examples of fluidised fermenters mentioned above break up the agglomerates which form and mix the particles at the base back into the bulk of the bed.

The best features of the beds reviewed above include regular circulation which ensures even distribution of additives, and high shear forces in the region of the base and walls to disrupt agglomerates and prevent the formation of stagnant areas. To combine these features in one bed, a new bed design was proposed.
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2.3 Design of the New Bed

To achieve sufficiently agitated conditions in the bottom of the bed to prevent these problems, but without resorting to the use of a mechanical stirrer, a bed was designed which combined some of the features of both the spouted bed and the Legros design. The prototype had a spouted bed distributor and a single set of auxiliary air jets at one level in the walls. The jets from the wall inlets entered the bed horizontally, so they would move any groups of particles which came to rest. It was anticipated that the shear forces exerted by the jets would also break up some of the agglomerates.

2.3.1 Construction Details

The proposed design was for a cylindro-conical bed 145 mm in diameter, made from mild steel with a perspex cylindrical section, with a central spout and four 10 mm diameter inlets through the vertical side walls, 15 mm above the conical section. Air was supplied to the wall inlets and to the spout through separate manifolds, and the flow to each could be controlled independently. In addition, one or more of the wall inlets could be shut off. The design is shown in figure 2.16.

2.4 Qualitative Experiments

Initial tests with foam cubes showed that the action of the jets prevented a build-up of static particles above the jets, but not in the conical section. To allow the jets to act closer to the base of the bed, further tests were done using a flat base-plate with a central spout 25 mm in diameter. Two different inlet orientations were tried. In the first of these (Fig 2.17a), the inlets were perpendicular to the wall, to push the particles towards the spout where they could be entrained into the upward flow of gas. However, when the bed was run there were still areas of stagnation between the jets, and channelling rapidly developed. The design was then modified and a second column was built, in which the inlets were set at an angle in the walls (at 45° to the tangent to the wall, though...
still parallel to the base). In this design the action of the jets was expected to accelerate the particles into a circular flow pattern in the lower part of the bed, sweeping away any stagnant areas between the jets; this orientation is shown in figure 2.17b.

2.4.1 Effectiveness with Test Particles

The tests on this second version of the prototype were very satisfactory. The dry particles were easily fluidised and experiments with wet and alginate-filled particles were also successful. The action of the jets easily established a circular flow of particles in the base of the bed, and particles were also entrained into the spout and carried up to the top of the bed with an equivalent flow of particles downward at the wall. Therefore the particles were well mixed and circulated frequently between the top and bottom of the bed. As the particles have a velocity component parallel to the base they are prevented from settling out at the bottom, and as the jets are small, high velocities can be used compared to the superficial velocity of the gas in the bulk of the bed; a high superficial velocity would carry material out of the top of the bed. No areas of stagnation were apparent anywhere in the bed.

2.4.2 Very Wet and Difficult Materials

At this stage, as a purely qualitative test, some wet, cohesive particles were tried. The particles were coated with a mixture of dextrin, cotton-seed protein and calcium carbonate, suspended in 5% starch solution, which is a substrate developed for solid-state culture of streptomycetes. It is usually granulated but in this case the particles were coated and then put into the rig without being dried first. The result was a mixture of particles and suspension with a sticky custard-like consistency. This was much more sticky than the material in the fermenter is expected to be, but the higher moisture content was used to compensate for the rapid drying in the bed. Using the highest air flow rates possible in the test rig, this material was lifted and circulated in the pattern
described as "toroidal" below. The circulation continued even when water was sprayed onto the top of the bed. On the basis of these results it was decided to investigate the hydrodynamics of this bed design quantitatively through further, more detailed experiments on this rig.

2.4.3 Flow Patterns In The Prototype Rig

Several circulation patterns have been observed in the prototype bed. These all develop from the superimposition of the standard spout-bed flow (Fig. 2.18a), and the circular flow induced by the secondary air (Fig. 2.18b). With a bed of dry particles these can be combined to give a spouted flow in which the downward path of particles in the annulus is spiral; the relative magnitude of the vertical and horizontal velocity components can be varied by adjusting the two air flows (Fig. 2.18c). This was also achieved in a flat-bottomed bed with a central spout. With the shallower bed of heavily coated particles, the flow was toroidal, i.e. in the form of a closed, horizontal, spiral path travelling up at the centre and down at the walls (Fig. 2.18d).

With a deep bed of the dry particles there is a tendency for the bed to segregate into a lower rotating section and an upper fountain (Fig. 2.18e). This effect is thought to be due to two forces acting on the particles: the electrostatic attraction between the particles and the perspex walls, and the drag force due to the upward flow of air in the annulus. Both of these oppose the down-flow of particles in the annulus, leading to the short-circuiting across the top of the bed. The electrostatics are irrelevant to the fermenter design, as the bed will be wet and humid, but the flow of secondary air poses a problem. The upward flow is reduced if there is no fluidising flow from the base of the annulus as seen in figure 2.18d; therefore the decision to use fluidising air must compare the benefit of aerating all the bed thoroughly to any benefit achieved by rapid recirculation.
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In deeper beds of dry particles the spout is submerged, and the upper part of the bed forms a bubbling fluidised bed; this can be regarded as a well mixed system in which the fluidised region exchanges material with the regularly circulating material in the base of the bed. The use of a concentric draught tube may promote the transfer of particles from the lower part of the bed to the surface, and improve the mixing throughout the bed.

2.5 Conclusions

Fermentation applications require a three-phase fluidised bed in which the particles are well mixed and the fermentation parameters such as temperature and moisture content are closely controlled. The particles in a fluidised bed fermenter need to be irrigated to maintain the moisture levels, and may become adhesive and cohesive due to the presence of both water and polysaccharides in the substrate and produced by the micro-organisms.

Standard designs for fluidised beds and spouted beds, and alternative designs available in the literature, are not effective for this type of application. The main problems are poor mixing and the development of static regions and aggregates in the base of the bed.

A novel bed has been designed which addresses these problems by introducing high velocity air jets at the base. These prevent the formation of static zones and break up any aggregates. In the next chapter, the hydrodynamics of this design will be discussed, and a model which can be used for scale-up will be introduced.
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Figure 2.1 Examples of Test Particles

a) cylinders, 38 ppi

b) cylinders coated with nutrient medium

c) coated cylinders after autoclaving

d) Puffed Wheat
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Figure 2.2 Fluidised Bed
Chapter 2: Preliminary Tests on Bed Designs

Figure 2.3 Bar chart of Preliminary Results: Fluidised Bed

Preliminary Bed Designs

Fluidised Bed

Superficial Velocity (m/s)

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<th>dry cylinders</th>
<th>coated cylinders</th>
<th>wet cylinders</th>
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- Initiate motion
- Stable mobilisation
Chapter 2: Preliminary Tests on Bed Designs

Figure 2.4 Bar chart of Preliminary Results: Fluidised Bed with Modified Distributor

Preliminary Bed Designs
Fluidised Bed: Small Distributor holes

![Graph showing fluidised bed designs with superficial velocity (m/s) for dry, coated, and wet cylinders, indicating initiate motion and stable mobilisation.](image-url)
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Figure 2.5 Spouted Bed
Chapter 2: Preliminary Tests on Bed Designs

Figure 2.6 Bar chart of Preliminary Results: Spouted Bed

Preliminary Bed Designs

Spouted Bed

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dry cylinders  coated cylinders  wet cylinders

- Initiate motion
- Stable mobilisation
Chapter 2: Preliminary Tests on Bed Designs

Figure 2.7 Tumbling Bed

14.5 mm

10 mm

14.5 mm
Chapter 2: Preliminary Tests on Bed Designs

Figure 2.8 Whirling Bed
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Figure 2.9 Bar chart of Preliminary Results: Whirling Bed
Figure 2.10 Annular Spouted Bed
Chapter 2: Preliminary Tests on Bed Designs

Figure 2.11 Bar chart of Preliminary Results: Annular Spouted Bed
Figure 2.12 Legros Bed

MOBILISING DEVICE

a) distributor
b) mobilising orifices \( (\phi 5 \text{ mm}) \)
c) accelerating orifice \( (\phi 25 \text{ mm}) \)
d) mobilising air supply
e) accelerating air supply
f) plenum chamber
Chapter 2: Preliminary Tests on Bed Designs

Figure 2.13 Comparative Bar Chart: Results with Dry Particles

Preliminary Bed Designs
Mobilisation of Dry Particles

<table>
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<th>Small Fluid</th>
<th>Conical</th>
<th>Whirling</th>
<th>Spouted</th>
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Legend:
- / Initiate motion
- - Stable mobilisation
Chapter 2: Preliminary Tests on Bed Designs

Figure 2.14 Comparative Bar Chart: Results with Coated Particles

![Bar Chart for Preliminary Bed Designs](image)

**Preliminary Bed Designs**

Mobilisation of Coated Particles

- Fluid
- Small fluid
- Conical
- Whirling
- Spouted

- **Initiate motion**
- **Stable mobilisation**
Chapter 2: Preliminary Tests on Bed Designs

Figure 2.15 Comparative Bar Chart: Results with Wet Particles

Preliminary Bed Designs
Mobilisation of Wet Particles

<table>
<thead>
<tr>
<th>Superficial Velocity (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>fluid</td>
</tr>
<tr>
<td>small fluid</td>
</tr>
<tr>
<td>conical</td>
</tr>
<tr>
<td>whirling</td>
</tr>
<tr>
<td>spouted</td>
</tr>
</tbody>
</table>

[Bars for each type of fluid motion are shown, with categories: fluid, small fluid, conical, whirling, spouted.]

- [ ] Initiate motion
- [ ] Stable mobilisation
Chapter 2: Preliminary Tests on Bed Designs

Figure 2.16 Prototype Bed Design
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Figure 2.17 Alternative Jet Orientations

(a)

(b)
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Figure 2.18 Flow Patterns in the Prototype Bed

a) Spout Flow Only
b) Jet Flow Only
c) Combined Flows
d) Toroidal Flow Pattern
e) Separation of Bed into Two Flow Zones
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3.1 Introduction
In order to investigate the hydrodynamics of the new design a series of qualitative tests was done. The prototype was used to measure the relationship between the amount of air required through the jets to mobilise the bed, and a number of variables. These included the amount of air supplied through the base, and the bed depth, bulk density and the number and size of the jet inlets.

This chapter covers the initial investigative experiments which established quantitively the relationship between the air flow supplied to the bed and the motion in the bed, particularly at the point at which motion was initiated. Based on these initial results, a theory was developed to predict the mobilisation point, and this is also discussed. A series of more focussed experiments was then done to validate the theory. In addition a second analysis is proposed, which concerns the interaction between the multiple jets in the base of the bed; this has implications for the scale-up of the design. These analyses have been used to calculate the mobilisation point for a larger prototype which was built to study the heat-transfer behaviour of the bed. The work on the larger unit is discussed in detail in chapter 4.

3.2 Prototype Rig
The prototype design is shown in figure 3.1, and consists of a perspex column of 145 mm diameter and 1 m height. Two types of base plate were used: a perforated plate distributor with 2 mm holes on a 14 mm triangular pitch, giving 1.4% free area, and a flat plate with a single 25 mm diameter central orifice, which is referred to below as the spout-plate distributor. Additional air inlets in the column wall allow jets of air to be injected into the bed. In the prototype design, four inlets, each of 10 mm internal diameter, were used; these were located at a height of 25 mm above the distributor and set at an angle of 45° to the radial direction as shown in figure 3.1.
Air for the wall inlets was supplied through a common manifold. The air supplies to the distributor and to the manifold could be controlled independently and, in addition, one or more of the inlets in the wall could be shut off.

As well as the two distributor plate designs, two other modifications to the rig have been tested. Firstly, a second column was made, in which the inlets were not horizontal but were oriented at 45° to the horizontal in a downwards direction, but with the angle between the inlets and the radial direction unchanged. Secondly, inserts were made for the wall inlets which reduced the diameter to 5 mm; these could be used with either column.

Two types of particle were used for the initial experiments. Polyester foam cylinders, 5 mm in diameter by 5 mm long, are the most promising support for the fermenting organism. For this work they were coated with a typical nutrient suspension and dried before use. "Puffed Wheat", which has a similar bulk density and a characteristic dimension of 8 mm, was used as an alternative particle, as it has the advantage that it requires no preparation and produces less fines: the loss of fines from the biological support particles had a significant effect on the bulk density during experiments. In a fermenting system the presence of water is expected to eliminate most of these losses, by making the particles less brittle.

3.3 Initial Experiments

3.3.1 General Observations

In the initial experiments the particle flow patterns in different configurations were studied qualitatively, by visual observation. As described in Chapter 1, cohesive and angular particles cannot be stably fluidised or spouted in conventional bed designs. This is observed in the prototype rig when no air is introduced through the wall inlets, particularly with the coated particles or with wetted particles. When air is supplied to the manifold it enters
the column through the inlets as four equal jets; the action of
the jets on the bed causes it to rotate about a vertical axis.
This rotation prevents the formation of areas of stagnant
particles on the base or against the walls of the bed, so that the
bed is not subject to defluidisation like a conventional fluidised
bed or to the channelling observed in a spouted bed. When air is
introduced through both the base plate and the walls of the bed,
stable motion is achieved with a range of combinations of the two
flows.

3.3.2 The Effect of Specific Parameters
The various flow regimes are shown in figure 3.2. The behaviour
in the bed can be described as the rotation induced by the action
of the jets superimposed on the bed behaviour associated with the
distributor type. Thus, with the perforated plate distributor,
the result is a rotating fluidised bed, while with the spout-plate
distributor a rotating spouted bed is obtained.

Bed Depth
The behaviour of the bed also varies with the depth of particles
used. For very shallow beds, where $H/D$, the height to diameter
ratio, is low, i.e. $H/D < 0.5$, and with either distributor plate,
the centrifugal action pushes the rotating particles outwards
against the walls of the column, and some or all of the air
entering through the distributor bypasses the bed through a
central void. For deeper beds, i.e. $H/D > 1$, two distinct
regions can be seen: a lower region where the jets are the major
influence on motion, and a region where the vertical flow of air
is predominant. Any central void does not extend to the surface
of the bed. With the perforated plate distributor, the upper part
of the bed is a slowly rotating bubbling fluidised bed while the
bottom of the bed is spinning rapidly. The boundary is not
distinct, there being a steady decrease in the horizontal
(spinning) velocity with height above the jets. With the spout
plate distributor there is a similar decrease in the horizontal
velocity with height, and the action of the spout imposes a steady
circulation on the particles. Particles are carried up through the diffuse, central, spout region and deposited on the top of a denser, annular region to return to the bottom of the bed following a spiral path. With a shallower bed and where the greater proportion of the air is introduced through the distributor, the frequency of vertical rotation is less than the frequency of the horizontal rotation.

Jet Diameter
When the jet diameter is reduced and all other parameters are kept constant, the volumetric airflow through the jets required to initiate motion in the bed is reduced.

Jet Orientation
The orientation of the jets also affects the amount of air required for mobilisation. When the jets are inclined downwards the air flow-rate required to initiate motion is higher. With the spout-plate distributor, the circulation time is reduced, as the downward velocity at the wall is increased.

Number of Jets Used
When fewer jets are used the minimum air flow rate required to initiate some visible motion of the particles in the bed is lower, all other parameters being constant. However, steady rotational motion is less likely to be established, and a range of behaviours may be observed, including channelling and the formation of a spout at the wall with purely local particle movement. If too few jets are used, any rotation which is established is unstable and channels rapidly develop; the air then bypasses the bed at the walls and motion stops. The instabilities are worst with only one jet used, and worse if two neighbouring jets are used compared to two opposite jets.

3.4 Quantitative Experiments
The first quantitative experiments were designed to provide data for comparison between the prototype design and a conventional
fluidised bed. In particular, it was decided to measure the superficial velocity at the mobilisation point and compare this with the minimum fluidisation velocity. For these experiments the coated particles were used.

3.4.1 Experimental Method
In each experiment the bed was set up and filled with a measured quantity of particles and the air flow rate through the distributor, $Q_D$, was set. The airflow rate to the jets was then slowly increased and the point at which the particles were first observed to move was noted. The mobilisation point was determined in this way for a range of flows through the distributor, from no flow to the minimum fluidisation velocity. The experiment was then repeated with different bed depths. The initial results are plotted in figure 3.3.

3.4.2 Results
The results show a considerable amount of scatter; this is partly due to the qualitative nature of the observations and partly due to inhomogeneities in the bed. The latter lead to maldistribution of the air from the wall inlets and, in the worst case, the bed fails internally allowing a channel to form so that some of the introduced air bypasses the bed; the airflow required through the wall inlets to initiate motion is then correspondingly higher.

Despite the amount of scatter it is still possible to observe trends in the data. The amount of air required for mobilisation increases markedly with depth when no air is supplied through the distributor, but this difference becomes negligible as the bed approaches the fluidised state.

The total sum of the air flow through the jets and the base is greater than the minimum fluidisation velocity, (0.43 ms$^{-1}$) at all points, see also section 2.2.3. However, as the particles do not fluidise well in the conventional manner, the comparison is theoretical, and the design should be assessed on the ease with
which difficult particles can be processed.

The total air required is plotted as a function of the air supplied through the base in figure 3.4. The least amount of air is required when no air is supplied through the base. A maximum occurs when the flow through the base is approximately $U_{mf}$ (0.007 m$^3$s$^{-1}$), the theoretical minimum fluidisation velocity. This variation in the total air flow required through the bed in order to mobilise it contrasts with the behaviour in a fluidised bed, in which the fluidisation velocity is independent of bed depth. The theoretical value of the minimum fluidisation velocity appears to be of little use in predicting the behaviour of this type of particle.

The pressure drop across a packed bed is a function of the superficial air velocity. Therefore, as the superficial air velocity is not constant at the mobilisation point, but varies according to the division of the air between the distributor and the jets, the pressure drop at mobilisation is not constant, even when the bed depth remains the same.

In addition, figures 3.3 and 3.4 show that the air flow required for mobilisation increases with bed depth. Therefore the mobilisation point is a function of the bed geometry as well as the fluid and particle properties.

In comparison the theoretical minimum fluidisation velocity is independent of the bed geometry. This indicates that the relationship between the air flow and the mobilisation point will not be similar to the relationship which defines fluidisation, which is discussed in chapter 2.

The variation in the mobilising flow through the jets with bed depth when no air is supplied through the base is shown in more detail in figure 3.5; in this case "Puffed Wheat" was used. This shows an upward trend with bed depth, but the value appears to
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tend towards an asymptote, at a value approximately equal to the minimum fluidisation velocity.

3.4.3 Conclusions from the First Quantitative Experiments
The different configurations can be compared quantitatively on the basis of the point at which the bed starts to move, which in most cases is the point at which the bed starts to rotate. This "mobilisation point" is reasonably reproducible for a given set of conditions in the bed and can be compared to the point of incipient fluidisation in a fluidised bed. However no simple relationship between the pressure drop across the bed and the bed weight is observed at this point, as would be seen in fluidised systems. Despite the lack of a measurement other than visual observation to determine this point, it is an important parameter, because stable mobilisation of the bed is observed at flow rates only marginally above the initial mobilisation point, and hence it defines the minimum operating conditions for a mobilised bed.

Because the mobilisation point can be used to predict suitable operating conditions in a "mobilised" bed, a correlation or analytical solution which can be used to predict this point provides a key to successful scale-up of this design of bed. As there are important differences between the minimum mobilisation velocity and the minimum fluidisation velocity, which indicate that the bed geometry has a marked effect on the former, the approach taken was based on an analysis of the stresses acting in the bed, rather than a modification of the theory of fluidisation, which is based solely on fluid and particle properties.

The theory which is presented below is based on the frictional stresses at the bed wall which resist motion, and the stresses exerted on the bed material by the flow of air through the jets, which promote rotation. At the same time the flow of air upward through the bed also exerts a stress on the bed and this can be defined as a pressure gradient which is a function of the superficial air velocity in the bed. By calculating the magnitude
of all the stresses acting in the bed, the point at which the resistance due to friction is balanced by the stress exerted by the jets can be found: this defines the mobilisation point.

3.5 Analysis of the Stress Distribution in the Bed
3.5.1 Forces acting in the Bed
When the jets enter the bed they exchange momentum with it, and hence impose a torque. Rotation of the bed is opposed by friction between the bed and the wall and base. Therefore we can propose the hypothesis that the bed rotates in a horizontal direction when the torque applied by the air flow through the jets is sufficient to overcome all resistance to motion due to friction or adhesion at the wall. We also make the assumption that, at the point of mobilisation, friction is mobilised over the entire wall and base surfaces of the bed and failure is over the entire wall surface.

Therefore at the mobilisation point:

\[ \text{Resisting torque due to friction on walls and base} = \text{Torque imposed by gas flow (3.1)} \]

3.5.2 Force exerted by the jets
The force exerted by the jets can be calculated if we assume that all the momentum of the gas is transferred to the bed, i.e. the angular momentum of the air leaving the bed is negligible. The torque arises due to this force acting at a distance \( y \) from the bed axis, where \( y \) is the perpendicular distance between the axis and the direction in which the jets enter the bed.

The torque exerted by the gas flow if all the momentum is exchanged is given by:

\[ T = N \times \frac{Ap\mu^2}{x} \times y \quad (3.2) \]

Torque number rate of change of momentum torque
of jets at a single jet arm
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If the total gas flow is \( Q \), then

\[
    u = \frac{Q}{N A} \tag{3.3}
\]

and therefore

\[
    T = \rho N A y \left( \frac{Q}{N A} \right)^2 = \frac{\rho Q^2 y}{NA} \tag{3.4}
\]

where \( y = R \sin 45^\circ \) and \( R \) is the bed radius, for jets set at \( 45^\circ \) to the tangent, as in this case.

**3.5.3 Torque Due to Friction**

If we assume that the friction is Coulombic, the shear stress due to friction, \( \tau_f \), acting at any point is a linear function of the normal stress \( \sigma \):

\[
    \tau_f = \mu \sigma + \tau_0 \tag{3.5}
\]

where \( \mu \) is the coefficient of friction, and \( \tau_0 \) is the adhesive stress, i.e. the frictional stress at zero normal load. For the purpose of this analysis adhesion is assumed to be negligible. Hence:

\[
    \tau_f = \mu \sigma \tag{3.6}
\]

The forces acting on the bed at the point of mobilisation are shown in figure 3.6. The friction acting at the wall can be resolved into a horizontal component \( \tau_{r\theta} \) which acts against the stress imposed on the bed by the jets, and a vertical component, \( \tau_{rz} \) which partially supports the weight of the bed. The resultant shear stress acts in a direction tangential to the surface of the bed and inclined at an angle \( \alpha \) to the horizontal.

From equation 3.5, the frictional stress is given by:

\[
    \tau_f = \mu \sigma_{rr} \tag{3.7}
\]

where \( \sigma_{rr} \) is the normal stress on the wall.
$\tau_f$ has vertical and horizontal component stresses:

$$\tau_{rz} = \tau_f \sin \alpha \quad \text{and} \quad \tau_{r\phi} = \tau_f \cos \alpha$$

(3.8)

As the vertical stress $\tau_{rz}$ is zero at the bed surface and then increases with depth, as is shown below, it is possible that the value of $\alpha$ will vary with depth. However, for this analysis, $\alpha$ has been assumed to be constant with depth. A sensitivity analysis on the results of this model, (section 3.6) shows that the value of $Q_{\text{min}}$ is not a strong function of $\alpha$, so that the exact functional form assumed for $\alpha$ is not critical.

The stresses acting on the base are shown in figure 3.7. The friction on the base acts in the opposite direction to the applied shear stress, which is tangential to the direction of motion and horizontal.

$$\tau_b = \mu_b \sigma_{zz}$$

(3.9)

3.5.4 Calculation of the Normal Stress at the Wall

The normal force acting on the walls and base has to be known in order to calculate the total resistance due to friction. The stress distribution in a bed of particles has been considered by Nedderman (1982); taking the simplest analysis, we can assume that $\sigma_{zz}$ and $\sigma_{rr}$, the normal stresses, are functions of the bed depth $z$ but do not vary with radial position, $r$.

The normal stresses $\sigma_{zz}$ and $\sigma_{rr}$ are assumed to be principal stresses in the ratio:

$$\sigma_{rr} = K \sigma_{zz}$$

(3.10)

where $K$ is a constant. The value of $K$ depends on further assumptions about the nature of the failure of the bed; Janssen (1895) used the following relationships to calculate $K$ for both passive and active failure of the bed.
In the case where the bed fails passively, i.e. the bed contracts inwards from the wall, K is given by the passive solution:

$$K = K_p = \frac{1 + \sin \phi}{1 - \sin \phi} \quad (3.11)$$

where \( \phi \) is the angle of internal friction of the material.

In the case where the failure is active, i.e. the bed expands outwards towards the wall, K is given by the active solution:

$$K = K_A = \frac{1 - \sin \phi}{1 + \sin \phi} \quad (3.12)$$

The bed will fail at one of these two limits; if for example the angle of internal friction \( \phi \) is 30°, then the values are:

$$K_A = 0.33 \quad \text{and} \quad K_P = 3.0 \quad (3.13)$$

For the particles used in these experiments a unique value of \( \phi \) was not determined. However Janssen's definition of K is useful as it allows us to compare experimental data with the two limiting cases and to then consider the validity of the estimated value of \( \phi \).

