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Part 2

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DESIGN OF LAMELLA SEPARATORS

PART II

by

D.J. BROWN, BSc, DIS

A doctoral thesis submitted in partial fulfilment of the requirements for the award of Doctor of Philosophy of Loughborough University of Technology

November 1986

Supervisor: Dr A.S. Ward
Director of Research: Dr B.W. Brooks

Department of Chemical Engineering

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ABSTRACT

A method has been developed to predict the minimum angle at which solids will flow down the lower inclined surface of a lamella separator. It is based on a frictional analysis and requires experimental data from a simple batch rig. This method should help to eliminate problems with sludge build-up in lamella separators. The minimum angle was dependent on particle size, shape and surface texture, liquid viscosity and surface tension and the type of settler material.

The degree of variation in the underflow solids concentration with time was also examined. It was found to be dependent on settler length, angle and a well-established flow regime. The mode of sludge flow and the mode of settler operation had no effect.

The continuous operation of a lamella separator using a real system was examined over a range of settler lengths, angles, plate spacings and operating modes. Two particulate systems were examined, limestone of size ranges 5-10 µm and 45-53 µm, both of which were dispersed in water. The requirements for steady state conditions and flow stability were met as closely as possible.

It was shown that the predictability of the Nakamura-Kuroda equation could be vastly improved by the careful selection both of the settler dimensions and in the case of the 5-10 µm limestone, of the operating mode. Eighty per cent agreement between this equation and the experimental results was obtained with 5-10 µm limestone at a settler length of 0.49m in the cocurrent supercritical mode. The most favourable conditions were short settler lengths, shallow angles and operation in the cocurrent supercritical mode. This mode was more stable than other modes and its efficiency was not greatly affected by change in settler variables. An optimum channel length to plate spacing ratio was found to exist but was dependent on channel spacing, particle size and mode of operation.
ACKNOWLEDGEMENTS

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- Mr R McTernan for his assistance in operating the continuous rig
- Mr G Boyden for photographic services
- My family and my boyfriend, Graham Heap, for their encouragement and support
- Janet Smith for typing the thesis.

DECLARATION OF ORIGINALITY

The experimental work reported herein is the responsibility of the author. The interpretation and theoretical arguments are likewise the author's except where acknowledgement is made of previously published work.
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CHAPTER 1
INTRODUCTION

A lamella separator is an enhanced rate gravity settler for clarifying and thickening suspensions. It consists of a tank containing a series of parallel inclined plates. The effect of the plates is to provide a large number of discrete settling chambers, with the result that the effective settling area can be more than ten times greater than the horizontally projected area of the separator. The advantages of a lamella separator over a conventional separator in addition to its compactness are lower capital, installation and operating costs and the potential for higher settling rates.

Most of the models developed to predict the behaviour of suspensions settling under the influence of inclined surfaces have either been too simplistic or have contained too any ad-hoc assumptions. Hence the present design methods are limited in their extent and accuracy and process engineers tend to rely heavily on empirical findings and past industrial experience. Extensive pilot plant experiments are often necessary which are both time consuming and costly. The more well tested design methods, reported in the literature, involve either imposing an improvement factor on the Coe and Clavenger procedure for conventional settlers or adding another term to the Yoshioka technique. The improvement factor is based on the Nakamura-Kuroda equation, developed to predict the rate of sedimentation in an inclined vessel. However the accuracy of these methods is only about 50%. More recently some new guidelines for the design scheme were proposed by Poh\textsuperscript{48} based on an improved understanding of the operation of the lamella settler. These included achieving laminar flow and steady state conditions, flow stability and a continuous sludge flow along the lower inclined surface. He showed that by adopting these guidelines the Nakamura-Kuroda equation was able to predict the maximum overflow in both batch and continuous operation over the following range of conditions:
\[ \Lambda = 0(10^4) - 0(10^7) \]
\[ \text{Re} = 0(1) - 0(10) \]

In addition an optimum plate length to plate spacing ratio was found to exist beyond which the design became uneconomic due to flow instabilities. Poh used an ideal system of glass beads dispersed in a mixture of Reofos 65/Reomol DBP in experiments to substantiate the theory.

An additional problem that has not been dealt with sufficiently in the literature concerns the flow of sludge down the lower inclined surface. If the settler angle is not steep enough then poor sludge flow and solids build-up can occur. The current design procedure is to overestimate the angle to ensure that this does not occur. However this may not give rise to the optimum design.