Consider an element of the bed as shown in figure 3.8. Resolving vertically we obtain:

$$\frac{\pi D^2}{4} \frac{d\sigma_{zz}}{dz} = \frac{\pi D^2 \gamma}{4} dz - \pi D \tau_{rz} \quad (3.14)$$

where \( \gamma \) is the weight density, \( \rho_b g \), and \( (\pi D^2 \gamma / 4)dz \) is the weight of the element.

Substituting for \( \tau_{rz} \) from equation (3.7) and rearranging gives:

$$\frac{d\sigma_{zz}}{dz} + \frac{4 \mu w \sigma_{rr} \sin \alpha}{D} = \gamma \quad (3.15)$$
And using the substitution for \( \sigma_{rr} \) from equation (3.10) we obtain:

\[
\frac{d\sigma_{zz}}{dz} + 4 \frac{\mu K}{D} \sigma_{zz} \sin \alpha = \gamma
\]  

(3.16)

This can be integrated to give \( \sigma_{zz} \) in terms of the bed depth, \( z \). In the general case we consider the effect of a surcharge, \( Q_0 \), on the stresses in the bed, i.e. \( \sigma_{zz} = Q_0 \) at the bed surface.

Rearranging gives:

\[
\frac{d\sigma_{zz}}{dz} = \gamma \left( 1 - a \sigma_{zz} \right)
\]  

(3.17)

where

\[
a = \frac{4 \mu K \sin \alpha}{\gamma D}
\]

hence:

\[
\frac{d\sigma_{zz}}{(1 - a \sigma_{zz})} = \gamma \, dz
\]  

(3.18)

Using the boundary conditions: \( z = 0 \) \( \sigma_{zz} = Q_0 \) \( z = z \) \( \sigma_{zz} = \sigma_{zz} \)

and integrating as follows:

\[
\int_{Q_0}^{\sigma_{zz}} \frac{d\sigma_{zz}}{(1 - a \sigma_{zz})} = \int_0^z \gamma \, dz
\]

(3.19)

we obtain:

\[
\frac{1}{-a} \left[ \ln \left(1 - a \sigma_{zz} \right) \right]_{Q_0}^{\sigma_{zz}} = \left[ \gamma z \right]_0
\]

(3.20)

and hence:

\[
\frac{1}{-a} \left( \ln \frac{(1 - a \sigma_{zz})}{(1 - aQ_0)} \right) = \gamma z - 0
\]

(3.21)

\[
\frac{(1 - a \sigma_{zz})}{(1 - aQ_0)} = \exp (-a \gamma z)
\]

(3.22)
which can be rearranged to give:

\[
\sigma_{zz} = \frac{1}{a} \left( 1 - \exp(-a \gamma z) \right) + Q_0 \exp(-a \gamma z) \quad (3.24)
\]

Substituting back for \( a \), gives:

\[
\sigma_{zz} = \frac{\gamma D}{4 \mu K \sin \alpha} \left[ 1 - \exp \left( \frac{-4 \mu K \sin \alpha}{D} \right) z \right] + Q_0 \exp \left( \frac{-4 \mu K \sin \alpha}{D} z \right)
\]

(3.25)

With no surcharge the second term on the right-hand-side is zero.

At great depths, as \( z \to \infty \) the exponential terms disappear and hence:

\[
\sigma_{zz} \to \frac{\gamma D}{4 \mu K \sin \alpha}
\]

(3.26)

and from the definitions of \( K \) and \( \mu_w \):

\[
\sigma_{rr} \to \frac{\gamma D}{4 \mu_w \sin \alpha} \quad \tau_f \to \frac{\gamma D}{4 \sin \alpha}
\]

(3.27)

In the case where \( \tau_{r\theta} \) is zero, (i.e. no flow through the jets and therefore no rotation), \( \alpha = 90^\circ \) and the solution for \( \sigma_{zz} \) reduces to:

\[
\sigma_{zz} = \frac{\gamma D}{4 \mu_w K} \left[ 1 - \exp \left( \frac{-4 \mu K}{D} z \right) \right] + Q_0 \exp \left( \frac{-4 \mu K}{D} z \right)
\]

(3.28)

which is the result obtained by Janssen (1895) for the failure of granular materials in hoppers.

Janssen's analysis and more recent analyses are discussed by Nedderman (1982), who indicates that more precise predictions of
the stress distribution would be obtained by using a more rigorous method such as the method of characteristics. However this method also depends on assumptions which are not easily verified, and therefore, in the light of the assumptions already made, the improved accuracy of the predictions does not justify the use of this more complicated analysis.

3.5.5 Calculation of the Torque Due to Friction
Having defined the stress distribution, we can calculate the torque due to friction which resists rotation of the bed.

Torque on Base
The friction on the base is due to the vertical stress exerted on the base by the bed weight. If we consider a small annular element of radius \( r \), where \( 0 < r < R \), and thickness \( dr \), then the torque acting on the element is \( 2\pi r^2 \tau_b dr \). The value of \( \tau_b \), the shear force on the base, is given by:

\[
\tau_b = \mu_b \left( \sigma_{zz} \right)_{z=H} \tag{3.29}
\]

As \( \sigma_{zz} \) is not a function of \( r \), integration of this expression between \( r = 0 \) and \( r = R \), gives the following expression for the total torque on the base:

\[
T_b = \frac{\pi D^3}{12} \mu_b \left( \sigma_{zz} \right)_{z=H} \tag{3.30}
\]

We can then calculate the torque on the base by inserting the value of \( \sigma_{zz} \) at \( z = H \).

\[
T_b = \frac{\pi D^3}{12} \mu_b \left( \frac{\gamma D}{4\mu K\sin\alpha} \left[ 1 - \exp \left( \frac{-4\mu K\sin\alpha}{w} \right) \right] \right)
+ Q_0 \exp \left( \frac{-4\mu K\sin\alpha}{w} \right) \tag{3.31}
\]
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Torque on Wall
To calculate the torque acting at the wall we have to consider the variation in the radial stress with depth. If we consider a horizontal element of the bed, depth dz, distance z from the top of the bed, friction acts at the wall over the area π D dz. For the cylindrical bed the torque arm radius is half the diameter, D/2. Therefore the torque required to overcome the friction at the wall is given by:

\[ dT = \frac{\pi D^2}{2} \tau_{r\theta} \, dz \]  

(3.32)

If we substitute for \( \tau_{r\theta} \) (equation 3.8), and then for \( \sigma_{rr} \) (equation 3.10), we get:

\[ dT = \frac{2\pi D^2}{4} \mu_w K \sigma_{zz} \cos \alpha \, dz \]  

(3.33)

Integrating over the bed depth H gives:

Total torque on wall =

\[ \frac{\pi D^3 \gamma}{8 \tan \alpha} \left\{ H - \frac{4 \mu W K \sin \alpha}{D} \left[ 1 - \exp \left( \frac{-4 \mu W K \sin \alpha}{D} \right) \right] \right\} \]

\[ + \frac{2\pi D^2}{4} \mu_w K \cos \alpha \left\{ \frac{4 \mu W K \sin \alpha}{D} Q_0 \left[ 1 - \exp \left( \frac{-4 \mu W K \sin \alpha}{D} \right) \right] \right\} \]

(3.34)

Hence the total torque due to friction acting on the walls and base of the bed is the sum of the two expressions given in equations 3.31 and 3.34.

3.5.6 The Effect of Drag Forces in the Bed
The above solution applies to the torque required to rotate a static bed of material in a cylindrical hopper where the weight of the bed is wholly supported by the walls and base. In the prototype bed the torque is applied by the action of air jets in the base of the bed; therefore there is an additional force acting on the bed due to the drag exerted by the upward flow of air.
through the bed. If air is also supplied through the base of the bed the total air flow through the bed must be accounted for in the force balance.

The drag force, due to the gas entering at the base, acts vertically and will support part of the weight of the bed, reducing the weight which is supported by the walls and base. In the extreme case the whole weight is supported and the bed fluidises. As is well known from work with packed and fluidised beds, the effect of the drag force on the bed weight can be measured as the pressure drop per unit height. Therefore to calculate the stress distribution correctly when air is flowing through the bed, we must account for the variation in the effective bed weight $\gamma$ by substituting a function of $U$, the superficial gas velocity in the bed.

The drag force in the bed is a function of the gas and particle properties and can be expressed as a function of the particle Reynolds number

$$\text{Re} = \frac{\rho_f U d_v}{\mu_f} \quad (3.35)$$

and the superficial velocity in the bed, $U$.

For Reynolds numbers greater than 1, as in this case, the Ergun equation can be used to calculate the pressure-drop-flow relationship:

$$\frac{\Delta P}{H} = f(\epsilon) \frac{\mu_f U^2}{d_v} + g(\epsilon) \frac{\rho_f U^2}{d_v} \quad (3.36)$$

Under laminar conditions the first term dominates and $\Delta P/H$ is a function of $U$. If the Reynolds number is large, the second term dominates and $\Delta P/H$ is a function of $U^2$. For intermediate flow both terms are significant. In this case the Reynolds number is 233, which is intermediate:
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To produce a more convenient functional form, we can simplify the equation to a power law, i.e.:

\[ \frac{\Delta P}{H} = K U^n \]  \hspace{1cm} (3.37)

where \( 1 \leq n \leq 2 \)

where \( K \) is a function of the fluid and particle properties and therefore constant in these experiments.

We can then measure the variation of \( \Delta P/H \) with \( U \) for the range of values of interest here and fit a straight line through a log-log plot of \( \Delta P/H \) v \( U \). The value of \( n \) is given by the slope of the graph and \( K \) can be calculated from the intercept with the \( x \) axis.

For the prototype bed, \( \Delta P/H \) has been measured and is given for puffed wheat by:

\[ \frac{\Delta P}{H} = 1452 \ U^{1.57} \]  \hspace{1cm} (for the range \( 0.199 \text{ ms}^{-1} \leq U \leq 0.520 \text{ ms}^{-1} \))  \hspace{1cm} (3.38)

and for dry coated particles by:

\[ \frac{\Delta P}{H} = 281 \ U^{1.35} \]  \hspace{1cm} (for the range \( 0.381 \text{ ms}^{-1} \leq U \leq 1.515 \text{ ms}^{-1} \))  \hspace{1cm} (3.39)

The results are shown in figures 3.9 and 3.10.

This modification to the effective bed weight is only valid if the air flow in the bed is evenly distributed. To test this assumption, the pressure-drop:flow relationship was measured for two cases: where the air is introduced evenly through a perforated distributor, and where it is introduced through a single central inlet. The results are plotted together in figure 3.11 and show that the effect of changing the point at which air is introduced has a negligible effect on the pressure-gradient:flow relationship in the prototype bed.

We can apply this modification to estimate the stress distribution
in the bed when air is supplied through the base of the bed or through the jets. If we consider the effect of air addition through the distributor on the stress distribution, we can substitute a form of $\gamma$ which is a function of the total flow rate through the bed.

$$\gamma_{mod} = \gamma - \frac{\Delta P}{H}$$  \hspace{1cm} (3.40)

or

$$\gamma = 1452 \ U^{1.57} \text{ (puffed wheat)}$$

$$\gamma = 281 \ U^{1.35} \text{ (coated particles)}$$

**Surcharge Form of the Equations**

When air is supplied through the jets, the modified form of $\gamma$ applies above the jets, Fig. 3.12; therefore, to calculate the frictional torque on the wall above the jets, equations 3.31 to 3.34 are used, taking the modified value of $\gamma_{mod}$ from equation 3.40 and setting $Q_0 = 0$. Below the jets the value of $\gamma$ is unaffected by the gas flow through the jets; therefore either $\gamma = \rho_b g$ as before, or, in the case where part of the air is supplied through the base, a different value of $\gamma_{mod}$ applies. The conditions in the base of the bed can therefore be estimated by recalculating the stresses from the level of the jets, using the appropriate value of $\gamma$ and accounting for the effect of the material above the jets by using the surcharge term:

$$Q_0 = \left( \sigma_{zz} \right)_{z=H-h_l}$$  \hspace{1cm} (3.41)

The total torque on the wall, $T_w$, is given by the sum of two terms, $T_{wa}$ and $T_{wb}$ for the regions above and below the jets.

We can then calculate $(\sigma_{zz})_{z=H}$ and hence $T_b$, the torque on the base.

The full equations are given below:
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Above the height of the jets:

\[
\left( \sigma_{zz} \right)_{z<(H-h_1)} = \frac{\gamma_{\text{mod}}}{R} \left[ 1 - \exp\left( -R \ z \right) \right] + Q_0 \exp\left( -R \ z \right) \quad (3.42)
\]

where \( R = \frac{4 \mu \ k \ \sin \alpha}{D} \)

So for \( z = (H-h_1) \), where \( h_1 \) is the distance between the jets and the distributor, setting \( Q_0 = 0 \), we obtain:

\[
\left( \sigma_{zz} \right)_{z=(H-h_1)} = \frac{\gamma_{\text{mod}}}{R} \left[ 1 - \exp\left( -R \ (H-h_1) \right) \right] \quad (3.43)
\]

Hence the torque on the walls above the jets is given by:

\[
T_{wa} = \frac{\pi D^3}{8 \ \tan \alpha} \ \gamma_{\text{mod}} \left( (H-h_1) - \frac{1}{R} \left[ 1 - \exp\left( -R(H-h_1) \right) \right] \right) \quad (3.44)
\]

Below the level of the jets, recalculating the vertical stress from the point \( z = (H-h_1) \) and substituting:

\[
z_1 = (z-(H-h_1)) \quad \text{and} \quad Q_0 = \left( \sigma_{zz} \right)_{z=(H-h_1)}
\]

We obtain:

\[
\left( \sigma_{zz} \right)_{z>(H-h_1)} = \frac{\gamma}{R} \left[ 1 - \exp(-R(z_1)) \right] + \left( \sigma_{zz} \right)_{z=(H-h_1)} \exp(-R(z_1)) \quad (3.45)
\]

and the torque on the wall below the jets is given by:

\[
T_{wb} = \frac{\pi D^3}{8 \ \tan \alpha} \left\{ \gamma \left\{ h_1 - R \left[ 1 - \exp\left( -R \ h_1 \right) \right] \right\} 
\right. \\
\left. \quad - \gamma_{\text{mod}} \left[ 1 - \exp\left( -R(H-h_1) \right) \right] \left[ 1 - \exp\left( -R \ h_1 \right) \right] \right\} \quad (3.46)
\]

Finally, the torque on the base is given by:

\[
T_b = \frac{\pi D^3}{12} \ \mu_b \ (\sigma_{zz})_{(z=H)} \quad (3.47)
\]
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\[ T_b = \frac{\pi D^3}{12} \mu_b \left( \frac{\gamma}{R} \left[ 1 - \exp(-R h_1) \right] + \frac{\gamma_{\text{mod}}}{R} \left[ 1 - \exp(-R(H-h_1)) \right] \exp(-R h_1) \right) \]  

(3.48)

Therefore we can calculate the stress distribution over the entire wall and base of the bed, and hence the frictional torque which has to be overcome to mobilise the bed. The variation in the normal stress on the wall (\(\sigma_{rf}\)) with height is plotted in figure 3.13 for the case where there is no flow through the base and a nominal flow through the jets.

3.5.7 Torque Balance

It can be seen that both the torque exerted by the jets and the torque due to friction are functions of \(Q_j\), the volumetric flow rate through the jets. Also, an increase in \(Q_j\), and therefore \(U\), reduces the torque due to friction through the effect on \(\gamma\).

Therefore, if we plot the torque exerted by the jets, and the torque due to friction for various bed heights, against the volumetric flow rate, \(Q_j\), the locus of the intersections of the torque due to friction and the frictional torque for each bed height gives the predicted value of the variation in the minimum mobilising velocity, \(Q_{\text{min}}'\), with bed height, (figure 3.14).

For any set of parameters the minimum flow rate for motion, \(Q_{\text{min}}'\), can be found graphically or by solving the equations numerically. The results are subject to two uncertainties. Firstly, the failure may be either active or passive; therefore the value of \(K\) is uncertain, though the two limiting cases can be calculated. This provides a range of values between the two limiting cases with which the experimental results can be compared. It should be noted however that the value of \(K\) also depends on the internal angle of friction, which is another parameter which is not easily measured for this particular material. Secondly the value of \(\alpha\), the direction of the net force at the wall is unknown, and it is...
possible that $\alpha$ will vary with depth in the bed. The effect of the assumptions made about the value of $\alpha$ can be estimated by a sensitivity analysis, calculating the variation in the theoretical result with variations in $\alpha$; this is discussed in the next section.

### 3.5.8 Parameters used in the Theoretical Model

For comparison with experimental results a number of the parameters in the theoretical model have been measured. Those that have not been measured have been estimated on the basis of the assumptions discussed below.

#### Wall Friction Coefficients

The coefficients of friction at the wall (perspex), and on the base (perforated aluminium), were measured using a hinged plate as shown in figure 3.15. The plate was lifted until the particles resting on it began to slide and the coefficient was taken to be $\tan \beta$ at that point. The measured values are shown in Table 3.1.

<table>
<thead>
<tr>
<th>Material</th>
<th>$\mu_w$ (perspex)</th>
<th>$\mu_b$ (aluminium)</th>
</tr>
</thead>
<tbody>
<tr>
<td>puffed wheat</td>
<td>0.33</td>
<td>0.36</td>
</tr>
<tr>
<td>coated foam</td>
<td>0.28</td>
<td>0.51</td>
</tr>
</tbody>
</table>

#### Internal Friction Coefficients

The angle of internal friction has not been determined for any of the materials used in these experiments, as there was no available equipment suitable for measuring it for particles of this size. For the purpose of the analysis the value $\phi = 30^\circ$ has been used to calculate $K_A$ and $K_P$ in each case. This is a typical value for $\phi$, and is in agreement with the empirical method of measuring the angle of repose, which gave a similar value.

#### Direction of the Net Stress at the Wall ($\alpha$)

As has been noted above, the direction of the stress at the wall
may vary with depth in the bed, as it is dependent on the vector sum of the vertical stress on the wall due to friction, which is a function of depth, and the stress due to the jets. The analysis is based on equating the force exerted by the jets with the horizontal component of the stress at the wall; this does not, however, allow us to determine the direction or magnitude of the net stress at the wall, except by making a deduction from visual observations. The effect of the value of \( \alpha \) is shown by a sensitivity analysis on the variation in \( Q_{\text{min}} \) with \( z \) (figure 3.16), for values of \( \alpha \) between 30° and 60°. As can be seen on the figure, the value of \( \alpha \) used does not have a large effect on the predicted result. Therefore for this analysis the value of \( \alpha = 45° \) was used throughout.

3.6 Detailed Quantitative Experiments

Both the early experiments and the analysis indicate that a major factor affecting the point of mobilisation and the value of \( Q_{\text{min}} \) is the depth of the bed. Therefore the more detailed quantitative experiments were based on measuring the variation in \( Q_{\text{min}} \) with bed depth, and the range of experiments carried out also examined the effect of other parameters, such as the particle bulk density and the jet diameter and orientation. The majority of these experiments were performed using either puffed wheat or coated foam particles. \( Q_{\text{min}} \) is defined as the amount of air which has to be supplied to the bed to initiate rotation.

The data from these runs are plotted alongside the theoretical values for comparison. As the value of \( K \) is uncertain, both the active and passive solutions are plotted, using the assumed value of 30° for \( \phi \), the internal angle of friction. The value of \( \alpha \) is assumed to be 45° throughout the bed.

3.7 Comparison of Results with Theory

3.7.1 General Observations

Figure 3.17 shows the variation in mobilising velocity with height for puffed wheat, the experimental data falling between the two
limiting theoretical cases, indicating that the theory can predict the way in which the mobilisation point varies with bed depth. In fig. 3.18 the equivalent data for coated particles are shown. In this case the data tend towards the active solution of the equations. However in the light of the degree of experimental error it is difficult to draw a final conclusion from the data presented here.

3.7.2 Reduced Jet Diameter

For these experiments the jet diameter was reduced to 5 mm by putting perspex inserts into the jet inlets. Data for puffed wheat and coated particles are shown in Figs. 3.19 and 3.20. The main effect of reducing the jet diameter is to reduce the amount of air required to spin the bed for a given height. This effect is predicted by the theory, as the rate of change of momentum increases as the square of the air velocity, (see equation 3.3) and the air velocity through the jets is increased as the diameter is reduced.

These data are also within the range described by the two limiting cases, and in this case neither the active nor the passive solution is closer to the experimental results over the whole experimental range. However it can be observed that the results are closer to the active solution at shallow bed depths, and tend towards the passive solution with deeper beds. This may indicate a change in the mechanism of failure in deeper beds, particularly as the jets act only on the base of the bed.

A similar trend is also observed in the data in fig. 3.17.

3.7.3 Response to Variations in the Particle Type

The data for puffed wheat and coated foam particles have already been compared above, and are in similar agreement to the theoretical results. An additional experiment was done with
another type of large particle: 15 mm diameter hollow plastic balls, which are used to insulate the surface of water baths. The data from this experiment are presented in figure 3.21. A similar trend is observed in these data, with a closer fit to the active solution being observed at shallow bed depths. These results demonstrate that the overall similarity between the experimental results and the theory occurs over a range of bulk densities and particle sizes.

3.7.4 Effect of Using Inclined Jets

The results of a series of experiments using the second bed design, with the jets inclined downwards at an angle of 45°, are shown in figure 3.22. This design was tested with both the puffed wheat and the coated particles; data from both the inclined jets and the horizontal jets are plotted for comparison.

From the theory it is expected that using inclined jets will increase $Q_{\text{min}}$, as the horizontal component of the torque exerted by the jets will be reduced by a factor of $\sqrt{2}$. There will also be a small increase in the vertical stress on the base due to the downward component of the rate of change of momentum from the jets, which will increase the frictional torque on the base, and an overall decrease in the frictional stresses as the airflow is increased, due to the increase in the vertical drag forces. The last two effects act in opposite directions and will cancel out to some extent.

A significant difference between the results for horizontal and inclined jets is seen with the coated foam particles, where the downward inclined jets increase the mobilisation velocity by about 200 l/min or 0.2 m/s. The data are replotted in figure 3.23, in which the data for the inclined jets have been divided by $\sqrt{2}$. For the coated foam particles the data are coincident in this figure, which shows that the change in the horizontal component has the greatest influence on $Q_{\text{min}}$; the other factors mentioned
above are either negligible, or they balance out in this case.

However with the puffed wheat the difference between the two cases is much smaller, and the two data sets differ by less than a factor of $\sqrt{2}$; therefore the modified data are not coincident. In this case the other factors may be more significant. In particular, the increase in the drag force, and concurrent reduction in the frictional resistance, may reduce the required torque.

These data highlight another difference between the two types of particles, in terms of the form of the mobilisation point curve. The theory predicts, and the experimental results demonstrate that the mobilisation velocity initially increases with bed depth, but the increase is asymptotic. Therefore there is a limiting value for $Q_{\text{min}}$, and for the prototype used in these experiments this approximates to the minimum fluidisation velocity. Over the range of bed depths used for this work, the coated particles are in the first part of the curve, where $Q_{\text{min}}$ increases rapidly with bed depth, while the puffed wheat quickly approaches the asymptotic value and $Q_{\text{min}}$ is constant for most of the experiments.

There is some evidence that the data from experiments with puffed wheat can be normalised by dividing by $\sqrt{2}$ at the smallest bed depths, while, at the larger bed depths, where the value of $Q_{\text{min}}$ has reached its asymptote, the values for the coated particles are not coincident when normalised in this manner. Therefore the results for the two particle types are not inconsistent, and the overall conclusion from this experiment is that below the asymptotic value, the horizontal component of the torque exerted by the jets is the major influence on $Q_{\text{min}}$.

The mobilisation point in the asymptotic part of the curve may be more strongly influenced by the superficial velocity in the bed, which at this point is supporting most or all of the bed weight.
rather than by the direction of the force exerted by the jets at the wall. At this point the assumption that the bed acts as a rigid body may not apply, especially in the case where the particles can be fluidised. Where the bed is fluidised, a different theory is required to predict the amount of air which is required through the jets to initiate rotation in the bed.

3.8 Conclusions
The theory proposed above provides a good model of the behaviour of beds of large light particles which are mobilised by the action of jets in the base of the bed. The theory has been tested on three types of particle. However the data available are not sufficient to discriminate between the active and passive solutions of the equations. The experimental data show that there is close agreement with the active solution for shallow beds of particles but with the deeper beds the experimental values are higher than those predicted from the active solution.

The predicted variation in the value of the mobilisation point as a function of jet diameter is also consistent with the experimental results, for two different particle types. The air flow rate required to mobilise the bed is lower with smaller jets, as the jet velocity, and hence the momentum carried by the jets, is increased by reducing the jet diameter at a constant flow rate.

The effect of using downward-inclined jets is to increase the mobilisation flow rate for the denser particles such as the coated foam particles. This is consistent with a decrease in the horizontal component of the momentum by a factor of $\sqrt{2}$; this applies only at superficial velocities below the asymptotic value (approximately equal to the fluidisation velocity). At superficial velocities close to the fluidisation velocity the
effect of inclining the jets is negligible. There appears to be no benefit in inclining the jets, though in applications in which the bed is mixed by using air through a spout in the base plate, the rate of circulation may be enhanced by encouraging downward motion of the particles at the wall.

For particles which are free to move, at superficial velocities above the minimum fluidisation velocity, the bed cannot be assumed to act as a rigid body. Therefore the above analysis cannot be used to predict the point at which rotation is initiated in a fluidised bed. In addition, for particles which are not free to move, the analysis of the stress distribution in the bed gives the result that the frictional torque reduces to zero when the weight of the bed is entirely supported by the drag forces. Therefore, in the case where the superficial velocity through the distributor is greater than the actual minimum fluidisation velocity, the predicted value of $Q_{\text{min}}$ tends to zero.

3.9 Jet Penetration Theory

3.9.1 Introduction

In the section above, in which the point of mobilisation was discussed, the bed of particles was considered to act as a rigid body. In practice however, the running condition of the bed for a commercial application would use air flow rates above the minimum mobilisation flow in order to ensure stable operation and good mixing characteristics, as is the case for commercial fluidised beds. In most cases this would also take the air flow above the theoretical minimum fluidisation velocity, either as air introduced through the base, or as an increased flow through the jets. A different analysis is therefore necessary when considering the mobilisation point of a bed which is already fluidised. One particular approach is presented below.

3.9.2 Theory

The interaction of the jets with a fluidised bed may be expected to differ from their interaction with a static bed, which, as
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discussed above, can be considered to act as a rigid body. The case of a free turbulent jet entering a stagnant volume of fluid (fig 3.24) has been well researched and it is observed that the jet dissipates its kinetic energy as turbulence in the boundary layer and entrains a volume of fluid much greater than the original volumetric flow of the jet from the surrounding fluid; equations to predict both the angle of dispersion $\alpha_d$, and the jet penetration length $L$ are available in the literature. The more complicated problem of jets in a fluidised bed has been compared to the free turbulent jet and described in a similar manner (Massimilla, 1985).

Of the two characteristic parameters the penetration length $L$ is of particular interest in our design, as multiple jets are used. By coincidence, the jets are set in the walls at an angle of $45^\circ$ to the tangent (originally to make accurate fabrication simpler) and therefore each jet is directed towards the point of origin of the next. Hence if the penetration length approaches or exceeds the inter-jet length, each jet will interact with the next and this may affect the behaviour of the bed.

Although the theory for jets in fluidised beds is not as well developed as that for free turbulent jets, a large number of studies of the jets entering a fluidised bed at the distributor have been made; unfortunately this work usually concerns vertical jets, and contributes little to this study. A few researchers, notably Zenz (1968), Merry (1971), and Shakhova (1972), have also correlated data for horizontal jets. Hence we have three correlations for the jet penetration length:

Zenz

$$\frac{L_{\text{max}}}{d_o} + 1.48 = 0.5 \log \left( \frac{\rho_p u_0^2}{\rho} \right)$$  \hspace{1cm} (3.49)

(in cgs units, gives $L$ in cm)

Zenz

Range of variables:

$\begin{align*}
&d_p \quad 0.05-2.0 \text{ mm} \\
&d_o \quad 8.0-19.0 \text{ mm} \\
&\rho_p \quad 2500 \text{ kg m}^{-3}
\end{align*}$
Merry
Range of variables:
\[ d_p \text{ 0.18-2.0 mm} \]
\[ d_o \text{ 2.5-14.0 mm} \]
\[ \rho_p \text{ 1000-2500 kg m}^{-3} \]

Shakhova
Range of variables:
\[ d_p \text{ 1.25-5.0 mm} \]
\[ d_o \text{ 2.0-10.0 mm} \]
\[ \rho_p \text{ 1500-2500 kg m}^{-3} \]
Merry

\[
\frac{L}{d_o} + 4.5 = 5.25 \left( \frac{\rho_f u_o^2}{(1-\varepsilon) \rho_p g d_p} \right)^{0.4} \left( \frac{\rho_f}{\rho_p} \right)^{0.2} \left( \frac{d_p}{d_o} \right)^{0.2} \quad (3.50)
\]

Shakhova

\[
\frac{L}{d_p} = 7.8 \left( \frac{\rho_f u_o}{\rho_p g d_p^{0.5}} \right) \quad (3.51)
\]

These have been plotted in figure 3.25 using the data for the puffed wheat system:

<table>
<thead>
<tr>
<th>Table 3.2 Parameters for the Puffed Wheat System</th>
</tr>
</thead>
<tbody>
<tr>
<td>orifice diameter</td>
</tr>
<tr>
<td>air density</td>
</tr>
<tr>
<td>solid density</td>
</tr>
<tr>
<td>gravitational acceleration</td>
</tr>
<tr>
<td>particle diameter</td>
</tr>
<tr>
<td>voidage</td>
</tr>
</tbody>
</table>

The differences between these correlations are marked, and it should be noted that they are derived from data from different ranges of particle sizes and densities, collected by visual observation.

3.9.3 Comparison with Experimental Results

Initial trials, see section 3.3.2, indicated that stable rotation was not established in the bed when less than four jets were used. The mobilisation becomes progressively more stable when more jets are used, particularly when the jets are opposite rather than neighbouring. These results confirm that stable rotation is a function of the jet spacing, and support the hypothesis that it is re-entrainment of part of the air from each jet, into the next, which stabilises the rotation.
The distance between two jets in the prototype jet design, (along the axis of the jet), is given by:

\[ x = \frac{R}{\cos 45°} \]

Therefore in the 145 mm diameter bed: \( x = 0.103 \) m

Comparing this length with the jet penetration lengths we can propose an alternative hypothesis for predicting the point of mobilisation in a fluidised bed. The asymptotic value of \( Q_{min} \) for puffed wheat is 0.0086 \( m^3 s^{-1} \) with horizontal jets, (see fig. 3.22). Although the correlations give different predictions, with this value of \( Q \), the jet penetration length as predicted by Merry is 0.0118 m which is comparable to the inter-jet distance.

With the inclined jets the length of the path between jets will be increased. If we assume that the jets are reflected off the base, then the total path between jets increases to 0.117 m, see figure 3.26. Using Merry's correlation, which is the middle value, and therefore representative, we can calculate the flow required for jets of these two lengths, and compare these to the experimental values:

<table>
<thead>
<tr>
<th>( L ) (m)</th>
<th>( Q_{min} ) (m(^3)s(^{-1}))</th>
<th>( Q_{min} ) (m(^3)s(^{-1}))</th>
</tr>
</thead>
<tbody>
<tr>
<td>Horizontal Jets</td>
<td>0.103</td>
<td>0.0076</td>
</tr>
<tr>
<td>Inclined Jets</td>
<td>0.117</td>
<td>0.0085</td>
</tr>
</tbody>
</table>

These data are comparable, although the jet length correlation underestimates the required flow rate in each case.
3.9.4 Discussion

If we accept that at mobilisation the jets are interacting, this has a number of implications for the design of the bed.

Firstly, re-entrainment of the gas (and some particles) from one jet into the next will stabilise the rotational flow in the base of the bed. Therefore in addition to predicting the point of mobilisation by using the torque balance discussed above, we are able to cross-check this result using the jet penetration length. This will give an indication of whether the resulting rotation is likely to be stable.

Secondly, if we want to ensure that the jets will re-entrain each other, we can use the predicted jet penetration length to fix a jet separation and orientation which is consistent with the jet penetration length for the design flow rate through the jets. This is particularly important for scale-up of the design.

Thirdly, the jet penetration length may allow us to predict the amount of air which needs to be supplied through the jets to spin a bed which is already fluidised, either by diverting part of the flow from the distributor, or by introducing additional air.

In section 3.7, it was shown that for superficial gas velocities below the minimum fluidisation velocity, the horizontal momentum of the jet is the main parameter which determines the mobilisation point. However, above the minimum fluidisation velocity, the inclination of the jets has been shown experimentally to have a smaller influence on the mobilisation point, and this indicates that the mobilisation point is not a direct function of the horizontal jet momentum. However the increase is consistent with the increased path length between jets.

If we consider the implications of the last point raised above, it may be the case that, when the weight of the bed is supported by the drag forces, the energy of the jets can be dissipated by
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turbulence and by random movement of the individual particles in the bed. Hence, for rotation to be established, it is not sufficient to use a flow rate higher than \( \dot{Q}_{\text{min}} \); but in addition, the jets must impinge on each other and re-entrain to form a complete rotational path in the base of the bed. Therefore, stable rotation cannot be established unless the jets are able to re-entrain in each other to a greater or lesser extent; this determines the combinations of jet orientation and jet length which will be effective.

3.9.5 Conclusions

Although there is some evidence, as discussed above, that the interaction of the jets is important for maintaining stable rotation in the bed, there is no reason to say that the jet penetration length has more influence on the mobilisation point than the torque exerted by the jets. The two theories are therefore complementary. The torque balance can be used to predict the mobilisation point of a static bed, which is acting as a rigid body, before it is fluidised; the jet penetration length can be used to predict the establishment of stable rotation in a fluidised bed, which is behaving as a fluid. Comparison of the two results may show whether the rotation initiated in a static bed will be stable or unstable.

3.10 Numerical Values for Scale-up to 225 mm Bed

In the next chapter the heat transfer characteristics of the spinning bed design will be discussed. For these experiments a new experimental bed was designed and built, using a stainless steel and glass bed 225 mm in diameter and 450 mm deep; this bed is operated with a bed depth of 225 mm. This represents a modest scale-up of the 145 mm diameter bed used for the initial experiments. Using the theory developed in this chapter it is possible to predict both the minimum mobilisation point and the jet penetration lengths for this new bed.
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3.10.1 Minimum Mobilisation Point

Using the same data used above for the puffed wheat particles in the smaller bed, and solving the equations graphically using a spreadsheet, the following results are obtained:

Table 3.4 Parameters for 225 mm bed and Puffed Wheat System

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>$\rho_b$</td>
<td>100 kg m$^{-3}$</td>
</tr>
<tr>
<td>$\mu_w$</td>
<td>0.325</td>
</tr>
<tr>
<td>$\mu_b$</td>
<td>0.364</td>
</tr>
<tr>
<td>$D$</td>
<td>0.225 m</td>
</tr>
<tr>
<td>$\alpha$</td>
<td>45°</td>
</tr>
</tbody>
</table>

The coefficients of friction were assumed to be similar in the larger bed.

Table 3.5 Solution of the Torque Balance for the 225 mm Bed

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Passive Solution ($K = 3.0$)</th>
<th>Active Solution ($K = 0.333$)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flowrate through jets</td>
<td>0.0234 m$^3$s$^{-1}$</td>
<td>0.0262 m$^3$s$^{-1}$</td>
</tr>
<tr>
<td>Superficial velocity in bed</td>
<td>0.588 m s$^{-1}$</td>
<td>0.659 m s$^{-1}$</td>
</tr>
<tr>
<td>$\sigma_{zz}(z = H-H_1)$</td>
<td>61.1</td>
<td>16.9</td>
</tr>
<tr>
<td>$\sigma_{zz}(z = H)$</td>
<td>83.2</td>
<td>33.6</td>
</tr>
<tr>
<td>Torque on wall above jets</td>
<td>0.078 N-m</td>
<td>0.153 N-m</td>
</tr>
<tr>
<td>Torque on wall below jets</td>
<td>0.033 N-m</td>
<td>0.035 N-m</td>
</tr>
<tr>
<td>Torque on base</td>
<td>0.090 N-m</td>
<td>0.036 N-m</td>
</tr>
<tr>
<td>Total Torque</td>
<td>0.179 N-m</td>
<td>0.224 N-m</td>
</tr>
<tr>
<td>Torque from jets</td>
<td>0.179 N-m</td>
<td>0.224 N-m</td>
</tr>
</tbody>
</table>
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The two solutions for a 225 mm deep bed are shown in Fig. 3.27 and 3.28; the solution curves for the active and passive solutions are shown in figure 3.29.

3.10.2 Jet Penetration Length

The equations for the jet penetration length can also be solved graphically, by plotting the jet length versus the jet flow rate and comparing this to the inter-jet distance given by:

\[
\text{inter-jet distance} = \frac{D}{2 \cos 45°}
\]  

(3.52)

For the inter-jet distance, 0.159 m, the three available correlations give the following results:

<table>
<thead>
<tr>
<th>Method</th>
<th>L (m)</th>
<th>Q (m³/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Zenz (1968)</td>
<td>0.159</td>
<td>0.00296</td>
</tr>
<tr>
<td>Shakhova (1972)</td>
<td>0.159</td>
<td>0.01708</td>
</tr>
<tr>
<td>Merry (1971)</td>
<td>0.159</td>
<td>0.01140</td>
</tr>
</tbody>
</table>

The results are also shown in figure 3.25.

It can be seen from the results that for the larger bed, the jet length correlations all predict a lower value of Qₘᵢₙ than that calculated from the torque balance. However the result from Merry's correlation is within the same order of magnitude. It should be noted that the jet penetration calculations are applicable to a fluidised bed, while the torque balance is based on the assumption that all the air enters the bed through the jets.

These results have not been tested directly by observation, as with the 145 mm bed. As the 225 mm bed was designed for heat transfer experiments, it was built from stainless steel, with a
Chapter 3: Hydrodynamics in the Novel Bed Design

jacketed wall; therefore it was not possible to observe the mobilisation point directly to confirm these results. However, some conclusions can be drawn from the heat transfer results, and these are discussed in detail in chapter 4.
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Figure 3.1 Prototype Design
Chapter 3: Hydrodynamics in the Novel Bed Design

Figure 3.2 Flow Regimes in Prototype Bed

a. Perforated plate distributor, deep bed
   Bubbling fluidised bed with rotation close to the distributor.

b. Perforated plate distributor, shallow bed
   Rotating spouting bed.

c. Spout plate distributor, deep bed
   Spouting bed with spiral particle flow downwards at the wall.

d. Spout plate distributor, shallow bed
   "Toroidal" flow: bed rotates about a vertical axis, and from centre to edge. Frequency of rotation about the vertical axis is less than that about the horizontal axis.
Figure 3.3 Mobilisation Point as a Function of $Q_J$ and $Q_D$ for Two Bed Depths

Mobilisation Point: Coated Particles

Jet Flow v Flow through Distributor

Flow through Jets (m$^3$/s) vs Flow through Distributor (m$^3$/s)

- $H = 110$ mm
- $H = 45$ mm

$U_{mf}$
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Figure 3.4 Total Air Required for Mobilisation vs Air Flow through Base

Mobilisation Point: Coated Particles

Total air flow through distributor

Flow through Distributor (m3/s)

- h = 45 mm
- h = 110 mm
Figure 3.5 Mobilisation Point: $Q_j$ v Bed Depth $h$: no flow through distributor

Mobilisation Point: Puffed Wheat

Air Flow v Bed Depth

Minimum mobilising air flow ($m^3/s$)

Bed Depth (m)
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Figure 3.6 Forces acting on the Bed at the Mobilisation Point
Figure 3.7  Stresses acting on the Base

\[ r, R, \sigma_z, \tau_{\phi z}, \delta_r \]
Figure 3.8  Stress Distribution in a bed of Granular Solids
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Figure 3.9 Pressure Drop per Unit Height v Superficial Velocity: Puffed Wheat

Pressure drop v Flow for Puffed Wheat

\[ \ln(\Delta P/H) \text{ v } \ln(U_b) \]
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Figure 3.10  Pressure Drop per Unit Height v Superficial Velocity: Coated Particles

Pressure drop v Flow: Coated Particles

\[ \ln(\Delta P/H) \text{ v } \ln(U_b) \]
Figure 3.11 Pressure Drop per Unit Height vs Superficial Velocity: comparison between different distributor types

Pressure Drop vs Flow Relationship
Effect of Distributor Type

\[ \ln \left( \frac{\Delta P}{H} \right) \]

- \( \square \) perforated
- \( + \) spout
- \( - \) regression

\[ \ln (U_b) \]

Values:
- 0.9
- 0.7
- 0.5
Figure 3.12 Net Vertical forces in the bed shown as a reduction in effective bulk density

\[ \gamma = \gamma_{\text{mod}} \]

\[ \gamma = \rho g \]

\[ h_1 \]
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Figure 3.13 Normal stress on the wall ($\sigma_{rr}$) vs bed depth:

$Q_j < Q_{min}$

Variation in $\sigma_{rr}$ with Depth
Jets 25 mm above base
Figure 3.14 Torque Balance: Torque v Volumetric Flowrate through Jets, no flow through the distributor.

Torque Balance (Puffed Wheat)

Torque v Superficial Velocity

Torque (N-m) : Friction and Jet Impedance

Air flow (m³/s)

H = 0.4m
H = 0.3m
H = 0.2m
H = 0.1m
H = 0.05m
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Figure 3.15 Method for Determining Coefficient of Wall Friction
Figure 3.16  Sensitivity Analysis on $\alpha$

Sensitivity Analysis on Alpha

(Alpha in Degrees)

Jet M obili ng Velocity (m/s)

0.02
0.019
0.018
0.017
0.016
0.015
0.014
0.013
0.012
0.011
0.010
0.009
0.008
0.007
0.006
0.005
0.004
0.003
0.002
0.001
0

0 0.04 0.08 0.12 0.16 0.2 0.24

bed height (m)
Figure 3.17 Experimental Results Plotted Against Theory: Puffed Wheat, d = 10 mm, no flow through the distributor
Figure 3.18  Experimental Results Plotted Against Theory: Coated foam, $d = 10$ mm, no flow through the distributor

Coated Particles, $d = 10$ mm

Experimental Results v Theory

Minimum Mobilising Velocity (m/s)

Bed Depth (m)

---
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Figure 3.19 Experimental Results Plotted Against Theory: Puffed Wheat, d = 5 mm, no flow through the distributor

Puffed Wheat, d = 5 mm

Experimental Results v Theory
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Figure 3.20  Experimental Results Plotted Against Theory: Coated Foam, $d = 5$ mm, no flow through the distributor

Coated Particles, $d = 5$ mm

Experimental Results v Theory

Minimum Mobilising Velocity (m/s)

Bed Depth (m)
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Figure 3.21 Experimental Results Plotted Against Theory:
Waterbath Balls, d = 10 mm no flow through the distributor.
PLASTIC BALLS, BALL DIAMETER = 15 MM

Experimental Results v Theory

Minimum Mobilising Velocity (m/s)
0.00 0.04 0.08 0.12 0.16 0.20 0.24 0.28
Bed Depth (m)
0.00 0.10 0.20 0.30 0.40 0.50 0.60 0.70 0.80 0.90 1.00 1.10 1.20 1.30 1.40 1.50 1.60 1.70 1.80

passive
active
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Figure 3.22 Experimental Results: horizontal and inclined jets, no flow through the distributor
Puffed Wheat d = 10 mm; Coated Particles d = 10 mm.

Effect of inclined jets
Puffed Wheat and Coated Particles

Minimum Mobilising Velocity (m/s)

Bed Depth (m)

CP horizontal   CP inclined   PW horizontal   PW inclined
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Figure 3.23 Experimental Results: horizontal and inclined jets, no flow through the distributor Puffed Wheat, Puffed Wheat \( d = 10 \text{ mm} \); Coated Particles \( d = 10 \text{ mm} \).

Values of \( Q_{\text{min}} \) for inclined jets divided by: \( \sqrt{2} \)

Effect of inclined jets / \( \sqrt{2} \)

Puffed Wheat and Coated Particles

Minimum Mobilising Velocity (m/s)

Bed Depth (m)

\( \square \) CP horizontal

\( \uparrow \) CP inclined

\( \diamond \) PW horizontal

\( \triangle \) PW inclined
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Figure 3.24 Free Turbulent Jet
Figure 3.25 Correlations for Horizontal Jet Penetration Length

Jet Correlations
Jet Length v Velocity at Orifice

Jet penetration length (m) vs. jet length at orifice (m/s)

- Merry
- Zenz
- Shakhova

Experimental orifice to orifice length
Figure 3.26  Extended Inter-jet Path Length with Inclined Jets
Figure 3.27 225 mm bed solution of Torque Balance (K = K_p)

Torque Balance Puffed Wheat (passive)
Torque v Air Flow (all via jets)
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Figure 3.28 225 mm bed solution of Torque Balance (K = K_A)
Figure 3.29 225 mm bed Torque balance solution curves

Solution of Torque Balance Equations

225 mm bed: Puffed Wheat

$q_{\min} (m^3/s)$ vs $H (m)$

- active
- passive
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4.1 Introduction

It is important to understand something of the heat transfer properties of the novel bed design described earlier, as many of the possible applications will require the addition or removal of heat from the mobilised bed; for example, in a fermenter the metabolic heat has to be removed, and in a drying application heat for evaporation has to be supplied to the particles via the gas or through the walls of the bed. In any application requiring heat addition or removal, there will be a heat source and a heat sink, which may include the particles, the gas, the walls of the bed or any immersed surfaces; in particular, the particles may be the source or sink, or merely a carrier medium between the gas and the surfaces in the bed. This chapter identifies both the dominant mechanisms of heat transfer and the magnitude of the coefficients of heat transfer in the novel bed design, by a study of the available literature and by a discussion of experimental results.

The most relevant area of work on heat transfer is that for gas-fluidised beds, which has been reviewed by Botterill (1975, 1986), and Xavier and Davidson (1985). A fluidised bed is notable for its high rates of heat transfer, both between the particles and the gas, and with submerged surfaces; this normally leads to very homogeneous temperature conditions throughout the bed. The work by LaNauze and Jung (1982, 1983) on combustion in fluidised beds provides some data on the behaviour of larger particles in a fluidised bed.

In the work described in this thesis the time-dependent aspects of the heat transfer mechanisms, which are a consequence of the regular circulation in the spinning bed, can be related to the heat transfer mechanisms in slugging beds as described by Xavier and Davidson (1985).

4.2 Heat Transfer in Fluidised Beds

Within a fluidised bed, heat transfer occurs by a combination of
different mechanisms. Between the gas and the particles, heat is exchanged by conduction and convection, and as the particles have a high surface area per unit volume of the bed and are intimately mixed with the gas, the gas and particles rapidly approach an equilibrium temperature. Heat is also exchanged between particles by conduction through both the contact points and the gas boundary layer surrounding each contact point.

Heat transfer at a submerged surface also occurs by a combination of particle and gas mediated mechanisms, see Fig. 4.1. Particle-convective surface-to-bed heat transfer was first described by Mickley and Fairbanks (1955). "Packets" of particles at the bulk bed temperature arrive at the surface and exchange heat with it. After a while, the mixing processes in the bed sweep them away and disperse the particles into the bed, where they exchange heat with the gas and other particles as described above. The high rates of heat transfer are due to the particle properties: as the solids heat capacity per unit volume is at least an order of magnitude greater than that of the gas, the movement of particles carries relatively higher quantities of heat around the bed. The process of gas-convective heat transfer with a surface is similar to that seen in a packed bed, with the gas exchanging heat at the surface and then percolating into the bulk of the bed. At elevated temperatures, heat transfer also occurs by radiation, and as a first approximation, the gas-convective, particle-convective and radiative heat transfer processes can be considered to be additive (Baskakov et al. 1973). Heat removed from the surface by the particles is also transferred to the gas throughout the bed by particle-to-gas heat transfer.

Theoretical or empirical models for these various mechanisms of heat transfer in fluidised beds have been developed, and these are reviewed by Botterill (1986) and Xavier and Davidson (1985). To simplify the analysis of a system with several parallel heat transfer mechanisms, it is often possible to identify a single
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dominant mechanism and hence, by measuring the overall heat transfer coefficient, to obtain an approximate value for the specific coefficient. For example, radiative heat transfer is usually taken to be negligible at bed temperatures below 900 K. The dominant mechanism can be identified in some cases by comparison of the published correlations for the various heat transfer coefficients.

For an application with more than one process in series, a similar comparison of the published correlations can sometimes be used to identify the rate-controlling step. To do this it is necessary to identify the route by which the heat is transferred. For example, in the transfer of heat between the gas and the wall, transfer between the gas and the particles is significant if particle convective heat transfer between the bed and the wall is the dominant mechanism. However, it is much less important if the dominant mechanism at the wall is gas convection. By calculating the relative magnitude of the three coefficients: gas-to-particle, gas-to-wall and particle-to-wall, the dominant mechanism between the bed and the wall can be predicted. In the case where heat is transferred via the particles to the wall, the rate-limiting step, either gas-to-particle or particle-to-wall, can be determined.

The general trends observed in fluidised beds can be applied to a wide range of operating conditions. However, in applying empirically-derived correlations for any of the heat transfer coefficients, it is necessary to consider the range of conditions over which the correlations apply. In many cases the operating conditions are outside the range of the correlations and it is necessary to use experiments to determine the absolute value of the heat transfer coefficients.

In the following sections the available correlations are applied to the novel bed described in earlier chapters; in an attempt to
determine the dominant heat transfer mechanisms and the magnitude of the heat transfer coefficients.