A detailed review of the literature available on lamella separators and on the design constraints proposed can be found in Chapters 2 and 3 respectively.

The present work aims to:
1. determine whether the design constraints proposed by Poh can be applied to real systems;
2. to establish guidelines for optimum operating conditions in real systems;
3. to examine the flow of sludge down the lower inclined surface in more detail and to develop a method for predicting the minimum angle at which sludge flow occurs;
4. to develop the design scheme further to include the following constraints:
   - Steady state constraint to enable the formation of steady state stratified viscous layers in the settling channel
   - Laminar flow constraint
   - Flow stability constraint to minimise the re-entrainment of particles from the suspension layer into the clear liquid layer
Sludge flow constraint to ensure a continual and rapid removal of sludge from the lower inclined surface

and also to consider design constraints imposed by the land area available and by the capital cost of the settler.

Chapter 4 describes the experimental programme and the experimental rigs developed to achieve the above mentioned objectives. The rigs comprise batch ones to study both the flow of sludge and the minimum angle at which sludge flow occurs, and also a continuously operated rig used to obtain comparative data.

The results from the batch tests on the sludge are described and discussed in Chapter 5, while in Chapter 6 the results from the continuous rig are examined.

A mathematical model for predicting the minimum angle at which sludge will flow is developed in Chapter 7. Following on from this the proposed design scheme is outlined.

The conclusions from the research work are presented in Chapter 8 and finally the recommendations for future work in Chapter 9.
### CHAPTER 2
#### LITERATURE REVIEW

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CHAPTER 2

LITERATURE REVIEW

The literature review is divided into five sections. The first section outlines the development of theoretical models for both batch and continuous inclined sedimentation. The existing design methods for sizing lamella separators are then discussed under three main categories which are in accordance with the approaches taken - the empirical, semi-empirical and theoretical approaches. Their main objective is to determine the total surface area required for sedimentation in order to achieve the desired throughput. The next section highlights the practical design considerations. This includes flow patterns and hydraulic conditions, geometric parameters, the design of settling channels, feed entry and sludge collector, materials of construction and the pretreatment of the suspension. Industrial applications and the advantages and disadvantages of a lamella separator compared with a conventional vertical separator are discussed in the final two sections.

2.1 THEORY

There have been a relatively small number of theoretical studies of settling under inclined surfaces and these can be divided neatly into batch models and continuous models.

2.1.1 Batch Inclined Sedimentation Models

The first significant work on inclined sedimentation\(^1\) was by Boycott in 1920. He was studying the sedimentation of blood corpuscles in tubes and observed that they settled faster in inclined tubes. Furthermore, the rate of sedimentation was higher in tubes of smaller bore and in tubes with a higher initial vertical height of suspension. These results are shown diagrammatically in Figure 2.1. Boycott attributed the phenomenon to an effect of the Brownian motion of the
(i) Effect of angle of tilt on settling rate

(ii) Effect of initial height of suspension on settling rate

(iii) Effect of tube diameter on settling rate

FIGURE 2.1: BOYCOTT'S OBSERVATIONS
lower corpuscles in the tube. Several investigators, including Berczella and Wastl\textsuperscript{2}, Linzenmeier\textsuperscript{3} and Lundgren\textsuperscript{4} subsequently advanced hypotheses to explain Boycott's observations but all achieved limited success. Their hypotheses are outlined below.

In 1923-24 Berczella and Wastl\textsuperscript{2} postulated that the friction between groups of particles which were rising along the upper part of the tube wall and falling along the lower part of the wall diminished as the angle of inclination of the tube from the vertical increased. This, they claimed, was responsible for the increase in the settling rate. Linzenmeier\textsuperscript{3} developed these ideas further. He suggested that the corpuscles which fell vertically collided with others sliding down the lower part of the tube wall, resulting in an 'avalanche'. This caused the suspension to circulate and coagulate more rapidly. The enhanced coagulation was thought to increase the settling rate. In 1928 Lundgren\textsuperscript{4} proposed a more logical explanation. He observed a stream of clear liquid flowing upwards beneath the upper inclined surface. From this he concluded that the liquid which was displaced by the settling particles was able to bypass the suspension of particles. Hence the hydrodynamic resistance to sedimentation was less and settling was quicker. Although this idea could explain the effect of altering the tube angle and bore, it could not account for the effect of altering the vertical height of the suspension.