### 4.2.1 Heat Transfer to Vertical Surfaces: General Observations

For any system in which heat is added or removed through the wall, or via vertical surfaces immersed in the bed, the coefficient of heat transfer between the bed and the wall needs to be known. This is a case where several heat transfer mechanisms operate in parallel and the relative importance of each mechanism depends on the conditions in the bed. Botterill (1986) states that particle convection is significant for particles in the size range 4 - 1000 µm, and gas convection is significant for particles greater than 800 µm in diameter, while radiative heat transfer becomes significant at temperatures above 900 K. Baskakov and Suprun (1972) found experimentally that the fraction of the heat transferred from the surface by the gas increased from 10 to 95% as the particle diameter was increased from 140 µm to 4 mm. Therefore for the system considered here it appears that gas convection may be the most significant process (as the particles are at the top of Baskakov's range), while radiant heat transfer will be negligible at fermenter operating temperatures, which are below 373 K. However, it should be noted that the particles used here are much less dense than the more conventional materials to which Botterill and Baskakov refer; therefore, for these particles, the size ranges over which the various regimes are valid will be different. This is discussed further below.

The variation in the value of the overall bed to surface heat transfer coefficient as the particle size is increased has been investigated by Baskakov et al. (1973), who measured the maximum overall heat transfer coefficient between a bed of corundum particles and an immersed cylindrical object, (see Fig. 4.2).

Figure 4.2 shows that the coefficient increases to a maximum as the particle type changes from group C to group A; this
corresponds to the change from a cohesive, channelling bed to a bed of freely moving particles. The minimum occurs at the point at which gas-convection replaces particle-convection as the dominant heat transfer mechanism, as discussed by Botterill, (see above). As the particle size is increased the minimum fluidisation velocity also increases, and the increase in gas flow through the bed leads in turn to the increase in the gas-convective component. At the same time the specific surface area per unit volume of bed and the number of contact points per unit area of wall both decrease, reducing the particle-convective component. In addition, when the particles are too large to be contained within the thermal boundary layer, the part of the particle surface outside the boundary layer remains at the bulk bed temperature and does not take part in heat transfer. Thus the particle-convective component of heat transfer decreases with increasing particle size.

This published trend highlights the difference between the particles used in this work and typical particles in a fluidised bed application. The characteristics of a fluidised bed are strongly influenced by the particle type as defined by Geldart, (see section 1.2). Therefore it is often more appropriate to compare particles on the basis of the Geldart classification rather than according to size. Hence, although the particles used in the experiments are relatively large, we must take account of the fact that they fall into group B under Geldart's classification, due to their low density.

Referring to figure 4.2, the predicted value of the overall bed to surface coefficient is in the range 200-250 W m\(^{-2}\)K\(^{-1}\) for particles of the size of those in this system: equivalent volume sphere diameter \(d_v\) 8.6 mm, whereas type B particles show values between 300 and 800 W m\(^{-2}\)K\(^{-1}\). It is therefore difficult to predict the values to be expected in this system, because of the unusually low particle density. The lower density will affect the
value of both the particle convective and gas convective components of heat transfer. High gas convective coefficients are related to the higher fluidisation velocities associated with heavier particles. Therefore the coefficient will be reduced for lighter particles. The particle-convective component depends on the particle specific heat and density, and is also reduced for lighter particles. The trends observed will still apply: the particle convective component will decrease as the particle size increases, while the gas-convective component will increase, though this trend may not be as marked as it is for denser particles, such as sand, grain, or plastic beads. The overall effect of using particles which are both large and of low-density is to make any prediction of the absolute values of the heat transfer coefficients likely to be an over-estimate. This is true whether we base the estimate on the Geldart classification, or on the particle size.

The consistent result from all the published work is that, with a particle above 1 mm in diameter, gas-convection will be the largest component of heat transfer between the bed and the wall. However, although we can neglect radiative transfer, we must still consider the absolute value of the particle-convective heat transfer coefficient \( h_{pc} \), although we expect it to be small compared to the gas convective coefficient \( h_{gc} \).

In a complementary study, the maximum heat transfer coefficient as a function of air flow rate has been studied by Todes (1965). Todes showed that as the air velocity is increased above \( U_{mf} \), the heat transfer coefficient increases at first, due to the increased mixing in the bed, but then decreases, as the excess air, passing through the bed as bubbles, starts to shroud the heat transfer surfaces. The maximum value of the heat transfer coefficient occurs closer to \( U_{mf} \) as the particle size is increased, corresponding to a move from group A through B to D.
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The published literature is consistent when discussing the trend in the relative importance of the different transfer mechanisms as the particle size is varied. However, there is relatively little experimental work which has involved particles greater than 1 mm in diameter, and none published on large light particles such as are considered in this work. Therefore the correlations which are discussed below, which may be the best available, have been used to estimate the relative importance of the mechanisms, rather than the absolute value of the coefficients. The results have then been used to design a suitable experimental programme to measure the rates of heat transfer in the bed.

4.2.2 Correlations for the Gas Convective Heat Transfer Coefficient to Vertical Surfaces

Correlations to evaluate $h_{gc}$ have been developed by a number of workers. Baskakov and Suprun (1972) measured the mass transfer coefficient for the evaporation of naphthalene into a fluidised bed and used these results to develop a correlation for the mass transfer coefficient.

$$Sh = 0.0175 \text{ Ar}^{0.46} \text{ Sc}^{0.33} \quad U > \text{ U}_m \quad (4.1)$$

$$Sh = 0.0175 \text{ Ar}^{0.46} \text{ Sc}^{0.33} \left( \frac{U}{U_m} \right)^{0.3} \quad \text{U}_m < U < \text{ U}_m \quad (4.2)$$

For systems in which the Reynolds number is high, the relationship between the Sherwood number and the Archimedes and Schmidt numbers can be used to derive the relationship between the Nusselt number and the Archimedes and Prandtl numbers by an accepted analogy. Hence Baskakov and Suprun were able to derive the following correlation.

$$Nu_{gc} = 0.0175 \text{ Ar}^{0.46} \text{ Pr}^{0.33} \quad U > \text{ U}_m \quad (4.3)$$

$$Nu_{gc} = 0.0175 \text{ Ar}^{0.46} \text{ Pr}^{0.33} \left( \frac{U}{U_m} \right)^{0.3} \quad \text{U}_m < U < \text{ U}_m \quad (4.4)$$
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where \( U_m \) is the superficial gas velocity giving maximum bed to surface heat transfer (from Todes 1965).

The Archimedes number \( Ar \) is given by:

\[
Ar = \frac{\rho_f d_v^3 \left( \rho_p - \rho_f \right) g}{\mu_f^2}
\]  

(4.5)

Botterill and Denloye (1978) worked with a bed at the point of incipient fluidisation, having proposed that the overall heat transfer coefficient in this state is equal to \( h_{gc} \). This is based on the fact that at incipient mobilisation the particle renewal rate at the wall is very low, i.e. the particles are not mixed back into the bed, so heat transfer by particle convection is negligible. The value of \( h_{gc} \) obtained is subject to some inaccuracy, as there is also some conduction through the stationary layer of gas close to each inter-particle contact point, and through the particles themselves. However the error only becomes significant when the particle thermal time constant is greater than its residence time at the surface, i.e. for small particles, typically less than 1 mm in diameter. The effect of lower density will be to reduce the thermal time constant. Therefore in the system considered here this error should be negligible.

The correlation derived by Botterill and Denloye (1978) is:

\[
\frac{h_{gc} d^{0.5}}{k_f} = 0.86 Ar^{0.39} \quad \text{for} \quad 10^3 < Ar < 2 \times 10^6
\]  

(4.6)  

(dimensions are m\(^{-0.5}\))

As the temperature is increased, the Archimedes number decreases due to changes in the fluid viscosity. Using the correlations described above for the gas-convective heat transfer coefficient and the parameter values set out below, gives the following values for our system:
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Table 4.1:
Parameters for Calculation of Heat Transfer Coefficients

<p>| | | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Puffed Wheat</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>equivalent volume sphere diameter $d_v = 8.6$ mm</td>
<td></td>
<td></td>
</tr>
<tr>
<td>particle density $\rho_p = 100$ kg m$^{-3}$</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Air at Ambient Conditions</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>fluid density $\rho_f = 1.29$ kg m$^{-3}$</td>
<td></td>
<td></td>
</tr>
<tr>
<td>fluid viscosity $\mu_f = 1.85 \times 10^{-5}$ N s m$^{-2}$</td>
<td></td>
<td></td>
</tr>
<tr>
<td>fluid thermal conductivity $k_f = 0.026$ W m$^{-1}$K$^{-1}$</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$g = 9.81$ m s$^{-2}$</td>
<td></td>
<td></td>
</tr>
<tr>
<td>particle specific heat capacity $C_p = 1047$ W kg$^{-1}$K$^{-1}$</td>
<td></td>
<td></td>
</tr>
<tr>
<td>particle thermal conductivity $k_p = 2.6$ W m$^{-1}$K$^{-1}$</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

$$Ar = \frac{1.29 (0.0086)^3 (100 - 1.29) \times 9.81}{(1.85 \times 10^{-5})^2} = 2.32 \times 10^6$$

Correlation of Todes:

$$Re_{opt} = \frac{Ar}{(18 + 5.22 \sqrt{Ar})}$$

$$Re_{mf} = \frac{Ar}{(1400 + 5.22 \sqrt{Ar})}$$

$$Re_{mf} = 248 \quad U_{mf} = 0.414 \quad m s^{-1}$$
$$Re_{opt} = 291 \quad U_{mf} = 0.486 \quad m s^{-1}$$

Values for the thermal properties of Puffed Wheat were not available; therefore the values for cellulose have been used, where required, throughout this work. These were taken from the Handbook of Chemistry and Physics (1977).
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Baskakov:

\[ \text{Nu}_{gc} = 0.0175 \text{Ar}^{0.46} \text{Pr}^{0.33} = 16.34 \quad U > U_m \]

\[ \text{Pr} = \frac{k_f}{C_p \mu_f} = 1.34 \]

\[ \text{Nu} = \frac{h_{gc} d_v}{k_f} \]

\[ h_{gc} = \frac{16.34 \times 0.026}{0.0086} = 49.4 \text{ W m}^{-2} \text{K}^{-1} \]

Botterill:

\[ \frac{h_{gc} d_v^{0.5}}{k_f} = 0.86 \text{Ar}^{0.39} = 261.3 \text{ m}^{-0.5} \]

\[ h_{gc} = 73.26 \text{ W m}^{-2} \text{K}^{-1} \]

These estimates of \( h_{gc} \) are significantly lower than typical measured values for fluidised beds of more conventional type B particles, (300-800 W m\(^{-2}\)K\(^{-1}\)), assuming gas convection to be the main mechanism of heat transfer. They are also lower than the values measured for larger particles, (150-250 W m\(^{-2}\)K\(^{-1}\)). However they are consistent with the expectation that the heat transfer coefficients will be lower with large, light particles than for conventional materials.

4.2.3 Correlations for Particle Convective Heat Transfer to Vertical Surfaces

The other possible mechanism of heat transfer between vertical surfaces and the bulk of the bed is particle convection. As discussed above, (section 4.2.1), there is a lack of published data on low density materials. However, it is possible to use the published correlations to obtain a value which can be compared to the values obtained for gas convection.
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In the earliest work on particle convective heat transfer, Mickley and Fairbanks (1955) analysed the case where aggregates of particles exchange heat with an isothermal surface while in contact with it. They assumed that the particles are at the bulk temperature when they arrive in the gas boundary layer, which is at the wall temperature. Using the analysis of the decay of a temperature gradient with time, they obtained the following expression for the instantaneous heat transfer coefficient:

\[
h_i = (k_{mf} \rho_{mf} C_{mf} / \pi t)^{1/2}
\]

where \( k_{mf}, \rho_{mf}, \text{ and } C_{mf} \) are the effective conductivity, density and heat capacity respectively of the particulate phase, and \( t \) is the time elapsed since the aggregate arrived at the wall. An average value for the coefficient can be obtained by integration if the distribution of the package residence times is known.

Such a theoretical analysis is presented by Mathur and Epstein (1974) for the heat transfer coefficient between the particles and the wall in a spouted bed, using the vertical velocity at the wall to predict the residence time \( t_r \), with the assumption that the particles remain at the wall until they reach the base of the bed and hence the residence time is constant. This is of interest because a spouted bed is typically used to process larger particles such as grain which are more comparable to the particles tested in the prototype bed.

The integrated expression for a spouted bed is:

\[
h_{pc} = 2 \left[ \frac{\rho_b C_p k_b}{\pi t_r} \right]^{1/2}
\]

where \( t_r \) is the residence time at the wall, and \( \rho_b, C_p \text{ and } k_b \) (density, heat capacity and thermal conductivity) are the bulk properties of the bed.
In a later analysis, for slug flow in a fluidised bed, Xavier and Davidson (1985) used an expression for the residence time based on the slug frequency, which is a function of the slug velocity and gas flow rate. The assumption is that the particles remain in contact with the wall until the passing of a slug mixes them back into the bed. The average heat transfer coefficient between the particles and the surface can then be calculated as above:

\[
\frac{h_p}{\dot{t}} = \left[ \frac{k_{mf} \rho_{mf} C_{mf}}{\pi t_r} \right]^{1/2}
\]  

The bulk properties of the bed, between slugs, are assumed to be those of a fluidised bed at the minimum fluidisation velocity \(U_{mf}\).

To evaluate this expression Xavier and Davidson use the best available correlations from the literature, i.e.

\[
k_{mf} = k_e^0 + 0.1 \rho_f C_d U_p \quad \text{(4.12)}
\]

\(k_e^0\) is evaluated from:

\[
\frac{k_e^0}{k_f} = \left( \frac{k_p}{k_f} \right) [0.28 - 0.747 \log_{10} C - 0.057 \log_{10} (kp/kf)]
\]

\(\text{(attributed to Krupicza (1967))}\)

The minimum fluidisation velocity is derived from the Ergun equation, (equation 3.26), and the bed properties are derived from the particles properties, i.e.:

\[
\rho_{mf}=\rho_p(1-e_{mf})C_{p}
\]

\(\text{(4.14)}\)
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Baskakov (see Xavier and Davidson 1985), added a second, time-independent, resistance term to account for the zone of variable properties adjacent to the wall; for instance, the voidage increases near the wall. This film coefficient can be expressed as:

\[ h_f = \frac{m_k f}{d_p} \]  \hspace{1cm} (4.15)

where the constant \( m \) is a function of the particle packing at the wall, which is \( 2\pi \) for cubic packing and can be taken as 6 for design purposes (Xavier and Davidson 1985).

Taking the two resistances in series, the particle convective heat transfer is given by:

\[ \frac{1}{h_{pc}} = \frac{1}{1/h_p} + \frac{1}{1/h_f} \] \hspace{1cm} (4.16)

This theory can be applied to any system where the particle residence time can be measured or predicted.

For the prototype design under consideration here, we can propose the hypothesis that the residence time at the wall is controlled by the action of the jets. That is to say, the particles rotate in a regular manner and remain in contact with the wall between the jets. When they reach the next jet, the re-entrainment of part of the air flow into the next jet, and the orientation of the jet, at \( 45^\circ \) to the wall, disrupt the flow of air along the wall and lift the particles off the wall, mixing them back into the bed.

Therefore as a first approximation we can calculate the residence time from the average velocity of the particles at the wall and the distance between the jets.

The particle velocity in the 145 mm diameter bed was measured
using a video film, during a series of experiments designed to measure the degree of mixing in the bed. The horizontal velocity is shown as a function of gas flow rate through the jets in figure 4.3. From the graph, a representative velocity for this bed can be taken to be 0.1 m s\(^{-1}\), and as the distance between any two of the four equi-spaced jets is:

\[
\frac{0.145 \, \text{m}}{4} = 0.11 \, \text{m} \tag{4.17}
\]
giving a residence time at the wall of the order of 1.1 s.

Table 4.2 Parameters For Determining \( h_{pc} \)

<table>
<thead>
<tr>
<th>Fluid and Particle Properties as used above</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Fluid properties</td>
<td></td>
</tr>
<tr>
<td>( C_f = 1006 ) J kg(^{-1})K(^{-1})</td>
<td></td>
</tr>
<tr>
<td>( k_f = 0.026 ) W m(^{-1})K(^{-1})</td>
<td></td>
</tr>
<tr>
<td>( \rho_f = 1.29 ) kg m(^{-3})</td>
<td></td>
</tr>
<tr>
<td>( C_p = 1047 ) J kg(^{-1})K(^{-1})</td>
<td></td>
</tr>
<tr>
<td>( k_p = 2.6 ) W m(^{-1})K(^{-1})</td>
<td></td>
</tr>
<tr>
<td>( \rho_p = 100 ) kg m(^{-3})</td>
<td></td>
</tr>
<tr>
<td>( d_v = 0.0086 ) m</td>
<td></td>
</tr>
<tr>
<td>bed voidage ( \varepsilon_{mf} = 0.4 )</td>
<td></td>
</tr>
<tr>
<td>( t = 1.1 ) s</td>
<td></td>
</tr>
<tr>
<td>( m = 6 )</td>
<td></td>
</tr>
<tr>
<td>( U_{mf} = 0.41 ) m s(^{-1})</td>
<td></td>
</tr>
</tbody>
</table>

The assumption is also made that the net flow of gas within a particle packet (as defined by Mickley and Fairbanks (1955)), is equal to \( U_{mf} \), despite the fact that the particles are travelling at an angle to the direction of the air flow.

Results are as follows:

\[
\begin{align*}
  k_e^0 &= 0.220 \, \text{W m}^{-1}\text{K}^{-1} \quad \text{equation 4.13} \\
  k_{mf} &= 0.681 \, \text{W m}^{-1}\text{K}^{-1} \quad \text{equation 4.12} \\
  h_p &= 222 \, \text{W m}^{-2}\text{K}^{-1} \quad \text{equation 4.11} \\
  h_f &= 18.14 \, \text{W m}^{-2}\text{K}^{-1} \quad \text{equation 4.15}
\end{align*}
\]
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hence:

\[ h_{pc} = 16.77 \text{ W m}^{-2}\text{K}^{-1} \]  

This value of the particle-convective heat-transfer coefficient is lower than the values for gas-convection, see section 4.2.2. The greatest influence on the value of \( h_{pc} \) is the film coefficient \( h_f \) (18.14 W m\(^{-2}\)K\(^{-1}\)), due to the large particle size; the reduction in the coefficient with increase in particle size predicted by this theory is consistent with the trend observed by Baskakov. The assumption made about the gas flow rate through the particle packet is irrelevant, as this affects the value of \( k_m \) which has only a minor influence on \( h_{pc} \).

4.2.4 Gas to Particle Heat Transfer

For those applications where the particles are the source of heat, such as fermentation, or the sink, as in drying processes, heat transfer between the particles and the gas is important, either because the gas carries the bulk of the heat into or out of the bed, or because the gas transfers heat between the particles and the wall.

The coefficient of heat transfer between the gas and the particles can also be predicted using published correlations and will determine the condition of the particles in the bed. In most applications gas-to-particle heat transfer is not the rate-determining step in heat transfer between the wall and the particles; the large particle surface area ensures that equilibrium between the particles and the gas is rapidly reached. Therefore the particles are usually at the same temperature as the gas leaving the bed.

Experiments on fluidised beds have shown that the measured gradient in the gas temperature at the inlet to the bed is too steep to allow differentiation between plug flow and well mixed conditions for the gas in the bed (Botterill, 1986). However a
plug flow model has been shown to give more consistent results when applied to the work of different investigators.

As was noted above, there are few published correlations for larger particles. The two examples below indicate the inadequacy of the available correlations when applied to particles larger than 1 mm in diameter.

LaNauze (1982) derived a correlation for $h_p$ from work done on mass transfer in fluidised bed combustors. From experiment he developed the correlation:

$$\text{Sh} = 2c + 0.69 \left( \frac{1}{\varepsilon} \right)^{1/2} \text{Re}^{1/2} \text{Sc}^{1/3} \quad (4.18)$$

where the Sherwood number is give by:  

$$\text{Sh} = \frac{R_m d_p}{D}$$

and the Schmidt number $\text{Sc}$ is given by:  

$$\text{Sc} = \frac{\mu_f}{\rho_f D}$$

($R_m$ is the mass transfer coefficient and $D$ is the diffusivity of oxygen in air.)

By using the mass-transfer/heat-transfer analogy mentioned above, the equivalent heat transfer correlation was derived:

$$\text{Nu} = 2c + 0.69 \left( \frac{1}{\varepsilon} \right)^{1/2} \text{Re}^{1/2} \text{Pr}^{1/3} \quad (4.19)$$

where the Prandtl number is given by:  

$$\text{Pr} = \frac{k_f}{C_p \mu_f}$$

Using the data for our system:

- $\varepsilon = 0.4$
- $\rho_f = 1.29$
- $U = U_{mf} = 0.41 \text{ m s}^{-1}$
- $\text{Pr} = 1.34$
- $\text{Re} = 248$
hence:

\[
\begin{align*}
\text{Nu} & = 19.8 \\
h_{\text{gp}} & = 59.7 \text{ W m}^{-2}\text{K}^{-1}
\end{align*}
\]

This is higher than reported values, (e.g. 6 - 23 W m\(^{-2}\)K, Geldart, 1986); it indicates the effect of the higher Reynolds numbers which occur due to the higher minimum fluidisation velocities necessary for larger particles.

Zabrodsky (1966) gives the following correlation for the maximum heat transfer coefficient between the particles and the gas:

\[
h_{\text{max}} = 35.8 \rho_p^{0.2} k_f^{0.6} d_p^{-0.36}
\]

This is recommended for particles up to 800 µm in diameter, hence:

\[
h_{\text{max}} = 35.8 (100)^{0.2} (0.026)^{0.6} (0.0086)^{-0.36} = 53.2 \text{ W m}^{-2}\text{K}^{-1}
\]

This is in good agreement with the value obtained above but it is also relatively high.

4.2.5 Conclusions

The results of the various correlations are summarised in Table 4.3.

Table 4.3 Heat Transfer Coefficients Derived from Published Correlations

<table>
<thead>
<tr>
<th>Coefficient</th>
<th>W m(^{-2})K(^{-1})</th>
<th>Reference</th>
</tr>
</thead>
<tbody>
<tr>
<td>h_{gc}</td>
<td>49.4</td>
<td>Baskakov and Suprun (1972)</td>
</tr>
<tr>
<td>h_{pc}</td>
<td>73.3</td>
<td>Botterill and Denloye (1978)</td>
</tr>
<tr>
<td>h_{gp}</td>
<td>16.8</td>
<td>Xavier and Davidson (1985)</td>
</tr>
<tr>
<td>h_{gp}</td>
<td>59.7</td>
<td>LaNauze (1983)</td>
</tr>
<tr>
<td>h_{gp}</td>
<td>53.2</td>
<td>Zabrodsky (1966)</td>
</tr>
</tbody>
</table>

As was discussed in the introduction, the correlations in the
literature provide limited information about the heat-transfer behaviour of fluidised beds of large light particles. The absolute values calculated for the heat transfer coefficients vary both from each other and from the typical ranges of values measured experimentally. However the relative magnitude of the coefficients does agree with the observed trends and this provides a basis for predicting the behaviour of this system.

The trends observed in the work with particles of increasing size indicate that gas-convection is likely to dominate the transfer of heat between the bed and the wall, with a small or negligible contribution from particle convection. Radiative heat transfer will be negligible at low temperatures. Within the bed the rate of heat transfer is likely to be high, due to the relatively high minimum fluidisation velocity. Therefore the bed is expected to be at the same temperature as the gas leaving the bed.

4.3 Experimental Work
4.3.1 Design of the Experiments
For the fermentation application discussed above, the critical parameter is the rate at which heat generated in the particles can be removed through the walls after the air has been saturated. Similarly, in a drying application the rate at which heat can be supplied through the wall to the particles is of interest.

As the heat transfer into the bed from the wall is expected to be dominated by gas convection, these experiments were designed to measure the overall heat transfer coefficient between the gas and the wall. To investigate the effect of increasing the air flow rate, measurements were made at three different flow rates. Hence, by using the approximation that the overall rate of heat transfer is due to gas-convection alone, a measurement of the gas-convective component of heat transfer can be made.

Although the above approximation can be made with some confidence,
the results from the previous section indicate that particle convection may contribute as much as twenty percent of the total heat transfer. The relative magnitude of the particle convective heat transfer coefficient is heavily dependent on the particle renewal rate at the wall of the bed. This can be increased by increasing the flow of air through the bed, thus promoting more mixing; however, increasing the air flow also increases the gas convective component, so the two effects cannot be separated.

With the novel bed design the effect of the jets on the heat transfer coefficients is of particular interest. To investigate this, the total air flow in the bed was kept constant, but the division of the air between the jets and the distributor was altered. The jets have a marked effect on the particle velocity at the wall, and will also increase the gas velocity relative to the wall in the jet area. The effect of this was measured at the three experimental flow rates. This allowed the variation in the overall heat transfer coefficient to be measured as the particle velocity at the wall is varied, without altering the superficial gas velocity in the bed.

These experiments also provide a test for the hypothesis discussed in section 4.2.3, that the particle residence time is a function of the particle velocity at the wall and the jet spacing. Hence, if we accept that the particle velocity is a function of the jet velocity as indicated in Fig. 4.3., then the particle residence time is a function of the jet velocity, and we can predict that the overall heat transfer coefficient will increase as the flow through the jets is increased, even when the total air flow rate supplied to the bed is kept constant. The magnitude of the increase will give some indication of the relative importance of particle convective heat transfer to the wall.

The consistency of the conditions in the bed, particularly the development of temperature gradients, is also of interest, as for
fermentation applications small variations in temperature can have a large effect on the productivity and viability of the culture.

4.3.2 Experimental Equipment and Method

For this work a second experimental rig was designed and built (see Fig. 4.4). This was slightly larger than the 150 mm bed used for the work on hydrodynamics, being 225 mm in diameter. The lower part of the bed was made from stainless steel with four 10 mm inlets set at 45° to the tangential, and this part was double-walled to form a water jacket 225 mm deep. The upper part of the bed was made from glass, also 225 mm deep, with a stainless steel lid with ports for thermocouples and other probes. The air supply was split just before the bed into two separately controlled streams, to the jets and to the manifold supplying the distributor plate; orifice plates were used to measure the total flow and the flow diverted to the jets. The distributor was also made from stainless steel with 2 mm holes set on a 14 mm triangular pitch. The air leaving the bed was taken through a cyclone, which collected most of the fines, before being vented to atmosphere.

Thermocouples were fitted to measure the temperature of the air entering and leaving the bed, and the temperature of the water entering and leaving the jacket. Thermocouples were also fitted to record the temperature of the inner wall of the steel section where it was in contact with the particles; these were at four different heights. The ambient temperature and the temperature of the water bath supplying the water jacket were also monitored. To measure the particle temperature, a "Heidolph" infra-red probe was fitted in the top of the bed to measure the surface temperature of the top layer of particles. The readings from all these sensors were fed into an Amstrad PC1512 computer and stored on floppy disc using a data-logging program written in TURBO Pascal.

For each run the bed was filled with the test material to within...
25 mm of the top of the steel section; when aerated the bed expanded so that the entire steel section was covered. The air flows were initially set so that all the air entered the bed through the distributor, the water supply to the jacket was turned on and the bed was allowed to run for approximately 45 minutes until it reached steady-state conditions. A set of readings was then taken. At the same time that the readings were taken, the fines in the cyclone were collected and weighed to determine the rate of attrition. The air-flows were then adjusted so that part of the flow was diverted through the jets but the overall flow was constant, and the bed was allowed to equilibrate for a further 20 minutes before another set of readings was taken. This procedure was repeated for the range of flows through the jets.

Each set of temperature readings was taken over a period of 100 seconds to eliminate the noise generated by the computer (the design of the Amstrad places the power supply uncomfortably close to the data ports), and to allow for the fluctuations in the measured surface temperature of the particles, which occur as material from within the bed is brought to the surface. Therefore for each reading a set of 25 values at four second intervals was taken and averaged.

4.3.3 Analysis
To calculate the overall heat transfer coefficient between the wall and the gas we need to know the rate at which heat is removed from the bed by the gas, \( Q_{\text{gas}} \), the gas mass flow rate \( M_{\text{gas}} \), the area of wall \( a_{\text{wall}} \) and the mean temperature driving force \( \Delta T_{\text{gw, mean}} \). The gas flow rate is calculated from the orifice plate reading, corrected for pressure, using a calibration curve obtained by calibrating the plates in situ against a rotameter open to the atmosphere. The heat transfer rate is calculated from the gas flow and measured temperature change in the gas stream:

\[
Q_{\text{gas}} = C_f M_{\text{gas}} \Delta T_{\text{gas}} \text{ J s}^{-1} \tag{4.21}
\]
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The temperature driving force is calculated from the four surface thermocouples; of these the lowest thermocouple is mounted on the weld at the base of the bed and records an atypically low value, while the other three are mounted equidistantly apart between the top and the base of the bed and show a small but significant temperature gradient. Typical values measured by each thermocouple are shown in Table 4.4.

Table 4.4 Run 14 Mean Temperatures at the Wall

<table>
<thead>
<tr>
<th>Thermocouple</th>
<th>Position</th>
<th>Mean temperature</th>
</tr>
</thead>
<tbody>
<tr>
<td>6</td>
<td>top</td>
<td>64.8</td>
</tr>
<tr>
<td>7</td>
<td>second bottom</td>
<td>58.1</td>
</tr>
<tr>
<td>8</td>
<td>second top</td>
<td>56.4</td>
</tr>
<tr>
<td>9</td>
<td>bottom</td>
<td>43.2</td>
</tr>
</tbody>
</table>

It was decided to take the mean of the top three thermocouples as the best estimate of the wall temperature. The gas temperature is taken as the mean of the measured inlet and outlet temperatures. The difference between these two means is approximately equal to the log mean temperature difference, as the difference between the inlet and outlet gas temperatures is much less than the difference between the wall temperature and the gas temperature; for example Table 4.5 shows the data for run 141:

Table 4.5 Temperature Data at Steady State, Run 141

<p>| | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>air inlet (T_{g,\text{in}})</td>
<td>28.49 °C</td>
</tr>
<tr>
<td>air outlet (T_{g,\text{out}})</td>
<td>34.38 °C</td>
</tr>
<tr>
<td>mean wall temperature</td>
<td>60.53 °C</td>
</tr>
<tr>
<td>(\Delta T) inlet</td>
<td>32.04 K</td>
</tr>
<tr>
<td>(\Delta T) outlet</td>
<td>26.15 K</td>
</tr>
<tr>
<td>mean (\Delta T)</td>
<td>29.10 K</td>
</tr>
<tr>
<td>log mean (\Delta T)</td>
<td>29.00 K</td>
</tr>
</tbody>
</table>

Hence:
Chapter 4: Heat Transfer

\[
\Delta T_{gw,\text{mean}} = \left( \frac{T_2 + T_3 + T_4}{3} - \frac{T_{q,\text{in}} + T_{q,\text{out}}}{2} \right) \quad (4.22)
\]

and the heat transfer coefficient is obtained from:

\[
h_{gw,\text{(overall)}} = \frac{Q_{\text{gas}}}{\Delta T_{gw,\text{mean}} \ a_{\text{wall}}} \quad (4.23)
\]

The accuracy of the data logging system is discussed in appendix B.

4.3.4 Closure of Heat Balance

To assess the amount of heat lost from the system during the heat transfer experiments, some tests were performed on the bed before the particles were loaded. The water jacket was not lagged and therefore, despite the polished surface the greatest loss of heat was expected to be by convection from the metal surface of the lower section. The results of the experiments are shown in Table 4.6:

<table>
<thead>
<tr>
<th>run</th>
<th>flows</th>
<th>temperatures</th>
<th>heat flows</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>air</td>
<td>water in C</td>
<td>water out C</td>
</tr>
<tr>
<td>1</td>
<td>0</td>
<td>2.2</td>
<td>62.51</td>
</tr>
<tr>
<td>2</td>
<td>535</td>
<td>2.2</td>
<td>63.14</td>
</tr>
<tr>
<td>3</td>
<td>1415</td>
<td>2.2</td>
<td>63.42</td>
</tr>
<tr>
<td>4</td>
<td>2774</td>
<td>2.2</td>
<td>63.35</td>
</tr>
</tbody>
</table>

The variation in the gas inlet temperature is due to a rise in the ambient temperature in the laboratory over the course of the day.

The four runs were performed under different conditions to compare the heat removed from the wall with the heat carried by the air leaving the bed. As can be seen from Table 4.6, a trend is observed, and the discrepancy between the observed heat flows
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decreases as the flow-rate is increased. There are a number of factors which will affect the results. Firstly, in the first experiment there was no air flow; therefore losses by convection occurred over both sides of the water jacket. Secondly, the flow past the thermocouple may not be fully developed at the lower flow rates; for example, 535 l/min is equivalent to 2 m/s in the 76 mm outlet pipe, equivalent to a Reynolds' number of 2500 for a 1 mm thermocouple junction. Unless the heat transfer coefficient between the thermocouple and the surroundings is high, the thermocouple will register something between the gas temperature and the temperature of the surroundings. Thirdly, the bed may not have had long enough to reach steady state before the first readings were taken, hence a proportion of the heat carried by the gas may have been absorbed by the upper part of the bed, i.e. by the glass section and the stainless steel top-plate, which are not heated. This effect would decrease for later runs as the bed approached the gas temperature, and hence would give the observed fall in the losses.

From the results presented above it follows that the later results in this series of experiments are likely to be most accurate; i.e. the losses from the bed are of the order of 55 W at these operating temperatures. Inaccuracies in the results due to the thermal equilibration of the rig will be minimised by allowing the rig to come to equilibrium before starting the experiments, and by performing each set of experiments sequentially, without allowing the equipment to cool down.

The results of the overall heat balance in the last run give an acceptable level of accuracy when measuring the heat transfer in the bed. The significant parameter for the heat transfer experiments is the heat transfer from the wall into the gas. Losses from the outside of the jacket can be estimated from the difference between the heat supplied by the water and the heat removed in the gas; these losses can be ignored in future
calculations if we use the heat flow in the gas stream as the basis for the heat balance, as the losses appear as the difference between these two heat flows.

4.3.5 Experimental Conditions

Data were collected for three runs, using a constant total gas flow rate for each run and varying the proportion of the gas diverted through the jets. The conditions chosen included a flow rate close to the minimum fluidisation velocity and two higher flow rates. The initial experiment in each set was performed with no air diverted to the jets, so the bed was fluidised as far as possible in the conventional manner.

The design of the experiments, based on steady state measurement of the rate of heat transfer to the gas, depends on good mixing and homogeneous conditions throughout the bed; therefore all the experiments were performed at flow rates above the minimum fluidisation velocity and also above the theoretical mobilisation velocity, ( see below ). No experiments were performed at lower flow rates, so that the results give no indication of heat transfer behaviour around the mobilisation point. Rather, the effect of the jets on heat transfer in a fluidised bed has been investigated and, in the context of these experiments, "mobilisation" is the point at which steady circulation is established in the fluidised bed.

The point of mobilisation for this configuration has not been determined by experiment as the metal water jacket prevents any direct observation of the flow patterns at the wall of the bed, though some indication of the bed behaviour is given by movement at the surface. However we can calculate the expected mobilisation point velocity for the conditions in this rig using the theory developed in Chapter 3, ( see also section 3.10 ). This approach gives a jet flow of $0.0234-0.0262 \text{ m}^3\text{s}^{-1}$, equivalent to a bed superficial velocity of $0.59-0.66 \text{ m s}^{-1}$, to initiate
rotation with no flow through the distributor.

For each run the variation in the steady-state overall heat transfer coefficient (between the wall and the gas) can be plotted as a function of the flow rate through the jets. From the collected data we can also plot the attrition rate, the temperature at fixed points on the wall, and the particle surface temperature as functions of the flow rate through the jets.

4.4 Results

The results are tabulated in appendix C and the measured heat transfer coefficients are plotted in figure 4.5. The results for the three different total flow rates are plotted together for comparison in Fig. 4.5a and shown separately in Figs. 4.5b to d. The results reveal a number of different trends, the most notable of which is a local maximum in the heat transfer coefficient. The effect of increasing the air flow diverted to the jets above the transition point is also shown in Fig. 4.5. There are no other abrupt changes in the rate of heat transfer, however there is a fall in the heat transfer coefficient as the flow through the jets is increased, and there is some indication, from run 14 and the last point in run 12, that the heat transfer rate increases again at the highest jet flow rates.

An initial examination of the results indicates that the particles appear to be reaching equilibrium with the gas, or are slightly hotter than the gas leaving the bed. This is consistent with a slight cooling of the gas between the bed and the point of measurement, and the size of the experimental errors, which are discussed in appendix B. The addition of particles to the bed has no significant effect on the rate of heat transfer through the walls of the bed, compared to the rates measured in the heat balance experiments, (see above). This is evidence that gas convection is the dominant heat transfer mechanism.
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For comparison the predicted value of the mobilisation point, calculated from the theory presented in Chapter 3, is marked on each graph, for the specific operating conditions.

4.5 Experimentally Determined Value of $h_{gw}$

4.5.1 No Flow Through the Jets

With no flow diverted to the jets, the bed operates as a conventional fluidised bed; the coefficient is plotted against the total air flow in Fig. 4.6. The heat transfer coefficient ranges between 57 and 88 W m$^{-2}$K$^{-1}$ under these conditions. There is no indication that the maximum value of the coefficient, observed by Todes' (1965), occurs within the range of flow rates considered; instead, the heat transfer coefficient continues to increase as the air flow is increased. This implies that the air flow is not high enough for gas to shroud the heat transfer surfaces.

4.5.2 With Flow Through the Jets

The range of values of the bed-to-surface heat-transfer coefficient, between 50 and 100 W m$^{-2}$K$^{-1}$, can be compared with the values generated from the published correlations, (49 and 73 W m$^{-2}$K$^{-1}$, Table 4.3), on the assumption that it is to be interpreted as the gas-convective coefficient. The overall coefficient is consistent with the bed-to-surface heat transfer coefficient predicted by the correlations, and it is also comparable to the value of 150-250 W m$^{-2}$K$^{-1}$ predicted by Baskakov et al for particles of this size, (section 4.2.1). The heat transfer in this bed is comparable to that in a spouted bed, where similar sized particles are involved. The slightly higher overall coefficient, compared to typical values for spouted beds, may be due to a higher value of the gas-convective heat transfer coefficient, as the experimental flow rates, which are above $U_{mf}$, are significantly higher than calculated values of the minimum spouting velocity.

In summary, the experimental data presented here suggest that
gas-convection is the major mechanism of heat transfer in fluidised beds of large light particles. The published correlations for predicting $h_{gc}$, described in section 4.2, give values which compare well with the values obtained by experiment. Therefore these correlations can be used to predict the value of the overall heat transfer coefficient in this system.

4.5.3 Variation in $h_{gw}$ as the Proportion of Air Through the Jets is Increased

It can be seen from the results that the variation in $h_{gw}$ as the proportion of the gas flow entering through the jets is increased is not linear: the value remains static or decreases by about 10% from the fluid bed value at low gas flow rates, before increasing abruptly to a maximum. The decrease is due to a combination of factors: air is being diverted from the distributor plate, reducing the excess air and bubble flow in the bulk of the bed and hence the degree of mixing, while any effect on the heat transfer coefficient, due to the increased flow through the jets, is not large enough to counteract this.

As observed in the introduction to this section, there will be a localised increase in $h_{gc}$ due to the increased gas velocity in the jet region. However, as the superficial velocity in the bed is constant, the gas velocity at the wall away from the jets will remain the same, i.e. over most of the wall of the bed.

In addition to any effect on the gas-convective component of heat transfer, there will also be an increase in the particle-convective component. The particle convective component is given by equations 4.11 and 4.16, and these show that the heat transfer coefficient increases as the particle residence time at the wall is reduced. A plot of the heat transfer coefficient against the residence time at the wall (from equation 4.16) is given in figure 4.7.
From figure 4.5 we observe that the increase in $h_w$ is in the range 10 to 20 $\text{W m}^{-2}\text{K}^{-1}$. This is directly comparable to the predicted value of the particle-convective heat transfer coefficient from the theory discussed above, and consistent with a residence time at the wall between one and two seconds, see figure 4.7. (16.8 $\text{W m}^{-2}\text{K}^{-1}$).

This appears to be consistent with the changes observed when the bed is mobilised: when the bed starts to spin, the particle residence time at the wall becomes a function of the particle velocity and the inter-jet distance. However, the predicted values of the mobilisation point, which are plotted on figure 4.5, are not coincident with the change in the heat transfer coefficient. Predicted values for the mobilisation point and the jet penetration lengths are shown in Table 4.7.

### Table 4.7 Predicted Values of Mobilisation Point and Jet Penetration Length

<table>
<thead>
<tr>
<th>Mobilisation Point</th>
<th>Flowrates</th>
<th>Jet Velocity</th>
</tr>
</thead>
<tbody>
<tr>
<td>Run</td>
<td>Total Air ($\text{m}^3\text{s}^{-1}$)</td>
<td>Through Jets ($\text{m}^3\text{s}^{-1}$)</td>
</tr>
<tr>
<td>12</td>
<td>0.0296</td>
<td>0.0170</td>
</tr>
<tr>
<td>13</td>
<td>0.0296</td>
<td>0.0169</td>
</tr>
<tr>
<td>14</td>
<td>0.0255</td>
<td>0.0262</td>
</tr>
<tr>
<td>16</td>
<td>0.0318</td>
<td>0.0099</td>
</tr>
</tbody>
</table>

Jet Length Correlations: inter-jet distance = 0.159 m

<table>
<thead>
<tr>
<th>Flowrate through Jets required to penetrate inter-jet distance</th>
<th>Jet Velocity</th>
</tr>
</thead>
<tbody>
<tr>
<td>(m$^3$s$^{-1}$)</td>
<td>(m s$^{-1}$)</td>
</tr>
<tr>
<td>Zenz</td>
<td>0.0029</td>
</tr>
<tr>
<td>Merry</td>
<td>0.0114</td>
</tr>
<tr>
<td>Shakhova</td>
<td>0.0171</td>
</tr>
</tbody>
</table>

(the correlations are also plotted in figure 3.24)
Chapter 4: Heat Transfer

From figure 4.5 it can be seen that in each case the predicted mobilisation point is higher than the flow via the jets when the increase in the heat transfer coefficient is observed. For run 12, where the lowest total flow rate was used, the predicted value for the jet flow is greater than the total flow. Therefore, from the theory, the bed would not have mobilised during this run, even when all the air was diverted to the jets.

These results indicate that the initial abrupt increase in the heat transfer coefficient does not correspond to mobilisation of the bed as discussed in Chapter 3. However, in these experiments the bed was fluidised before any air was diverted to the jets; therefore the bed is not expected to behave as a rigid body, as was assumed in developing the mobilisation point theory. Momentum from the jets will be transferred to the individual particles, but not to the bed as a whole; this will mainly occur near the base, where the jets are acting. Therefore the rate of change of momentum required to initiate rotation in the region local to the jets is expected to be lower than that required to rotate the whole bed. The increased velocity of the particles in this region will reduce the residence time of individual particles at the wall, and any increase in the particle replacement rate at the wall will lead to an increase in the particle-convective heat-transfer coefficient, and hence the overall coefficient.

4.5.4 Variation in $h_{gw}$ Above the Transition Point

The predicted values from the jet penetration-length correlations, of the flow required to stabilise rotation in the base of the bed, also differ from the experimental values. However, the value obtained from Merry's correlation, which was found to be consistent with the initiation of stable rotation, (see section 3.9), corresponds to the point at which the overall heat transfer coefficient appears to start increasing again (at $Q_j=0.012 \text{ m}^3\text{s}^{-1}$) which is seen in runs 12 and 14. Therefore the stabilisation of the rotation which occurs when the jets are able
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to "link", may improve the heat transfer by reducing the spread of particle residence times and extending the rotating region to the higher parts of the bed, ensuring that all of the bed is well mixed.

The results from earlier measurements, which are plotted in figure 4.3, indicate that there is a minimum jet velocity for particle movement of about 5 m/s. Although this has not been confirmed at low jet velocities, or in the larger rig, the results for heat transfer are consistent with these findings.

4.5.5 Effect of Varying the Total Air Supplied to the Bed

At lower jet flow rates the effect of the overall gas flow rate is more marked — higher gas flow rates give higher overall coefficients. This is mainly due to supplying excess air above that required for fluidisation to the bed. In a fluidised bed of large particles such as this, the excess gas passes through the bed as bubbles, which promote particle mixing in the bed, and the higher superficial gas velocity also increases the gas convective coefficient between the bed and the wall. Therefore the overall coefficient is higher.

As air is diverted to the jets, the mixing due to the bubbles in the bulk of the bed is reduced. At the same time, the air entering through the jets may bypass the bed at the wall. This reduces the degree of mixing between the gas near the wall and that in the bulk of the bed. This may account for the reduction in heat transfer above the transition point for the higher overall gas flow rates.

For the lower overall flow rate used in run 14, the heat transfer in the jet region appears to have a more significant effect on the overall heat transfer rate. In this case the overall coefficient increases as the air flow through the jets is increased. These results are consistent with an increase in the particle-convective
component as discussed above.

The final result from run 12 shows the heat transfer coefficients converging at high flow rates through the jets. Therefore it appears that, at high jet flow rates, the jet flow has a dominant effect on the overall heat transfer coefficient in all cases. This is due to the enhanced gas-convective component at the wall in the vigorously rotating region, and the particle convective component as discussed above. The major part of the heat transfer will therefore occur in this region and the overall gas flow rate is less important. However, more experimentation would be required to prove this.

4.5.6 Conclusions

The mobilisation point for a static bed, which is predicted by using the "rigid body" assumption, cannot be used to predict the effect of jets on a bed which is already fluidised. For a rigid body the mobilisation velocity would fall to zero as the bed approaches a fluidised state, as the weight of the bed would be fully supported.

The heat transfer coefficients measured in the prototype bed are low compared with those found in the majority of fluidised beds. This is due to the particle type used; higher coefficients would be obtained with smaller, denser particles, as are commonly used in fluidised bed applications. The main mechanism of heat transfer between the bed and the wall is gas-convection. Correlations based on data from typical beds can be used to predict both gas-convective and particle-convective heat transfer coefficients, and the results are consistent with the trends predicted for large particles by Baskakov. The coefficients are larger than those observed in spouted beds.

The introduction of part of the air through the wall jets changes the heat transfer behaviour and a step increase in the heat
transfer coefficient is observed; this abrupt increase in the observed heat transfer coefficient corresponds to a change in the bed behaviour which reduces the particle residence time at the wall to a value in the region of one second. However, this point is lower than the mobilisation point for a static bed, and may indicate the initiation of localised rotation in the base of the bed, rather than rotation of the bed as a whole. The increase in the heat transfer coefficient is mainly due to the change in the particle convective component, from a negligible value to 10-20% of the total coefficient.

Although a fall in the heat transfer coefficient is observed as the jet velocity is further increased, at higher jet flow rates the heat transfer coefficient starts to increase again. This point is consistent with the jet penetration length, as predicted by Merry, being equal to the inter-jet distance. This further increase in the heat transfer coefficient is due to shorter particle residence times at the wall and the extension of rotation into the higher parts of the bed. At the highest jet flow rates used, the coefficients appear to converge, so that the coefficient is a function of the jet velocity, rather than the overall flow rate. This indicates that, for high jet velocities, the heat transfer occurs mainly in the region around the jets.
4.6 Drying in the Prototype Spinning Bed

4.6.1 Introduction

In a fermenter application the novel fluid bed would comprise a three-phase system, in which wet particles are fluidised by the gas stream. Water is added to provide the correct culture conditions and the performance of the fermenter depends on maintaining both the correct moisture content and temperature in the bed.

The addition of water affects several aspects of the operation of the bed. On a trivial level, the particles are heavier, which increases the air flow required to mobilise the bed; they may also be more cohesive (and adhesive), due to the nutrient medium on the particle surface. More importantly, the heat transfer within the bed is altered by the evaporative heat transfer at the particle surface. This increases the capacity of the air to remove heat from the bed, and hence we expect the overall rate of heat transfer to be higher than that observed in a mobilised bed of dry particles. The effect and limitations of evaporative cooling for temperature control in fermenters have been discussed in Chapter 1. If evaporation occurs, water has to be simultaneously added to the system to maintain the optimum bed conditions.

In this section of the work the drying rate in the prototype bed was measured over time, and the heat transfer to wet particles was studied. The effect of adding water to the bed to maintain constant conditions was also examined. Although the experiments were designed to provide information about the conditions in a fluidised bed fermenter, the results also shed light on the effectiveness of the novel bed design as a dryer for very wet or cohesive particles.

4.6.2 Theory of Gas-Fluidised-Bed Drying

The use of fluidised beds as dryers has been reviewed by Reay.
Drying is discussed either as a batch process, in which the particles become drier with time, or as a continuous process in which particles are continually added to and removed from the dryer. In the latter case, although the average conditions in the process may be constant, the individual particles are undergoing batch drying, with a variable drying period. This differs from the conditions in a fermenter in which the rate of water removal from the particles is equal to the rate of water addition, and an individual particle does not become drier over time.

In a batch dryer the rate of drying varies over the drying period as shown in figure 4.8. Initially the drying rate is constant, as the surface of the particles is saturated with water, and the rate of evaporation is limited by external factors such as the transfer of heat to the particle, mass transfer across the boundary layer and the humidity of the air; this is usually referred to as the constant rate period.

With porous materials, the drying rate begins to change at a point known as the critical moisture content. When the surface water has been removed, water begins to migrate from within the particle to the surface. Within the particle the main transport process may be liquid diffusion, capillary flow, vapour diffusion in pores, or a combination of these in series or parallel; various mechanisms are discussed by Legros (1986) in his thesis on the drying of cut lamina tobacco. In some materials, subsequent drying is limited by the rate at which internal water migrates to the surface, and as can be seen above, if the maximum drying rate is a function of the drying conditions – air flow rate, temperature etc. – then the critical moisture content is also a function of the drying conditions. Alternatively, if the internal mass transfer is fast, the drying rate will begin to fall when the surface begins to dry out, reducing the area available for mass transfer. In this case the critical moisture content will depend more on particle properties, such as size and surface area, than
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on the drying conditions.