Clearly at this stage there was a need for a fundamental model to satisfactorily explain inclined sedimentation as well as to elucidate its commercial potential.

Based on experimental observations, Ponder\textsuperscript{5} derived a kinematic model which gave a quantitative prediction for the sedimentation rate. This was further developed by Nakamura and Kuroda\textsuperscript{6} twelve years later. Their model, which was originally devised for sedimentation in an inclined square section tube set on its edge, depended on two vital assumptions. These were that:
i) only the downward facing surface accelerated sedimentation

ii) the particles in the settling suspension tended to keep the same distance apart until they alighted upon a solid surface or upon other particles.

They argued that, since the thickness of the liquid layer beneath the upper inclined wall was generally observed to be very small and independent of time, the portion of clarified liquid formed beneath this surface should be instantaneously added to the clear fluid above the horizontal surface. The mathematical development of this model is outlined below.

At the start of settling all particles on a surface denoted by the line CAB (shown in Figure 2.2(a)) were assumed to settle with an initial velocity \( v \) for an elemental time \( dt \) and reach a hypothetical surface DFH. The velocity \( v \) was assumed to have the same value at all points and hence \( AF = BH = CD \). Thus the shaded area ABGFEC represented the volume of clear liquid displaced by the particles in time \( dt \). The volumes represented by CDE and BGH were considered to be negligibly small and were neglected to simplify the mathematics. In reality, however, the particles would not take up the surface shown as EFG because of the density and height difference between the suspension at FG and the liquid at the point E. An instantaneous rearrangement would take place giving a new clear liquid suspension interface at plane A'B' shown in Figure 2.2(b). Nevertheless the two volumes of clear liquid shown must be equal and so the area AA'BB' must equal the area ABGFEC. It was assumed that the initial height of the interface AB was \( h \) and that this fell to a final value \( (h-dh) \) after the elemental time \( dt \). Then, by equating the two areas, a mathematical relationship was obtained which related the enhanced rate of sedimentation to the suspension properties and settler dimensions. Thus

\[
\frac{dh}{dt} = v(1 + h\cos \alpha /b) \quad (2.1)
\]
FIGURE 2.2: NAKAMURA-KURODA INCLINED SEDIMENTATION MODEL
Similar derivations were made to describe the enhanced rate of sedimentation in an inclined tube of circular section

\[-\frac{dh}{dt} = v(1 + 1.27 \ h \cos \alpha/b) \] \hspace{1cm} (2.2)

and for a square section tube resting on one corner

\[-\frac{dh}{dt} = v(1 + 1.414 \ h \cos \alpha/b) \] \hspace{1cm} (2.3)

From these equations it can be seen that, in a square section tube resting on one corner, the settling rate should be higher than in a circular tube which, in turn, should give a higher rate than the simple plane lamella. This prediction was verified experimentally.

The Nakamura-Kuroda equations apparently represent an upper limit to the rate of sedimentation although Zahavi and Rubin\(^7\) and Kinosita\(^8\) have reported higher rates under certain conditions. In general however, later workers including Graham and Lama\(^9\), Vohra and Ghosh\(^10\) and Saida and Ghosh\(^11\) have found that the sedimentation rate was less than predicted by the above equations. They proposed the insertion of empirical coefficients to account for this discrepancy. Graham and Lama\(^9\) introduced a correction factor \(F\) which was determined as a function of concentration for each initial height of suspension and angle of inclination. \(F\) had a greater value for more dilute and more concentrated suspensions and was approximately independent of angle in the range 30° to 70°. Thus

\[-\frac{dh}{dt} = Fv(1 + h \cos \alpha/b) \] \hspace{1cm} (2.4)
Their work was carried out with a system of calcium carbonate and water. Vohra and Ghosh\textsuperscript{10} and Ghosh\textsuperscript{12} introduced a correction factor to the term resulting from the contribution of the ascending clear liquid under the inclined surface. Hence

$$\frac{dh}{dt} = v(1 + Fh \cos \alpha / b) \quad (2.5)$$

However, the correction factors were found to be far from constant and their value was difficult to predict accurately.