Conditions Within the Particles
For the purpose of fermentation each particle in the fluidised bed may be considered to act as a bioreactor, in which the correct operating conditions have to be maintained. Depending on the organism, the required moisture content may be above or below the critical moisture content; in either case it is important not to have temperature or moisture gradients within the particles.

If evaporation is occurring, water must be added to maintain the optimum moisture content. Even when water is continually sprayed into the bed, an individual particle experiences water addition periodically. Therefore a degree of mixing is required in the bed, so that the range over which the moisture content varies between water additions is within the required operating conditions. Above the critical moisture content the particle surface is saturated, so the conditions within the particle will remain relatively constant between additions. In the falling rate period, the internal conditions vary in response to the drying time, so the frequency of addition is more important.

Heat transfer to the particle and mass transfer to the particle both occur in series, firstly across the boundary layer at the surface of the particle, and secondly within the particle. The conditions within the particle will be uniform if the rates of transfer within the particle are higher than the external transfer rates, even in the falling rate period. However, if the addition of moisture is intermittent, transient gradients will appear between water additions.

It is possible to calculate whether the internal or external resistance to mass transfer is limiting; if the latter is the case, the conditions inside a particle will be constant over the particle diameter, though varying in time, even below the critical
moisture content.

In their discussion of heat and mass transfer processes for particles in fluids, Clift et al (1978) show that the external resistance is dominant if:

\[
\frac{\overline{X} D_p}{D} >> \text{Sh} \quad \text{where} \quad \text{Sh} = \frac{k d}{D} \tag{4.24}
\]

where \( \overline{X} \) is Henry's coefficient, \( k \) is a coefficient of mass transfer or

\[
\text{Bi} = \frac{h_{\text{ext}} d_p}{k_p} \ll \frac{c_f}{c_p} \tag{4.25}
\]

where \( D \) is the diffusion coefficient in the fluid, and \( D_p \) is the diffusion coefficient in the particulate phase.

This analysis shows a strong correlation between particle size and the significance of the internal resistance; it is apparent that the internal resistance can be reduced by reducing the particle size.

It is clearly of importance to estimate the Blot number for the experimental system. In the absence of specific data for the coated particles or puffed wheat used in these experiments, the published values for cellulose have been used, as the nearest organic material for which values are available in the literature.

<table>
<thead>
<tr>
<th>Property</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Particle diameter</td>
<td>( 8.6 \times 10^{-3} ) m</td>
</tr>
<tr>
<td>Particle thermal conductivity ( k_p )</td>
<td>2.6 W m(^{-1})K(^{-1}) (cellulose)</td>
</tr>
<tr>
<td>Particle heat capacity ( c_p )</td>
<td>1339 J kg(^{-1})K(^{-1}) (cellulose)</td>
</tr>
<tr>
<td>Fluid heat capacity ( c_f )</td>
<td>1050 J kg(^{-1})K(^{-1})</td>
</tr>
<tr>
<td>External heat transfer coefficient ( h_{\text{ext}} )</td>
<td>59.7 W m(^{-2})K(^{-1}) (section 4.2.4)</td>
</tr>
<tr>
<td>Blot number:</td>
<td>0.20 (from equation 4.25)</td>
</tr>
<tr>
<td>Ratio of heat capacities:</td>
<td>1.3</td>
</tr>
</tbody>
</table>

In this case the Blot number and the ratio of heat capacities are
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of a similar magnitude. Therefore although the ratio of heat capacities is higher, suggesting external control, neither of the resistances is dominant and intra-particle gradients may arise.

Equilibrium Moisture Content
In a fluidised bed the instantaneous drying rate of a particle is a function of the particle surface moisture. If the external resistance is limiting this is representative of the moisture level throughout the particle, even in a batch drying process. However, if the internal resistance is limiting, there will be an internal moisture gradient and the mean moisture concentration will be higher than the surface concentration.

When designing a fermentation process to operate at a specific water loading, this effect becomes significant. If the external resistance is limiting, the batch drying curve can be integrated to find the instantaneous drying rate at the required water loading, and hence the water addition rate can be set accordingly. If the internal resistance is limiting, the intra-particle gradients will decay over time if water is added continuously, and the equilibrium moisture content will be less than required. Furthermore, if the water is added to an individual particle intermittently, the slow rate of diffusion into the particle from the surface will create additional moisture gradients. In the latter case the specific water addition rate required to maintain the desired equilibrium moisture content can only be found by experimentation.

4.6.3 Experimental Equipment and Method
The experiments on drying were done in the same jacketed rig as the heat transfer experiments described above. Heat was supplied through the wall of the bed as before. To monitor the drying performance of the bed, two humidity probes were installed on the inlet and outlet gas streams. The electronic probes, manufactured by "Vaisala", were mounted in branches on the inlet pipe and the
top of the bed, and were fed by a controlled bleed from the main flow through a ball valve. For the initial batch drying experiments, a measured charge of wetted material was introduced to the bed. The material was mobilised and the temperature and relative humidity of the air flow into and out of the bed were logged, as well as the bed wall temperature, the water inlet and outlet temperatures, and the particle surface temperature (measured by an IR sensor as described in section 4.3.2).

Calculation of Drying Rate

The drying rate was measured by monitoring the change in relative humidity across the bed. This can be used to obtain the instantaneous rate, expressed as the weight of water removed per second by the air stream. The specific drying rate, expressed as the weight of water removed per unit weight of bed per second can be derived from this, and as the same weight of particles was used for each run, conclusions about the bed drying rate are valid for the specific drying rate as well.

Accuracy of the Measurement System

The accuracy of the measurement system was checked by performing a mass balance over the first run, by numerical integration of the water removal rate. The results are summarised in Table 4.8.

<table>
<thead>
<tr>
<th>Description</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Initial water loading (g)</td>
<td>100</td>
</tr>
<tr>
<td>Total water removed (g)</td>
<td>91.1</td>
</tr>
<tr>
<td>Residual water (g)</td>
<td>10.1</td>
</tr>
<tr>
<td>Total water removed plus residual (g)</td>
<td>101.2</td>
</tr>
</tbody>
</table>

Calculation of Heat Transfer Rates

The heat transfer from the bed consists of the latent heat taken up by the water when it evaporates, and the sensible heat carried by the air-stream. The mean heat removal rate in the constant
rate period was calculated from the rate of water removal and the
temperature change in the gas measured across the bed during the
experiments. The effective heat transfer coefficient is
calculated from the sum of the two heat fluxes; the sensible heat
removed in the air-stream was calculated using the method
described in section 4.3.3, (equation 4.21). From the
experimental results, the mean gas and particle temperature is
taken to be 25 °C and the mean wall temperature is 54 °C. The
wall area of the bed available for heat transfer is 0.164 m².

Equilibrium Experiment

The results from these batch experiments were compared with an
equilibrium drying experiment. For this the bed was charged and
run as before, but additional water was sprayed into the bed
continuously, using a spray nozzle submerged in the bed.

4.6.4 Results

The data from a typical batch drying run are shown in figure 4.9,
in which the temperature and relative humidity of the inlet and
outlet air are plotted against time. The inlet relative humidity
is plotted to confirm that the inlet conditions remain constant
during the experiment and to provide a base line for calculating
the relative humidity increase across the bed. During the course
of the experiment there is little variation in the inlet
conditions.

In the initial part of the experiment the gas leaving the bed has
a measured relative humidity of 97%, so that the air can be
considered to be saturated within the range of experimental error
(see appendix B); this is confirmed by the results in which a
constant rate drying period followed by a falling rate drying
period can be identified. Graphs of drying rate versus time,
Figs. 4.10 to 4.13, and drying rate versus moisture content
(Figs. 4.14 to 4.18) are given below.
Four batch drying experiments were carried out, with slightly different experimental conditions. The operating conditions for these runs are tabulated in Table 4.9.

<table>
<thead>
<tr>
<th>Run No.</th>
<th>water added (kg)</th>
<th>initial air flowrate ($\text{m}^3\text{s}^{-1}$)</th>
<th>air through</th>
<th>jets</th>
<th>base</th>
</tr>
</thead>
<tbody>
<tr>
<td>291</td>
<td>0.1</td>
<td>0.0229</td>
<td>yes</td>
<td>no</td>
<td></td>
</tr>
<tr>
<td>30</td>
<td>0.2</td>
<td>0.0417</td>
<td>yes</td>
<td>yes</td>
<td>yes</td>
</tr>
<tr>
<td>303</td>
<td>0.1</td>
<td>0.0288</td>
<td>yes</td>
<td>yes</td>
<td>yes</td>
</tr>
<tr>
<td>304</td>
<td>0.1</td>
<td>0.0293</td>
<td>yes</td>
<td>yes</td>
<td>yes</td>
</tr>
</tbody>
</table>

In all but run 291, which ended when the particles were blown out of the bed, the air flow rate was adjusted during the run to maintain uniform mobilisation in the bed.

The data for the four batch drying runs are plotted as drying rate versus time in Figs. 4.10 to 4.13, and despite the noise apparent in the data, it is possible to identify a short period of constant rate drying followed by a falling rate period during each run. The later part of each run was affected by the need to reduce the air flow to prevent the particles from being blown out of the bed as they became drier. This has been accounted for in calculating the drying rate from the flow rate and relative humidity data.

The constant rate period is measured from the point at which the bed begins to spin, shortly after the air is turned on. Drying continues until the air supply is stopped, either because drying is complete or because the bed has defluidised. The results from the four runs are summarised in Table 4.10.
### Table 4.10. Summary of Drying Runs: Results

<table>
<thead>
<tr>
<th>Run No.</th>
<th>291</th>
<th>30</th>
<th>303</th>
<th>304</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mass of particles (kg)</td>
<td>0.065</td>
<td>0.065</td>
<td>0.065</td>
<td>0.060</td>
</tr>
<tr>
<td>Initial water loading (kg)</td>
<td>0.100</td>
<td>0.200</td>
<td>0.100</td>
<td>0.100</td>
</tr>
<tr>
<td>Initial flow rates</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>total (measured (m³ s⁻¹))</td>
<td>0.0229</td>
<td>0.0417</td>
<td>0.0288</td>
<td>0.0293</td>
</tr>
<tr>
<td>jets (measured (l/min))</td>
<td>0.0217</td>
<td>0.0224</td>
<td>0.0197</td>
<td>0.0175</td>
</tr>
<tr>
<td>base (calculated) (l/min)</td>
<td>0.0012</td>
<td>0.0193</td>
<td>0.0091</td>
<td>0.0118</td>
</tr>
<tr>
<td>Constant rate period</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>start (mins)</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>finish (mins)</td>
<td>0.8</td>
<td>2.9</td>
<td>0.4</td>
<td>0.6</td>
</tr>
<tr>
<td>water removed (kg)</td>
<td>0.0214</td>
<td>0.1291</td>
<td>0.0117</td>
<td>0.0193</td>
</tr>
<tr>
<td>rate (x 10⁻³ kg s⁻¹)</td>
<td>0.446</td>
<td>0.742</td>
<td>0.488</td>
<td>0.536</td>
</tr>
<tr>
<td>moisture in air (kg water/kg air)</td>
<td>0.0151</td>
<td>0.0138</td>
<td>0.0131</td>
<td>0.0142</td>
</tr>
<tr>
<td>Critical moisture content (kg/kg)</td>
<td>0.94</td>
<td>1.00</td>
<td>1.15</td>
<td>1.05</td>
</tr>
<tr>
<td>falling rate period</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>start (mins)</td>
<td>0.8</td>
<td>2.9</td>
<td>0.4</td>
<td>0.6</td>
</tr>
<tr>
<td>finish (mins)</td>
<td>2.7</td>
<td>4.1</td>
<td>3.0</td>
<td>3.8</td>
</tr>
<tr>
<td>water removed (kg)</td>
<td>0.0423</td>
<td>0.0443</td>
<td>0.0625</td>
<td>0.0766</td>
</tr>
<tr>
<td>rate of fall (x 10⁻⁶ kg s⁻²)</td>
<td>1.31</td>
<td>3.52</td>
<td>1.12</td>
<td>1.43</td>
</tr>
<tr>
<td>Maximum rate of heat transfer (mean rate: constant rate period)</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>latent heat (W)</td>
<td>1009</td>
<td>1657</td>
<td>1091</td>
<td>1148</td>
</tr>
<tr>
<td>sensible heat (W)</td>
<td>73</td>
<td>161</td>
<td>60</td>
<td>83</td>
</tr>
<tr>
<td>total (W)</td>
<td>1082</td>
<td>1808</td>
<td>1151</td>
<td>1231</td>
</tr>
<tr>
<td>effective heat transfer coefft. (W m⁻² K⁻¹)</td>
<td>227</td>
<td>380</td>
<td>242</td>
<td>259</td>
</tr>
</tbody>
</table>

### 4.6.5 Discussion of Results

**Drying Rate**

The rate of drying during the constant rate period is a strong function of the gas flow rate. As the air leaving the bed is saturated, and the drying rate per unit volume of gas is
relatively constant, at about 0.014 kg water/kg air, this is consistent with the drying rate being limited by the capacity of the air to carry water vapour.

The critical moisture content is similar for the runs using similar flow rates, but higher when the gas flow rate is higher, in run 303. The rate of mass transfer from the interior of the particle to the particle surface decreases as the surface layer of the particle becomes depleted. The critical moisture content is reached when the rate of migration of water to the particle surface equals the constant drying rate. Therefore the higher value for run 303 occurs because of the higher rate at which water is removed in the air stream during the constant rate period.

The rate of fall of the drying rate in the falling rate period also appears to be a function of the total air flow rate, (see Table 4.6). This arises because the time taken to dry the particles after the critical moisture content is reached is a function of the internal resistances, not the external conditions. Therefore the falling rate period is expected to have approximately the same duration if the critical moisture content is similar. However, the rate of fall is steeper for run 303 because the critical moisture content is higher.

The difficulties encountered in trying to control the bed towards the end of each run make it difficult to estimate the real duration of the falling rate period.

The drying rate in the falling rate period can be plotted as a function of the particle moisture content, and this is shown in Figs. 4.14 to 4.17. When the graphs, which are plotted on the same scale, are superimposed, (see Fig. 4.18) the drying rate is seen to be a linear function of the particle moisture content, and can be expressed as:
water removed (kg/s) = 0.0045 (bed moisture content(kg)) + 0.0002

Despite the uneven nature of the results, this demonstrates that the drying rate during this part of the run was controlled by the particle moisture content, rather than the gas flow rate and relative humidity.

Heat Transfer
The increase in measured heat transfer coefficients indicates the effect of evaporative cooling on the heat transfer performance of the bed. The overall heat transfer coefficients between the wall and the gas are between 227 and 380 W m⁻² K⁻¹ in the drying experiments compared to between 50 and 100 W m⁻² K⁻¹ in the heat transfer experiments using dry particles. At similar air flow rates e.g. at 0.0288 m³ s⁻¹, the coefficient is a factor of 2.5 higher with evaporative cooling.

The increase in the overall heat transfer coefficient is achieved through the latent heat removed by the evaporation of water into the gas stream. The latent heat taken up by the water is in addition to the sensible heat which the air-stream takes up when it comes into equilibrium with the wall and the bed.

The capacity of the air to absorb heat depends on the air flow
Chapter 4: Heat Transfer

rate and relative humidity, which results in a limitation being placed on the extent of evaporative cooling for fermentation applications. Some extra drying capacity can be obtained by drying the air before it enters the bed, but the extent to which this can be done is limited, and once the air leaving the bed is saturated the total cooling duty can only be further increased by increasing the flow of air through the bed. Increasing the temperature gradient at the wall would increase the sensible heat carried by the gas, but this has a relatively small effect on the overall rate of heat transfer. The implications of this result will be discussed in more detail in chapter 5.

4.6.6 Equilibrium Drying Run

In the discussion of the theory of drying in a fluidised bed above, it was shown that the conditions within the particle during batch drying depend on the mechanism of control of the drying rate. If the drying rate is controlled by the rate of diffusion within the particle, gradients arise between the centre of the particle and the surface. However, if transport across the boundary layer into the gas stream is limiting, gradients within the particle do not develop, and the moisture within the particle is evenly distributed.

In order to differentiate between these two conditions for the system above, it was decided to perform an equilibrium drying experiment. The aim of this was to determine whether, during the falling rate period, gradients in moisture content developed within the particles. This is important, for two reasons. Firstly, in specifying the operating conditions, because if the mass transfer is internally controlled, it is impossible to predict the moisture addition rate required to maintain the optimum moisture level in the bed, if this is below the critical moisture content, from a batch drying experiment. Secondly, from a biological point of view, moisture gradients within the particle will lead to some parts of the particle varying from the optimal
moisture content, and this will affect the organisms, and reduce the overall productivity of the bioreactor.

During the experiment water was continuously added to the bed at a rate of 14 g/min which corresponds to a drying rate between the constant drying rate and the lowest rate observed in the falling rate period for the first batch drying run. This drying rate corresponds to a bed water content of 30 g during run 291. However it also corresponds to a value of 7.4g taken from the other batch drying runs, when this is calculated using the generalised function described in section 4.6.5.

When water is added continuously in this way, the bed will reach an equilibrium state where the drying rate is equal to the water addition rate. If gradients are present during the batch runs, these should decay during the equilibrium run and the equilibrium bed moisture content should be lower than the moisture content at the same instantaneous drying rate during a batch run. However if drying is externally controlled, the bed moisture at the same drying rate will be the same for the equilibrium run and the batch run.

For this experiment heat was supplied to the bed through the jacketted wall as before. The initial moisture content was 1.75 g water per gram (dry weight) of particles. The water delivery system consisted of a spray nozzle mounted 4 cm deep in one side of the bed, fed from a peristaltic pump. Subsequent experiments showed that this was critically dependent on good particle circulation to distribute the water throughout the bed. However a suitable replacement was not available.

Results
The run lasted for 36 minutes; at the end of the run the particles were reweighed and found to contain 2.5 g of water. The humidity and temperature readings during the run are shown in figure 4.19,
Chapter 4: Heat Transfer

which shows that the outlet relative humidity remained high throughout the run. Within the limits of experimental error (see below) the outlet air can be taken to be saturated initially; later in the run the relative humidity of the outlet air falls below the saturation value. As the run proceeded the particles became lighter and the air flow rate was reduced to prevent particle carryover.

The total moisture added and the moisture removed are plotted against time in figure 4.20. As the water removal rate exceeds the water addition rate towards the end of the run, and integration of the rates suggests that more water was removed than was added in total, this indicates that the measurements show an error of approximately +/-30% However the rate does reach a steady value, at around 18 g min⁻¹, which suggests that equilibrium had been reached. The principal source of error in the results arises due to errors in the relative humidity measurements. The instantaneous drying rate is shown against time in figure 4.21; this shows that the drying rate decreases over time as the air flow is reduced until it reaches equilibrium where the drying rate approaches the water addition rate.

Discussion of the Results of the Equilibrium Drying Experiment
During the initial part of the run the outlet air was saturated, as the excess water in the particles was removed, but during the run the outlet relative humidity fell, indicating that the rate of drying was not limited by the relative humidity of the air.

The equilibrium water content, at 0.044 g water per gram (dry weight) of particles, is lower than both the value taken from batch run 291 and the rate calculated from the drying rate function. This proves that drying during the batch experiments described above was controlled by internal diffusion within the particles.
Therefore, where it is essential to control the moisture content throughout the particle at a constant value, the bed should either be run above the critical moisture content, so the drying rate is controlled by the humidity of the air, or smaller particles should be used. As the Blot number is a function of the particle diameter, reducing this should improve the control of the moisture content; for example, reducing the particle diameter to 1.5 mm would reduce the Blot number to one twentieth of the ratio of the specific heats. With smaller particles, further experiments will be required to confirm that the size predicted by the Blot numbers is suitable.

4.6.7 Drying Experiments: Overall Conclusions

The drying experiments demonstrate that the prototype bed can be used as an effective dryer: foam particles are mobilised even at high water contents corresponding to ten times the particle dry weight. These results are consistent with the earlier quantitative experiments described in Chapter 2 in which sticky coated particles were successfully mobilised. Therefore the design is effective for drying difficult materials.

The particles can be mobilised over a range of moisture contents, and in batch drying the particles are seen to exhibit a constant drying rate period and a falling drying rate period. The rate of drying is initially limited by the saturation of the air, indicating that heat transfer between the particles and the gas is effective. The saturation point of the air would limit the amount of cooling provided by evaporation to a fluidised bed fermenter and this may have to be augmented by cooling through the walls or internal heat transfer surfaces.

The possibility of accurately controlling the moisture level in a fermentation application depends on the moisture level required and the critical moisture content of the particles. The system will be easier to control if the desired moisture level is above
Chapter 4: Heat Transfer

the critical value. In addition the culture conditions in the particle are improved if the moisture content is above the critical value, as the organisms do not experience cycling between wet and dry conditions between water additions.

The critical moisture content of the particles is itself a function of the drying rate in the constant rate period, and it is reached when the supply of moisture to the surface falls below the rate at which water can be removed by the air. Therefore the critical moisture content can be manipulated by varying the relative humidity of the drying air, to reduce or increase the maximum drying rate. This approach could be used to improve the control of lower moisture levels in the bed.

For smaller particles lower moisture contents can be controlled if the internal resistance to mass transfer is less than the external resistance. This can be predicted from the Blot number, using equation 4.25, i.e. the external resistance is limiting when:

\[ B1 = \frac{h_{ext}}{k_p} \frac{d_p}{c_f} \ll \frac{c_f}{c_p} \]

Smaller particles have a number of advantages in addition to the improved moisture control. The fluidisation velocity for smaller particles is less, reducing the amount of air required. In addition, diffusion paths are shorter for nutrients, dissolved gases and waste products, which provides better control of mass transfer.