Pearce\textsuperscript{13} proposed several reasons for the deviations in settling rate. He observed a difference between the batch settling behaviour of very dilute and concentrated suspensions. With very concentrated suspensions the upward facing surface appeared to support the suspension and thus hindered the particles falling away from the downward facing surface. However, at low solids concentrations the stream of clear liquid could become turbulent and result in the remixing of the clear liquid and suspension. He suggested that the tendency of a material to floc was an important factor. If it was non-flocculating then he predicted that the actual settling rate would be less than the theoretical value and that this deviation would increase with settling rate. This was because the sedimentation process in the suspension layer was hindered by convection currents whose circulation rate increased with an increase in the settling rate. This resulted in a decreased efficiency of the settling process. However when a suspension with a tendency to floc was subjected to convection currents in the inclined vessel, flocculation could either occur more rapidly than in a vertical tube or the flocs could be broken down. If the latter occurred there were two possible effects:
i) For fairly dilute mobile suspensions the settling would be slower than expected. Deviations from theoretical values would increase both as the settling rate increased and the circulation of the suspension became more vigorous. Furthermore the probability of the smaller flocs becoming entrained into the clear liquid stream would increase dramatically.

ii) If the suspension was thixotropic with a small initial stiffness which enabled the lower part of the vessel to support the suspension and hinder the descent of the particles away from the upper part of the wall, then floc breakage would make the suspension more mobile and would increase the rate of settling by reducing the hindering effect.

Zahavi and Rubin\textsuperscript{7} presented a rather complicated model which extended the geometric reasoning of Nakamura and Kuroda. Their model required both a constant for the enhanced sedimentation effect of a given fluid-particle system and the settling rate vs concentration data for vertical vessels. The constant represented the fixed average rate of clear liquid generation per unit area of the downward facing inclined surface in the suspension. Although they achieved good to fair agreement between the theory and their experimental data on clay suspensions the influence of inclination angle and suspension concentration was not completely determined.

They also described\textsuperscript{7} the settling process qualitatively based on experimental observations with dye injected at various points in the settling suspension. They found that the clear liquid layer originated from the suspension itself and not from the settling of solids under the inclined plane. For a right-angled trapezoid, a thin layer of clear liquid was first formed under the upper inclined plane. Additional clear liquid was then supplied from two main sources:

i) from liquid moving up through the suspension because of the increase in the solids concentration at the bottom. They called this the settling effect;
ii) at a given horizontal plane there existed a local pressure difference between the suspension and the clear liquid layer due to the higher density of the suspension. The pressure difference caused liquid to flow from the suspension into the clear liquid layer. This was called the 'filtration effect' and was thought to be the most important effect.

Working with mono-dispersed polymer suspensions Oliver and Jenson\textsuperscript{14,15} observed that the clear liquid formed beneath the upper inclined face was not as suggested by Nakamura and Kuroda but was in fact a roughly-triangular shaped channel (see Figure 2.3). They subsequently developed a mathematical model using the observed profile with the addition of a simple convection term containing an empirical function of concentration and angle of inclination. Mathematical solutions describing the profile of the clear liquid channel as a function of time were obtained on an analogue computer and the general agreement with experiment was fair. However the model seemed to break down at higher concentrations.

\begin{figure}[h]
\centering
\includegraphics[width=\textwidth]{sedimentation_model.png}
\caption{OLIVER AND JENSON'S INCLINED SEDIMENTATION MODEL}
\end{figure}
Sarmiento and Uhlherr investigated the batch settling behaviour of lightly flocculated red mud in inclined tubes. The batch tests revealed the existence of at least two settling regimes. They thought that the two regimes were likely to be analogous to the hindered settling and channelling regimes which are observed in vertical tubes because the liquid-particle and particle-particle mechanistic reactions were similar in nature, even though incline settling is under the additional influence of settling convection. Furthermore they proposed that under the correct conditions the compression regime could also occur. They introduced a correction term to the Nakamura-Kuroda equation which was solely a function of the angle of inclination and claimed that the equation could be applied over a wide range of concentrations provided that the main settling mechanism in both the inclined and vertical batch settling tests was channelling.

From a design standpoint the results of Sarmiento and Uhlherr and of Pearce (discussed earlier) highlighted the importance of using flocculation as a pretreatment step to produce strong and fast settling flocs in order to optimise the actual separation process. In addition a possible explanation for the deviations from theory of the actual settling rates of flocculated suspensions under inclined surfaces was put forward. In general the treatment of non-flocculated suspensions is expected to produce more compacted, higher bulk density sludges than flocculated suspensions which tend to be light and bulky.