Therefore a suitable system for use as a gas-fluidised bed fermenter would use small support particles. It requires good control of the relative humidity of the inlet air, and an accurate and effective water addition system which adds water evenly to all the particles. Additional heat removal might be required and could be provided through the use of a cooled jacket or cooled...
Chapter 4: Heat Transfer

internal fins. Any internals could be placed in the upper part of the bed, outside the vigorously spinning region, to avoid disturbing the flow patterns at the base.
4.1 Mechanisms of Heat Transfer in a Gas-Fluidised-Bed

- Particle convection: particle to wall
- Gas convection: gas to wall
- Particle to gas
- Radiation:
Chapter 4: Heat Transfer

4.2 Variation in Bed To Wall Heat Transfer Coefficient with Particle Size: carborundum particles, (Baskakov 1973)

![Diagram showing variation in heat transfer coefficient with particle size.](image-url)
Chapter 4: Heat Transfer

4.3 Horizontal Particle Velocity v Gas Velocity at the Jets

6" Bed Horizontal Velocity of Particles

particle velocity v jet velocity

jet velocity (m/s)

particle velocity (m/s)
Chapter 4: Heat Transfer

4.4 New Bed for Heat Transfer Experiments
Chapter 4: Heat Transfer

4.5 Measured Heat Transfer Coefficients v Flow Diverted to the Jets

All Runs Plotted on the Same Axes

Gas-convective Heat Transfer

Coefficient versus flow through jets

heat transfer coefficient (W/m² K)

flow through jets (m³/s)

+ total = 0.030 m³/s

△ total = 0.0296 m³/s⁻¹

○ total = 0.032 m³/s

□ total = 0.025 m³/s⁻¹
Chapter 4: Heat Transfer

4.5 Measured Heat Transfer Coefficients v Flow Diverted to the Jets

b Run 12 and Run 13: total flowrate = 0.0296 m$^3$s$^{-1}$

Runs 12–13 Gas–convective Heat Transfer

Heat transfer coefficient (W/m$^2$/K) vs flow through jets (m$^3$/s)

at mobilisation point flow through jets is 0.0170 m$^3$s$^{-1}$
Chapter 4: Heat Transfer

4.5 Measured Heat Transfer Coefficients v Flow Diverted to the Jets

Run 14: total flowrate = 0.0255 m$^3$s$^{-1}$

Run 14 Gas—convective heat transfer coefficient versus flow through jets

At mobilisation point flow through jets is 0.0262 m$^3$s$^{-1}$
4.5 Measured Heat Transfer Coefficients v Flow Diverted to the Jets

Run 16: total flowrate = 0.0318 m$^3$s$^{-1}$
4.6 Heat Transfer Coefficient as a Function of the Total Air Flow Rate: no air diverted to the jets

\[ h_{gw} \text{ vs Superficial Velocity in Bed} \]

No air Diverted to the Jets

\[ h_{gw} \text{ (W/m}^2\text{K)} \]

Superficial Velocity (m/s)
Chapter 4: Heat Transfer

4.7 Particle Convective Heat Transfer Coefficient v Residence Time at the Wall

Particle-Convective Heat Transfer
Coefficient v Residence Time at Wall

![Graph showing the relationship between particle convective heat transfer coefficient and residence time at the wall]
4.8 Theoretical Batch Drying Curve

Chapter 4: Heat Transfer
4.9 Experimental Data:
Temperature and Relative Humidity v Time, run 291

Experimental Data: Drying Run 291

% RH and Temperature v Time

%RH out

%RH in

temp out
temp in

time (mins)

% RH : Temperature (°C)
4.10 Drying Rate v Time: run 291

Drying run 291
water removal rate v time

Drying rate as a function of time for run 291.
Chapter 4: Heat Transfer

4.11 Drying Rate v Time: run 30

Drying run 30
water removal rate v time

water removal rate (kg/s)

0.0001
0.0002
0.0003
0.0004
0.0005
0.0006
0.0007
0.0008
0.0009
0.0010

0
1
2
3
4

time (minutes)
4.12 Drying Rate vs Time: run 303

Run dry303

water removal rate vs time

water removed (kg/s)

0 0.0001 0.0002 0.0003 0.0004 0.0005 0.0006 0.0007 0.0008 0.0009 0.001

0 1 2 3 4 5 6 7 8

time (mins)
Chapter 4: Heat Transfer

4.13 Drying Rate v Time: run 304

Run dry304

water removal rate v time

water removal rate (kg/min)

Time (mins)
Chapter 4: Heat Transfer

4.14 Drying Rate as a Function of Particle Moisture Content:
run 291

Drying run 291
water removal rate v bed moisture
Chapter 4: Heat Transfer

4.15 Drying Rate as a Function of Particle Moisture Content: run 30

![Drying run 30](image)

Drying run 30
water removal rate v bed moisture
4.16 Drying Rate as a Function of Particle Moisture Content:
run 303

Run dry303
water removal rate v bed moisture

![Graph showing water removal rate as a function of bed moisture for run 303. The x-axis represents bed moisture (kg) and the y-axis represents water removal rate (kg/a). The graph shows a trend where water removal rate increases with increasing bed moisture.]
Chapter 4: Heat Transfer

4.17 Drying Rate as a Function of Particle Moisture Content:
run 304

Run dry304

water removal rate v bed moisture

water removed (kg/s)

bed moisture (kg)
Chapter 4: Heat Transfer

4.18 Drying Rate as a Function of Particle Moisture Content: runs 291, 30, 303, 304, superimposed

Drying Runs Superimposed

water removal rate vs bed moisture content

![Graph](image_url)
Chapter 4: Heat Transfer

4.19 Experimental Data: Equilibrium Drying Run

Equilibrium Drying Experiment
Temperature and %RH v time

% RH out
% RH In
Temp In
Temp Out

Time (minutes)
4.20 Moisture Added (cumulative) and Moisture Removed (cumulative) v Time

Equilibrium Drying Run
Water added and water removed v time

water (cumulative) (kg)

0.8
0.7
0.6
0.5
0.4
0.3
0.2
0.1
0

0 20 40

time (mins)

water removed

water added
4.21 Instantaneous Drying Rate v Time: Equilibrium Drying Run

Equilibrium Drying Experiment

Water removal rate versus time

kg water/s

time (minutes)
Chapter 5: Conclusions and Recommendations for Further Work
Chapter 5: Conclusions and Recommendations for Future Work

Contents

5.1 Introduction
5.2 Summary of the Work Completed for this Project
5.3 Results
  5.3.1 Hydrodynamics
  5.3.2 Heat Transfer
  5.3.3 Principles For Scale-up
5.4 Future Work
  5.4.1 Hydrodynamics
  5.4.2 Spouted Spinning Bed
  5.4.3 Scale-up
  5.4.4 Fermentation
  5.4.5 Drying
5.5 Industrial Applications
Chapter 5: Conclusions and Recommendations for Future Work

5.1 Introduction
This work was started as a continuation of an internal ICI research project into the use of a gas-phase-continuous three-phase fluidised bed for fermentation. The aim of the project at that stage was to address two problems identified by the team at ICI. In the first instance, it had been identified that it would be difficult to prevent agglomeration and adhesion of particles to the walls of the bed, while allowing even distribution of moisture and maintaining the optimum conditions of moisture. It had been recognised at ICI that the optimum growing conditions required a moisture level which was close to the wet quenching point of the fluidised bed which they were using; therefore the operating window between achieving optimum growth conditions and loss of fluidisation was narrow. In the second instance, a study of the heat balance in the bed indicated that evaporative cooling would be insufficient to remove metabolic heat if the bed design was scaled up; therefore an investigation of the heat transfer properties of the bed, to determine the coefficient of heat transfer between the bed and the wall was required.

5.2 Summary of the Work Completed for this Project
An initial survey of the published literature indicated a range of development activity in the area of fluidised bed fermentation. A number of different culture systems had been studied at a laboratory scale, and a variety of products had been produced. However the work in this area has been limited compared with the huge amount of literature, mainly from the food industry, which refers to solid state fermentation in static or stirred fermenters. The phenomenon of fluidisation is also covered by a large range of published literature, within which the study of three-phase fluidisation, particularly gas-fluidised three-phase fluidisation, forms only a minor part. The majority of work with three-phase fluidised beds concerns dryers. Therefore, although there is a lot of material of indirect interest to this project, the published material which is of direct interest in this field is limited. In particular, very little practically useful work has been carried out on the problem of cohesive beds and maintaining fluidisation with cohesive and adhesive materials.
Based on the literature it was possible to assess the factors affecting the success of a solid state fermentation system, and to use these to define the optimum operating conditions in a three-phase fluidised bed for a fermentation application.

In the first part of the work various alternative distributor designs were tested with wet or cohesive particles, in an attempt to identify a design which would have a wider operating window and could handle wetter particles than the conventional fluidised bed design. In general these were designs which have been developed for other particle and gas contacting duties such as drying processes. On the basis of this work a modified design was developed which prevented adhesion and agglomeration in the base of the bed. This was achieved by introducing some of the fluidising air through jets in the walls of a cylindrical bed, and inducing a vigorous rotation of the lower part of the bed about a vertical axis. This bed was tested qualitatively by using it to mobilise a porous foam particle coated in a thick viscous suspension of microbial nutrients.

This spinning bed design represents a departure from designs described in the literature. Therefore a study was made of the forces acting in the bed and the forces exerted by the jets, to provide a method of predicting the point of mobilisation of the bed. An analysis was developed, which used Janssen’s approach to the calculation of the stress distribution in hoppers to predict the frictional forces at the wall resisting rotation of the bed, and therefore permitted calculation of the tangential force from the jets required to initiate motion.

In the second part of the work some of the heat transfer properties of the bed were studied; gas convection is the principal mechanism of heat transfer for systems using large particles (> 1 mm diameter), as was the case in this work. The overall coefficient of heat transfer between the bed and the wall was obtained by measuring the heat transfer rate between the heated wall and the gas. This provided an estimate of the gas
convective heat transfer coefficient between the wall and the bed, which indicates the scope for removing heat from the bed via submerged surfaces.

A review of the results in the context of possible industrial applications has been prepared at the request of ICI. This includes a summary of the design correlations and a simple cost comparison between conventional and fluid bed processes.

5.3 Results
This work has produced a design for a novel fluidised bed which is suitable for three-phase fluidisation with water as the dispersed phase. This makes it suitable for fermentation applications as well as more conventional uses including drying processes.

The operating window has been widened by developing a design which keeps the particles in motion throughout the bed, and breaks up loose agglomerates. This prevents wet quenching at moisture contents above the wet quenching point for a conventional fluidised bed.

Heat transfer coefficients have been measured for the new design, and the mechanisms of heat transfer in the bed have been investigated. Suitable correlations have been identified from the literature for predicting changes in the heat transfer coefficients with different particles.

5.3.1 Hydrodynamics
An analysis of the hydrodynamics has been developed in order to predict the mobilisation point of the bed. This theory agrees with experiment for a range of different materials and particle sizes. A method has also been developed, based on the jet penetration length theories, to predict the conditions of stable mobilisation (rotation) of a fluidised bed.

5.3.2 Heat Transfer
Experiments on the heat transfer properties of the bed have confirmed that gas convection is the main mechanism of heat
transfer between the particles and the wall of the bed. The heat transfer coefficient is between 50 and 100 W m\(^{-2}\)K\(^{-1}\). Heat transfer between the bed and the wall is enhanced by up to 20 W m\(^{-2}\)K\(^{-1}\) when the bed is mobilised, due to an increase in the particle convective heat transfer coefficient when the particle residence time at the wall is reduced.

5.3.3 Principles For Scale-up
The mobilisation point of a static bed can be predicted from the calculated frictional forces acting at the wall and base of the bed, balanced by the torque exerted by the jets. Correction must be made for the drag force exerted by the upward flow of air through the bed. This has been tested in beds with four equally-spaced jets, for diameters of 145 mm and 225 mm and aspect ratios (H/D) between 0.5 and 2.0. For beds outside this range it is recommended that further pilot trials are used to confirm these predictions.

The minimum air flow required through the jets in order to mobilise the bed can be reduced by supplying a flow of fluidising air to the base of the bed. The drag force exerted by the air reduces the frictional resistance at the wall, and hence the torque required to spin the bed.

The size and number of the jets can be varied, so long as the torque calculated from the analysis can be applied to the bed with the available air supply. It is recommended that the total air supplied to the bed should not be more than twice the minimum fluidising velocity for non-cohesive particles of the same size and density as the material being used, to prevent carryover from the top of the bed.

A suitable jet spacing, which will give an indication of the number of jets required can be chosen based on Merry's correlation for the jet penetration length. The jets should be arranged so that each jet is angled to interact with the next, to promote entrainment. Entrainment of the jets stabilises the flow in the base of the bed and promotes stable mobilisation. This approach
can also be used to specify the jet arrangement required to promote rotation in the base of a standard bed of fluidised particles, if this is required.

5.4 Future Work
5.4.1 Hydrodynamics
The theory which has been developed uses a number of approximations to derive the theoretical model. There is scope for more detailed work, using materials for which the angle of internal friction for example, has been determined. In addition, a more rigorous analysis of the wall friction, including the determination of the direction of the resultant of the wall friction in the spinning bed, would provide a more complete understanding of the mobilisation point.

5.4.2 Spouted Spinning Bed
It is possible to modify the bed by using a spout distributor, rather than the perforated plate used during this study. This promotes regular circulation between the top and the base of the bed, and will produce very consistent conditions throughout the bed. However the use of a spout plate complicates the analysis of the force balance in the bed and the theory needs to be modified in order to extend it to cover this case.

5.4.3 Scale-up
This study has only covered a limited range of bed sizes, suitable for laboratory tests. For most practical applications a larger fermenter or dryer would be used and therefore tests on larger units are needed to confirm and extend the theory on a wider range of bed and particle sizes.

The extension of the work to very large systems is likely to require a change in the approach - to multiple spinning cells for example - to extend the influence of the jets into the centre of a large bed. There may be scope to apply a torque on the bed via angled jets in a modified base-plate.
5.4.4 Fermentation
Work on fermentation in the novel design was not carried out. For any proposed application, comparison between the fluidised bed and a conventional submerged culture or solid state system is required to assess the advantages, if any, of the fluidised system. Of particular interest would be a comparison with a fermenter system based on a conventional gas-phase-continuous fluidised bed, as described in the literature, to assess whether the higher liquid loading possible with this design confers a significant advantage.

For any potential fermentation application a range of development work is required: to determine the effect of temperature and moisture gradients on the organism and to set the optimum particle size. The appropriate down-stream processing methods will also have to be developed.

5.4.5 Drying
Other possible applications for the novel bed include drying, in particular where materials are initially too wet to be processed in a conventional fluidised bed and require a pre-treatment stage. Tests on the bed are necessary to identify quantitatively the maximum liquid loading which the bed can handle, either in batch or continuous processing.

5.5 Industrial Applications
A comparison between the conventional process and a gas-fluidised-bed fermenter is given in appendix C. The comparison indicates that the process represents a process intensification, and there is a significant reduction in the size of process plant required. Therefore there is a possibility for cost savings in the capital cost of a fermentation process plant with this technology. There may also be revenue savings, these are dependent on the reduced volumes of raw materials and effluent which need to be handled.
APPENDICES
APPENDIX A

Experimental Results from Steady State Heat Transfer Experiments
Appendix A

RUN 12

Measured Temperatures: steady state

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Calculated Values

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#### RUN 13

**Measured Temperatures: steady state**

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Appendix A

RUN 14

Measured Temperatures: steady state

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Appendix A

RUN 16

Measured Temperatures: steady state

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APPENDIX B

Estimation of the Magnitude of the Experimental Errors in the Heat Transfer Experiments
Air Flowrates
Orifice Plates
The orifice plates were calibrated in situ as the pipe runs were not long enough to allow extrapolation from the design calculations. The upstream orifice plate in the three inch pipe was calibrated against a rotameter, open to the atmosphere, which was fixed to the end of the supply pipe before the bed and downstream cyclone were connected. The orifice plate in the two inch pipe which measures the total flowrate to the bed, and the orifice plate in the one inch section which measures the amount of air which is being diverted through the jets at any time, were calibrated against the three inch orifice plate when the rest of the system was connected.

The correlation for each orifice plate can be expressed as follows:

\[ \text{flow at pressure } P_1 = x \cdot y \]

where \( x \) is an experimentally determined coefficient and \( y \) is the theoretical coefficient given by:

\[
y = \left( 1 - \beta \frac{\Delta P}{P_1} \right) \sqrt{\Delta P \left( \frac{P + P_1}{P} \right)}
\]

In which \( \beta \) is a function of the diameter ratio of the orifice and \( P \) is the barometric pressure.

For the equipment used:

- 3" OP \( x = 32.68 \) standard deviation = 1.70
- 2" OP \( x = 47.90 \) standard deviation = 1.25
- 1" OP \( x = 14.57 \) standard deviation = 0.52

determined by linear regression of the flow plotted against \( y \).
Appendix B

Water Flowrate

The water flowrate was measured using a rotameter; this could be read to +/- 0.5 divisions. For the heat transfer experiments high flowrates were used and the total error was of the order of 29 +/- 0.5 divisions or 2.2 +/- 0.04 l/min.

Temperatures

Thermocouples

The thermocouples on the rig were all type K thermocouples. These were calibrated against an ice point to measure any offsets and checked against water at 80 °C to test the consistency of the readings. The characteristics of each thermocouple are given below.

The offset of the infra-red probe was estimated by comparison with the thermocouples.

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</tr>
<tr>
<td>4</td>
<td>bare wire</td>
<td>-2.0 air outlet</td>
</tr>
<tr>
<td>5</td>
<td>3mm stainless sheath</td>
<td>-2.7 water bath</td>
</tr>
<tr>
<td>6</td>
<td>bare wire</td>
<td>0.5 inner surface of bed (top)</td>
</tr>
<tr>
<td>7</td>
<td>bare wire</td>
<td>1.25 inner surface of bed (2nd bottom)</td>
</tr>
<tr>
<td>8</td>
<td>bare wire</td>
<td>1.25 inner surface of bed (2nd top)</td>
</tr>
<tr>
<td>9</td>
<td>bare wire</td>
<td>-0.30 inner surface of bed (bottom)</td>
</tr>
<tr>
<td>10</td>
<td>3mm stainless sheath</td>
<td>0.0 ambient</td>
</tr>
<tr>
<td>11</td>
<td>IR temperature probe</td>
<td>-1.3 surface of particles in bed</td>
</tr>
</tbody>
</table>

The type K thermocouples used had a given accuracy of +/- 3 K in 1000 or 0.3%. The resolution of the electronics in the data logger was +/- 0.24 K. Therefore the accuracy of the temperature readings is of the order of +/- 0.5 K. To eliminate any extraneous noise from the Amstrad computer used for the data
logging, an average of ten readings was taken for each temperature measurement.
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C.1 Introduction
This short chapter contains a brief summary of the data relevant to an industrial application of this type of fluidised bed. It summarises the correlations used to model the bed behaviour, and the ranges within which the parameters have been tested. A rough assessment has also been made of this technology in comparison to submerged culture, concentrating on the culture parameters, and sample costs are provided for the main plant items in two alternate processing trains.

C.2 Possible Commercial Applications
This technology could be applied to a range of solids handling processes. As well as the wide variety of possible fermentation applications discussed in chapter 1, it also has potential for drying applications, particularly where the particles are initially too cohesive to be handled by conventional fluidised beds. Other possible uses include gas-phase polymerisation, particle coating processes and food processing applications.

C.3 Operating Parameters
The design has been tested on a range of different particles and conditions. The main design parameters are listed below, with the range of values studied in the trials. In most cases the process is not limited to the trial range and the given values do not imply that the bed could not be operated successfully under different conditions; where possible wider limits or a method for scale-up are suggested.

Particle Size
The range tested was 5 to 15 mm diameter. Under Geldart's classification the test particles were all either type B or type D. This classification is considered to be the best method for assessing whether particles can be mobilised in this system; i.e. it is likely that smaller, heavier particles which also fall into group B or D could also be mobilised. Other particles are not excluded, but they have not been tested.

Particle Morphology
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Spherical, cylindrical, ellipsoidal and cubical particles were used in trials. All were successfully mobilised, though more angular particles required a higher velocity to initiate motion. Solid and porous particles were both successfully mobilised.

Moisture Content
Porous foam particles were successfully mobilised at a water content of 200g water to 60g dry weight of particles (approximately 2 litres bulk volume). At this loading most of the water is held within the particles rather than on the surface.

Foam particles have also been mobilised after coating with an equal volume of nutrient 'slurry'. This material had a custard-like consistency and the particles were cohesive due to a surface layer of liquid.

The total amount of water is less significant than the water held on the particle surface.

Vessel Volume
The two experimental units were 150 mm and 225 mm in diameter. Beds of particles were successfully mobilised at height to diameter ratios of 0.2 to 2 in the smaller bed and 0.2 to 1.5 in the larger bed. Observations indicate that with the jets at the base the vigorously rotating region may be limited to one bed diameter above the distributor; however the bed above this height will also rotate, though more slowly, and a zone of conventional bubbling fluidisation may form above one bed diameter above the distributor.

Jet Position
To be effective the jets must enter the bed at the base where they will prevent the particles settling on the distributor. In these trials four wall jets 15 mm above the distributor were used.

The jets enter the bed at an angle to the wall to promote rotation in the bed, and for the trials each jet was directed towards the entry point of the next. This is considered to be optimal for the
initiation of rotation. For larger beds more jets could be used, and the number required could be estimated by using the correlation for the jet penetration length to calculate a maximum practical inter-jet distance.

In larger diameter beds the particles may settle out towards the centre of the distributor if a central spout is not used.

Directional perforated plate is available from manufacturers such as Conidur Ltd., and this could be fabricated to promote rotation over the entire distributor surface.

Heat Load

Heat transfer coefficients between the bed and the wall are dependent on particle size as discussed in chapter 4. The coefficient increases as the particle size is reduced. Measured gas to wall heat transfer coefficients for 8 mm particles were between 80 and 100 W m\(^{-2}\)K\(^{-1}\).

Heat transfer by evaporation is limited by the humidity of the fluidising air but can provide a significant additional cooling effect in drying or fermentation applications.

Heat Transfer Surfaces

In the rotating region the only suitable heat transfer surfaces are concentric with the bed - either the external bed wall or a concentric cylinder. If a deeper bed is used, other surfaces such as cold fingers, or panels, with no horizontal surfaces, could be used in the upper, fluidised region of the bed.

C.4 Design Correlation Summary

The correlations which have been used through the thesis to model aspects of the spinning fluidised bed are summarised below.

Minimum Fluidisation Velocity (see section 2.2.3)

*Correlation of Wen and Yu*
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\[
U_{mr} = \frac{\mu}{\rho_f d_v} \left( 1135.7 + 0.0408 \cdot Ar \right)^{\frac{1}{2}} - 33.7
\]

\[
Ar = \frac{\rho_f d_v^3 \left( \rho_p - \rho_f \right) g}{\mu_f^2}
\]

Minimum Mobilisation Point (see sections 3.5.5 and 3.5.6)
Mobilisation occurs when the torque exerted by the jets equals the frictional torque on the sides and base of the bed.

Torque Exerted by the Jets

\[
T_j = \frac{Q^2 \rho_f}{N A} y
\]

Where the jets are at the bottom of the bed, the correlations for the torque due to friction have the following form:

Torque on the base

\[
T_b = \frac{\pi D^3}{12} \mu_b \left( \frac{\gamma_{mod} D}{4 \mu_w K \sin \alpha} \left[ 1 - \exp \left( \frac{-4 \mu_w K \sin \alpha}{D} \right) \right] \right)
\]

Torque on the wall

\[
T_{wa} = \frac{\pi D^3}{8 \tan \alpha} \gamma_{mod} \left( H - \frac{1}{R} \left[ 1 - \exp(-RH) \right] \right)
\]

where

\[
R = \frac{4 \mu_w K \sin \alpha}{D}
\]

\[
\gamma_{mod} = \gamma - \frac{\Delta P}{H}
\]

The bed will rotate when the torque exerted by the jets exceeds the frictional torque.

Details of the more complex versions of these correlations, for the case where the jets are part-way up the wall, are discussed in sections 3.5.5 and 3.5.6.
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Jet Penetration Length (section 3.9)

Correlation by Merry

\[
\frac{L}{d_o} + 4.5 = 5.25 \left( \frac{\rho_f u_o^2}{(1-\varepsilon) \rho_p g d_p} \right)^{0.4} \left( \frac{\rho_f}{\rho_p} \right)^{0.2} \left( \frac{d_p}{d_o} \right)^{0.2}
\]

Other correlations by Shakhova and Zenz, given in section 3.9 may be appropriate for other particle sizes.

Heat Transfer Coefficients (section 4.2)

Correlation by Baskakov and Suprun

\[
Nu_{gc} = 0.0175 \, Ar^{0.46} \, Pr^{0.33} \quad U > U_m
\]

\[
Nu_{gc} = 0.0175 \, Ar^{0.46} \, Pr^{0.33} \left( \frac{U}{U_m} \right)^{0.3} \quad U_m < U < U_m
\]

where \( U_m \) is the superficial gas velocity giving maximum bed to surface heat transfer (from Todes 1965).

Correlation by Botterill and Denloye

\[
\frac{h \, d_p^{0.5}}{k_f} = 0.86 \, Ar^{0.39} \quad \text{for } 10^3 < Ar < 2 \times 10^6
\]

(dimensions are \( m^{-0.5} \))

Correlation by LaNauze for \( h_{gp} \)

\[
Nu = 2c + 0.69 \left( \frac{1}{\varepsilon} \right)^{1/2} Re^{1/2} Pr^{1/3}
\]

where the Prandtl number is given by:

\[
Pr = \frac{k_f}{C_p \mu_f}
\]

Correlation by Zabrodsky for \( h_{gp} \)

\[
h_{gp} = 35.8 \rho_p^{0.2} k_f^{0.6} d_p^{-0.36}
\]
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Correlation by Xavier and Davison for \( h_{pc} \)

\[
h_{pc} = \frac{1}{(1/h_p) + (1/h_f)}
\]

\[
h_f = \frac{m k_f}{d_p}
\]

\[
h_p = \int_0^t h_1 \frac{dt}{t_r} = 2 \left[ \frac{k_{mf} \rho_{mf} C_{mf}}{\pi t_r} \right]^{1/2}
\]

Table C.1 Summary of Values of Heat Transfer Coefficients Derived from Published Correlations

<table>
<thead>
<tr>
<th>Coefficient</th>
<th>( W \text{ m}^{-2} \text{K}^{-1} )</th>
<th>Reference</th>
</tr>
</thead>
<tbody>
<tr>
<td>( h_{gc} )</td>
<td>49.4</td>
<td>Baskakov and Suprun (1972)</td>
</tr>
<tr>
<td></td>
<td>73.3</td>
<td>Botterill and Denloye (1978)</td>
</tr>
<tr>
<td>( h_{pc} )</td>
<td>16.8</td>
<td>Xavier and Davidson (1985)</td>
</tr>
<tr>
<td>( h_{gp} )</td>
<td>59.7</td>
<td>LaNauze (1983)</td>
</tr>
<tr>
<td></td>
<td>53.2</td>
<td>Zabrodsky (1966)</td>
</tr>
</tbody>
</table>

C.5 Example Fermentation: Penicillin Production Process
The penicillin production process has been used to make a comparison between the conventional submerged process and solid-state fermentation using a spinning bed fermenter. The new process provides opportunities for process intensification.

A flowsheet for a typical plant is shown in Figure C.1. Typical process parameters are shown in table C.2.

Table C.2 Process Parameters For Penicillin Production

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>temperature</td>
<td>23-28 °C</td>
</tr>
<tr>
<td>oxygen usage</td>
<td>0.4-0.8 mmol l(^{-1}) min(^{-1})</td>
</tr>
<tr>
<td>aeration</td>
<td>0.5-1.0 m(^3)m(^{-3}) min(^{-1})</td>
</tr>
<tr>
<td>power input (including aeration)</td>
<td>1-4 W l(^{-1})</td>
</tr>
<tr>
<td>heat generation</td>
<td>124 kcal (mol O(_2))(^{-1})</td>
</tr>
<tr>
<td>yield of penicillin (actual)</td>
<td>0.03-0.05 g (g glucose)(^{-1})</td>
</tr>
</tbody>
</table>
Appendix C: Implications for an Industrial Application

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>max product concentration</td>
<td>10-30 g l(^{-1})</td>
</tr>
<tr>
<td>max biomass concentration</td>
<td>2.5 g l(^{-1})</td>
</tr>
<tr>
<td>batch size</td>
<td>40-200 m(^3)</td>
</tr>
<tr>
<td>batch cycle time</td>
<td>8 days</td>
</tr>
</tbody>
</table>

The spinning bed process represents a process intensification. The biomass concentration is increased, from 2.5 g/l, to as much as 10-40 g/l by the addition of supports; this provides at least a four-fold reduction in the fermenter volume. The hold-up volume of broth is further reduced by up to 40% due to the increased voidage. Broth will have to be added to provide extra nutrients and to make up for evaporation losses during the fermentation, however it can be assumed that some of the nutrients can be supplied at higher concentrations and therefore, overall the broth preparation and sterilisation equipment can be reduced in size by at least a factor of two.

Revised process parameters are shown in table C.3.

Table C.3 Revised Process Parameters for Penicillin Production Using Fluidised Bed Solid State Culture

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>maximum biomass concentration</td>
<td>20 g/l</td>
</tr>
<tr>
<td>productivity</td>
<td>40 g/l after 2 days</td>
</tr>
<tr>
<td>specific productivity</td>
<td>1g penicillin/g biomass/day</td>
</tr>
<tr>
<td>oxygen consumption</td>
<td>4.8 mmol/l reactor volume/min</td>
</tr>
<tr>
<td>heat output</td>
<td>519 kJ/mol O(_2)</td>
</tr>
<tr>
<td></td>
<td>250 kW</td>
</tr>
<tr>
<td>superficial velocity</td>
<td>0.4 m/s</td>
</tr>
<tr>
<td></td>
<td>2.5 (U_{mf} = 1 \text{ m s}^{-1})</td>
</tr>
</tbody>
</table>

C.6 Economic Implications and Relative Costs

A number of comprehensive texts are available for fermentation plant design. The calculations required to develop and cost a fermentation process, from inception to industrial scale plant, are discussed by Atkinson (1983), and by Bailey and Ollis (1986). It is apparent that the economics of any specific process are strongly dependent on the raw materials required and the value and
purity of the end product. The main cost may lie in the capital cost of plant, media costs, extraction process running costs or as in many other areas of effect chemicals, with the distribution and marketing costs of the final product.

This section discusses the main areas in the process where a spinning bed process fermentation differs from submerged culture, with estimates where possible of the relative costs. Penicillin production has been used as the model process, as it represents an example of large-scale fermentation with a complex recovery train, and data on the process and costs are readily available. This is a high value product; the implications for the production of low-value products are discussed at the end of the section.

Key Areas
The fermenter design will have implications for the size and cost of the fermenter. It will also affect the volume of broth required, which will affect the preparation and sterilisation facilities, and the running costs. As was stated above, a conservative estimate is that the broth preparation and sterilisation facilities can be halved. The largest impact on the process when replacing a submerged culture with a spinning bed process will be seen in the downstream processing stage, as the initial concentration steps can be reduced or eliminated. The largest impact on the costs will be the reduction in land area and civil costs obtained by reducing the size of plant required.

Downstream Processing
A typical extraction train is shown in Figure C.2. This shows filtration and three extraction stages to produce a product rich broth, followed by evaporation and washing stages. Each extraction stage results in a 90% reduction in the product stream volume.

For the solid substrate product, the filtration stage is replaced by centrifugation, which yields a concentrated product stream. Only one or possibly two extraction stages are then required before the product is sufficiently concentrated for evaporation.
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and crystallisation. This is a significant saving as it is the high volume stages which are eliminated.

Sample Costings

For this example a unit producing 100 kg/day of penicillin (before recovery) has been used. In a submerged culture a 40 m³ vessel is required, with a batch time of 8 days. Based on the figures given above the following data are available:

<table>
<thead>
<tr>
<th>Maximum biomass concentration</th>
<th>2.5 g/l</th>
</tr>
</thead>
<tbody>
<tr>
<td>Productivity</td>
<td>20 g/l after 8 days</td>
</tr>
<tr>
<td>Specific productivity</td>
<td>1 g penicillin/g biomass/day</td>
</tr>
<tr>
<td>Oxygen consumption</td>
<td>0.6 mmol/l/min</td>
</tr>
<tr>
<td></td>
<td>0.4 moles s⁻¹</td>
</tr>
<tr>
<td>Heat output</td>
<td>519 kJ/mol O₂</td>
</tr>
<tr>
<td></td>
<td>207 kW</td>
</tr>
<tr>
<td>Aeration</td>
<td>0.75 m³ m⁻³ min⁻¹</td>
</tr>
<tr>
<td></td>
<td>30 m³ min⁻¹</td>
</tr>
</tbody>
</table>

Using the same specific productivity we can make the following assumptions for the spinning bed system:

<table>
<thead>
<tr>
<th>Reactor yield</th>
<th>100 kg/day</th>
</tr>
</thead>
<tbody>
<tr>
<td>Specific productivity</td>
<td>1 g penicillin/g biomass/day</td>
</tr>
<tr>
<td>Biomass concentration</td>
<td>20 g/l reactor volume</td>
</tr>
<tr>
<td>Batch time</td>
<td>2 days</td>
</tr>
<tr>
<td>Maximum penicillin concentration</td>
<td>40 g/l reactor volume</td>
</tr>
<tr>
<td>Reactor voidage</td>
<td>0.4</td>
</tr>
<tr>
<td>Reactor working volume</td>
<td>5 m³</td>
</tr>
<tr>
<td>Total reactor volume (with headspace)</td>
<td>6 m³</td>
</tr>
<tr>
<td>Oxygen consumption</td>
<td>0.24 mmol/g biomass/min</td>
</tr>
<tr>
<td></td>
<td>0.4 moles s⁻¹</td>
</tr>
<tr>
<td>Heat output</td>
<td>519 kJ/mol O₂</td>
</tr>
<tr>
<td></td>
<td>207 kW</td>
</tr>
<tr>
<td>Aeration</td>
<td></td>
</tr>
<tr>
<td>Vessel dimensions</td>
<td></td>
</tr>
<tr>
<td>Diameter</td>
<td>2 m</td>
</tr>
<tr>
<td>Height</td>
<td>2 m</td>
</tr>
<tr>
<td>Superficial velocity</td>
<td>assume Uₘᶠ</td>
</tr>
<tr>
<td></td>
<td>0.4 m/s</td>
</tr>
</tbody>
</table>

252
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use $2.5 \text{ U} = 1 \text{ m s}^{-1}$
flow $3.1 \text{ m}^3 \text{s}^{-1} = 180 \text{ m}^3 \text{min}^{-1}$
specific flow $= 30 \text{ m}^3 \text{m}^{-3} \text{min}^{-1}$

Due to the batch nature of the reaction, a typical production unit has several fermenters serviced by a common extraction train. The costings are therefore based on a train to process the product from four reactors.

Plant Item Capacities: Submerged Culture

<table>
<thead>
<tr>
<th>Plant Item</th>
<th>Per Unit</th>
<th>Overall</th>
</tr>
</thead>
<tbody>
<tr>
<td>fermenter</td>
<td>$40 \text{ m}^3$</td>
<td>$40 \text{ m}^3 \times 4$</td>
</tr>
<tr>
<td>air compression</td>
<td>$30 \text{ m}^3 \text{min}^{-1} ; 3 \text{ barg}$</td>
<td>$120 \text{ m}^3 \text{min}^{-1} ; 3 \text{ barg}$</td>
</tr>
<tr>
<td>rotary Filter</td>
<td>$5 \text{ m}^3 \text{ day}^{-1}$</td>
<td>$20 \text{ m}^3 \text{ day}^{-1}$</td>
</tr>
<tr>
<td>extractor</td>
<td>$5 \text{ m}^3 \text{ day}^{-1}$</td>
<td>$20 \text{ m}^3 \text{ day}^{-1}$</td>
</tr>
<tr>
<td>ion exchange column</td>
<td>$1250 \text{ l day}^{-1}$</td>
<td>$5000 \text{ l day}^{-1}$</td>
</tr>
<tr>
<td>extractor</td>
<td>$1250 \text{ l day}^{-1}$</td>
<td>$5000 \text{ l day}^{-1}$</td>
</tr>
<tr>
<td>evaporator</td>
<td>$250 \text{ l day}^{-1}$</td>
<td>$1000 \text{ l day}^{-1}$</td>
</tr>
</tbody>
</table>

Plant Item Capacities: Spinning Bed

<table>
<thead>
<tr>
<th>Plant Item</th>
<th>Per Unit</th>
<th>Overall</th>
</tr>
</thead>
<tbody>
<tr>
<td>fermenter</td>
<td>$6 \text{ m}^3$</td>
<td>$6 \text{ m}^3 \times 4$</td>
</tr>
<tr>
<td>air compression</td>
<td>$180 \text{ m}^3 \text{min}^{-1} ; 3 \text{ barg}$</td>
<td>$720 \text{ m}^3 \text{min}^{-1} ; 3 \text{ barg}$</td>
</tr>
<tr>
<td>centrifuge</td>
<td>$2.5 \text{ m}^3 \text{day}^{-1}$</td>
<td>$10 \text{ m}^3 \text{day}^{-1}$</td>
</tr>
<tr>
<td>extractor</td>
<td>$250 \text{ l day}^{-1}$</td>
<td>$1000 \text{ l day}^{-1}$</td>
</tr>
<tr>
<td>ion exchange column</td>
<td>$50 \text{ l day}^{-1}$</td>
<td>$200 \text{ l day}^{-1}$</td>
</tr>
<tr>
<td>evaporator</td>
<td>$50 \text{ l day}^{-1}$</td>
<td>$200 \text{ l day}^{-1}$</td>
</tr>
</tbody>
</table>
Main Plant Item Costs

1. **FERMENTER**
   stainless steel, jacketed, with agitator
   40 m$^3$ x 4
   812

2. **COMPRESSOR**
   filtered air sterile conditions
   7200 m$^3$ hr$^{-1}$; 3 barg
   197

3. **ROTARY VACUUM FILTER**
   stainless steel
   20 m$^3$ day$^{-1}$
   150

4. **CENTRIFUGAL EXTRACTOR**
   stainless steel, sterile conditions
   20 m$^3$ day$^{-1}$
   147

5. **ION EXCHANGE COLUMN**
   stainless steel; sterile conditions
   5000 l day$^{-1}$
   30

6. **CENTRIFUGAL EXTRACTOR**
   stainless steel, sterile conditions
   5000 l day$^{-1}$
   79

7. **CENTRIFUGAL EXTRACTOR**
   stainless steel, sterile conditions
   1000 l day$^{-1}$
   73

8. **SOLVENT EVAPORATOR**
   stainless steel
   200 l day$^{-1}$
   10

**TOTAL**
1498

*Costs taken from ICI Newcost database indexed to April 1992*
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Figure C.4 Flowsheet for Spinning Bed Process

Main Plant Item Costs

<table>
<thead>
<tr>
<th>Item Description</th>
<th>Cost (£ 000)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1. FERMENTER stainless steel, jacketted, with agitator</td>
<td>308</td>
</tr>
<tr>
<td>2. COMPRESSOR filtered air sterile conditions</td>
<td>1305</td>
</tr>
<tr>
<td>3. CENTRIFUGAL EXTRACTOR stainless steel, sterile conditions</td>
<td>196</td>
</tr>
<tr>
<td>4. CENTRIFUGAL EXTRACTOR stainless steel, sterile conditions</td>
<td>73</td>
</tr>
<tr>
<td>5. ION EXCHANGE COLUMN stainless steel; sterile conditions</td>
<td>10</td>
</tr>
<tr>
<td>6. SOLVENT EVAPORATOR stainless steel</td>
<td>5</td>
</tr>
</tbody>
</table>

**TOTAL** 1897

Discussion

The main plant item costs are higher for the spinning bed process.
than for the submerged culture. This is mainly due to the increased compression requirements. The higher MPI cost is likely to be offset by reduced land and civil costs as most of the plant items are smaller. Significant savings may also be available in the broth preparation area.

In this example, based on penicillin, the concentrations and final yield calculated for solid-state culture are probably higher than could in effect be achieved. Penicillin production and penicillin strains have been optimised in submerged culture over a period of many years, and there is less potential for improvement than most other fermentation processes. However, for the same reason, more process data is available and this is why it has been used as the basis for this general example. The intensification possible with solid-state culture has been estimated based on data in the books by Bailey and Ollis (1986) and Atkinson (1983), and for specific systems, as discussed in chapter 1, higher relative productivities may be achieved.

In any process where final concentrations in the broth could be increased using this process, a larger reduction in cost could be made by reducing the size of the extraction train required.

Low Value Products
Concentration and drying are major proportion of the cost for low-value fermentation products. Depending on the choice of support, some solid products could be used straight from the reactor, by adding a drying cycle to the end of the fermentation stage; e.g. yeast or mycoprotein. Conversely, a liquid product used in the form of a crude extract, could be sufficiently concentrated after a simple centrifugal separation of the solids to be used directly.

C.7 Experiments Required for Process Assessment and Scale-Up
The process has not been scaled up from the bench-scale units used for the initial experimentation. For a drying application, the required tests are straight-forward: confirmation that the correlations can be used to design a larger unit, and measurements
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of the drying rate and effective heat transfer rate at the desired scale or at a suitable intermediate pilot scale.

Fermentation Applications
The following series of tests is not exhaustive but indicates the main areas which require experimentation before scale-up of a fermentation process. These experiments are independent of the tests required to scale up the operation of the bed to larger units. The availability of a working unit of the desired size is assumed.

Laboratory Scale
(a) Measurements of the specific growth rate, heat evolution rate and nutrient and aeration requirements. These could be done in either submerged or solid state culture.

(b) Shake flask experiments to match the organism to a suitable support medium (and particle size). These could be carried out with irrigation with nutrient media if required. The culture should also be tested for resistance to dessication.

(c) Measurements of product generation rate and final yield in solid-state culture (shake flasks). Optimisation of the growth conditions and culture period.

(d) Development of an effective inoculation method. This may be done from suspended culture or by spore suspension. Good dispersion onto the supports is critical.

(e) Determine required scale of production for economic process.

Pilot Scale
(f) Pilot scale tests (200mm diameter bed) to confirm that the culture grows as expected, producing the desired products and is not subject to abrasion or agglomeration, or to undue dessication, cell damage or cell death.

(g) Further scale-up: depending on the required final fermenter
Appendix C: Implications for an Industrial Application

size, more pilot-scale tests may be required.

C.8 Conclusions
This system has potential for a number of commercial applications. A preliminary estimate of the costs of full scale production indicates that the capital cost is similar and may be less than for a conventional process. Running costs are more process dependent.
Appendix C: Implications for an Industrial Application

Figure C.1 Flowsheet for Conventional Submerged Culture Plant
(from Atkinson 1983)

[Diagram of flowsheet for conventional submerged culture plant]

Laboratory inoculum

Raw material storage

Medium preparation

Continuous sterilizer

Evaporator (double effect with recompression)

Concentrate-holding tank

Spray dryer

Holding bin

Formulation materials

Blender

Holding bin x 2

Bagger

Packaging and shipping

Seed tank x 3

Production fermenter x 4

Culture-holding tank

Seeds

Wash water

Condenser water

Steam

Process water

Power

Sterile air

Exhaust air

Process water

Steam

Tower water

Sterile air

Process water

Power

Steam

Tower water

Sterile air

Process water

Power

Steam

Tower water

Sterile air

Process water

Power

Steam

Tower water

Sterile air

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Appendix C: Implications for an Industrial Application

Figure C.2 Penicillin Downstream Extraction Train
(from Atkinson 1983)
## Nomenclature

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Definition</th>
</tr>
</thead>
<tbody>
<tr>
<td>$a$</td>
<td>constant in integration $= 4 \mu_w K \sin \alpha / (\gamma D)$</td>
</tr>
<tr>
<td>$a_{\text{wall}}$</td>
<td>wall area</td>
</tr>
<tr>
<td>$A$</td>
<td>cross-sectional area of jet</td>
</tr>
<tr>
<td>$A_B$</td>
<td>cross-sectional area of bed</td>
</tr>
<tr>
<td>$Ar$</td>
<td>Archimedes number</td>
</tr>
<tr>
<td>$Bi$</td>
<td>Biot number</td>
</tr>
<tr>
<td>$C_f$</td>
<td>specific heat of gas</td>
</tr>
<tr>
<td>$C_{mf}$</td>
<td>specific heat of bed at minimum fluidisation</td>
</tr>
<tr>
<td>$C_p$</td>
<td>specific heat of particle</td>
</tr>
<tr>
<td>$d_o$</td>
<td>orifice diameter</td>
</tr>
<tr>
<td>$d_p$</td>
<td>particle diameter</td>
</tr>
<tr>
<td>$d_v$</td>
<td>diameter of volume equivalent sphere</td>
</tr>
<tr>
<td>$D$</td>
<td>bed diameter</td>
</tr>
<tr>
<td>$D_D$</td>
<td>coefficient of mass diffusion in gas</td>
</tr>
<tr>
<td>$D_p$</td>
<td>coefficient of mass diffusion in particle</td>
</tr>
<tr>
<td>$ff_c$</td>
<td>flowability factor</td>
</tr>
<tr>
<td>$F_H$</td>
<td>adhesion force</td>
</tr>
<tr>
<td>$g$</td>
<td>acceleration due to gravity</td>
</tr>
<tr>
<td>$h_{\text{ext}}$</td>
<td>heat transfer coefficient external to particle</td>
</tr>
<tr>
<td>$h_f$</td>
<td>gas mediated heat transfer coefficient between particle &quot;packet&quot; and wall</td>
</tr>
<tr>
<td>$h_i$</td>
<td>instantaneous heat transfer coefficient</td>
</tr>
<tr>
<td>$h_{gw}$</td>
<td>overall heat transfer coefficient between wall and gas</td>
</tr>
<tr>
<td>$h_{gc}$</td>
<td>gas convective heat transfer coefficient</td>
</tr>
<tr>
<td>$h_{gp}$</td>
<td>gas to particle heat transfer coefficient</td>
</tr>
<tr>
<td>$h_{\text{max}}$</td>
<td>maximum bed to wall heat transfer coefficient</td>
</tr>
<tr>
<td>$h_p$</td>
<td>mean heat transfer coefficient between particle and wall</td>
</tr>
<tr>
<td>$h_{pc}$</td>
<td>particle convective heat transfer coefficient</td>
</tr>
<tr>
<td>$h_r$</td>
<td>radiative heat transfer coefficient</td>
</tr>
<tr>
<td>$h_l$</td>
<td>height of jet inlets above distributor</td>
</tr>
<tr>
<td>$H$</td>
<td>bed depth</td>
</tr>
<tr>
<td>$H/D$</td>
<td>aspect ratio of fluidised bed</td>
</tr>
<tr>
<td>$k_b$</td>
<td>thermal conductivity of bed</td>
</tr>
<tr>
<td>$k_r$</td>
<td>thermal conductivity of gas</td>
</tr>
</tbody>
</table>
Nomenclature

\( k_{\text{mf}} \) thermal conductivity of bed at minimum fluidisation

\( k^o \) thermal conductivity of bed independent of fluid flow

\( k_p \) thermal conductivity of particle

\( K \) constant (Molerus)

\( K \) constant in Janssen's equation

\( K_A \) constant for active solution of Janssen's equation

\( K_P \) constant for passive solution of Janssen's equation

\( L \) jet penetration length

\( L_{\text{max}} \) maximum jet penetration length

\( m \) constant in particle to wall heat transfer

\( n \) power law constant

\( N \) number of jets

\( \text{Nu}_{\text{gc}} \) Nusselt number for gas convective heat transfer coefficient

\( \Delta P \) pressure drop

\( Pr \) Prandtl number

\( Q \) volumetric gas flowrate

\( Q_D \) volumetric gas flowrate through the distributor

\( Q_j \) volumetric gas flowrate through the jets

\( Q_{\text{min}} \) minimum volumetric gas flowrate through jets for mobilisation

\( Q_0 \) surcharge weight on bed, at surface

\( Q_{\text{gas}} \) heat removed in gas stream

\( r \) radial position in bed

\( R \) bed radius

\( R = \frac{4 \mu w K \sin \alpha}{D} \) = constant in Janssen's equation

\( Re \) Reynolds' number

\( Re_{\text{min}} \) Reynolds' number at minimum fluidising velocity

\( Re_{\text{opt}} \) Reynolds' number at optimum velocity for heat transfer to wall

\( Sc \) Schmidt number

\( Sh \) Sherwood number

\( t \) elapsed time
Nomenclature

\( t_r \) particle residence time at the wall
\( T_{g,\text{in}} \) temperature of gas entering bed
\( T_{g,\text{out}} \) temperature of gas leaving bed
\( T_w \) temperature at wall
\( T_o \) uniform bed temperature outside thermal boundary layer
\( \Delta T_{\text{gas}} \) change in gas temperature across the bed
\( \Delta T_{gw,\text{mean}} \) mean bed to wall temperature difference
\( T \) torque
\( T_b \) frictional torque on base
\( T_{wa} \) frictional torque on wall above jets
\( T_{wb} \) frictional torque on wall below jets
\( u \) gas velocity
\( u_b \) bubble velocity
\( u_o \) jet velocity
\( U \) superficial velocity in bed
\( U_m \) velocity of maximum bed to surface heat transfer (Todes)
\( U_{mf} \) minimum fluidisation velocity
\( U_{ms} \) minimum spouting velocity
\( v_z \) vertical velocity of particles at wall
\( W \) bed weight
\( x \) interjet distance
\( y \) distance between point of action of jet force and bed axis
\( z \) vertical position in bed, measured from the bed surface
\( z_l \) vertical position in bed, measured down from the jet inlets

\( D_i \) diameter of inlet (spouted bed)
\( \Xi \) Henry's Law coefficient
\( k \) mass transfer coefficient
<table>
<thead>
<tr>
<th>Symbol</th>
<th>Definition</th>
</tr>
</thead>
<tbody>
<tr>
<td>$\alpha$</td>
<td>angle to the horizontal of net frictional force at the wall</td>
</tr>
<tr>
<td>$\alpha_b$</td>
<td>thermal diffusivity of bed</td>
</tr>
<tr>
<td>$\alpha_d$</td>
<td>angle of jet dispersion</td>
</tr>
<tr>
<td>$\beta$</td>
<td>angle used to measure coefficients of friction</td>
</tr>
<tr>
<td>$c$</td>
<td>bed voidage</td>
</tr>
<tr>
<td>$c_m$</td>
<td>packed bed voidage</td>
</tr>
<tr>
<td>$c_{mf}$</td>
<td>bed voidage at minimum fluidisation</td>
</tr>
<tr>
<td>$\gamma$</td>
<td>specific bed weight ($= \rho_b g$)</td>
</tr>
<tr>
<td>$\gamma_{mod}$</td>
<td>effective specific bed weight accounting for drag forces</td>
</tr>
<tr>
<td>$\tau$</td>
<td>shear stress</td>
</tr>
<tr>
<td>$\tau_0$</td>
<td>adhesive force</td>
</tr>
<tr>
<td>$\tau_f$</td>
<td>frictional stress at wall</td>
</tr>
<tr>
<td>$\tau_{rz}$</td>
<td>vertical shear stress component</td>
</tr>
<tr>
<td>$\tau_{r\theta}$</td>
<td>horizontal shear stress component</td>
</tr>
<tr>
<td>$\phi$</td>
<td>angle of internal friction of bed material</td>
</tr>
<tr>
<td>$\rho_b$</td>
<td>bed bulk density</td>
</tr>
<tr>
<td>$\rho_{mf}$</td>
<td>bed bulk density at minimum fluidisation</td>
</tr>
<tr>
<td>$\rho_f$</td>
<td>fluid density</td>
</tr>
<tr>
<td>$\rho_p$</td>
<td>particle density</td>
</tr>
<tr>
<td>$\mu$</td>
<td>coefficient of friction</td>
</tr>
<tr>
<td>$\mu_b$</td>
<td>coefficient of friction for material of base</td>
</tr>
<tr>
<td>$\mu_w$</td>
<td>coefficient of friction for wall material</td>
</tr>
<tr>
<td>$\mu_f$</td>
<td>fluid viscosity</td>
</tr>
<tr>
<td>$s$</td>
<td>sphericity factor</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>normal stress</td>
</tr>
<tr>
<td>$\sigma_{rr}$</td>
<td>normal stress on wall</td>
</tr>
<tr>
<td>$\sigma_{zz}$</td>
<td>normal stress on base</td>
</tr>
</tbody>
</table>
Nomenclature

Dimensionless constants

Archimedes number:

\[ Ar = \frac{\rho_f d^3_v \left( \rho_p - \rho_f \right) g}{\mu_f^2} \]

Biot number:

\[ Bi = \frac{h_{ext} d_p}{k_p} \]

Prandtl number:

\[ Pr = \frac{k_f}{C_p \mu_f} \]

Nusselt number:

\[ \text{Nu}_{gc} = \frac{h_{gc} d_v}{k_f} \]

Reynolds number:

\[ Re = \frac{\rho_f u d_p}{\mu_f} \]

Sherwood number:

\[ Sh = \frac{k_p d_p}{D} \]
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