It is apparent that all the models discussed above were based on kinematic and geometric considerations only. The fluid dynamic aspects, which must have significant effects, have virtually been ignored. Consequently there are two serious limitations to these models:

1) they cannot provide information about flow characteristics, such as the state of motion and concentration distribution within the suspension and the clear liquid layer formation which together affect the enhanced rate of sedimentation
ii) their range of validity is indefinable. Therefore any mathematical equations derived from the models cannot be used for design purposes with any degree of confidence since it is impossible to tell under what conditions, if any, they are expected to apply.

To rectify the deficiencies highlighted above, attempts have been made recently by Hill\textsuperscript{17,18} and subsequently by Acrivos and Herbolzheimer\textsuperscript{19} to develop more fundamental models using the principles of continuum mechanics. Hill established that the enhancement of sedimentation in inclined planes resulted from a naturally occurring settling convection\textsuperscript{8}, as observed by Kinosita, and was caused by the three following mechanisms:

i) The suspension density gradients, allied with particle concentration gradients, were not parallel to the hydrostatic pressure gradients. Thus convection acted towards establishing a downward gradient of density

ii) The return flow of liquid displaced by settling particles channelling through regions of lower particle concentration

iii) Particles collecting near to and flowing down along the upward facing, inclined surfaces.

Hill\textsuperscript{18} found that particle momentum transfer was the dominant mechanism and experiments were designed in which the other mechanisms could not account for the significant convection. This meant that the mathematical model subsequently developed was limited to defining trajectories of particles in very dilute suspensions where the settling velocity could be described by Stokes Law. The fluid motion was described by a continuity and Navier-Stokes equation to which a term which accounted for the momentum transfer of the settling particles was added. Using dimensional analysis it was shown that aside from the shape and the angle of inclination of the settler, the settling process was governed by two dimensionless parameters.
1. $N_{Re}$, a sedimentation Reynolds number:
   This represented the significance of inertial forces to viscous forces in any convective flow. It was defined mathematically as
   \[
   N_{Re} = h \nu \rho_f / \mu \tag{2.6}
   \]

2. $N_{Gr}$, a sedimentation Grashof number:
   This represented the significance of the gravitational forces relative to the viscous forces in any convective flow. It was defined mathematically as
   \[
   N_{Gr} = h^3 g \rho_f (\rho_p - \rho_f) \omega / \mu^2 \tag{2.7}
   \]
   where 
   - $h$ = characteristic length of macroscale motion (which Hill took to be the initial height of the suspension)
   - $\omega_0$ = initial volume fraction of particles
   - $\nu$ = vertical settling velocity of individual particles at $\omega_0$
   - $\rho_p$ = density of particles
   - $\rho_f$ = density of fluid
   - $\mu$ = viscosity of fluid
   - $g$ = gravitational constant

The model predicted that in order to achieve the most rapid sedimentation, $N_{Re}$ should be made as small as possible and $N_{Gr}$ as large as possible. Furthermore, using experimental results and numerical solutions from their mathematical model Hill et al were able to establish a range of values over which for the Nakamura-Kuroda equations could be applied. This incorporated the dual limit that $N_{Gr} \to \infty$ and $N_{Re} \to 0$. However the range and accuracy of their numerical solutions and experimental data was somewhat limited.

Acrivos and Herbolzheimer\textsuperscript{19} subsequently attempted to theoretically verify the semi-empirical findings of Hill. Applying the principles of continuum mechanics, a model was developed for describing quantitatively the sedimentation of small particles in inclined
channels where the length of the channel was of the same magnitude as the channel spacing. The model treated the settling suspension as an effective fluid and assumed that the flow was laminar and the particle Reynolds number was small - both assumptions being realistic for most industrial applications. It was found that the enhanced rate of sedimentation was indeed dependent on two parameters in addition to the vessel geometry i.e.

1) Re, a sedimentation Reynolds number which was typically small
2) $\Lambda$, the ratio of a sedimentation Grashof number to the Reynolds number which is typically very large.

By means of an asymptotic analysis it was reaffirmed that, as $\Lambda \to \infty$ and for a given settler geometry, the enhanced rate of sedimentation could be accurately predicted with the use of the Nakamura-Kuroda equation. Deviations from this equation were thought to be due to a flow instability which caused the particles to resuspend and thereby reduced the efficiency of the process. The model also produced an expression for the thickness of the clear liquid layer formed beneath the downward facing surface as well as velocity fields in the clear liquid and suspension layers. Under the conditions of their experiments, the agreement with theoretical predictions was excellent.

Acrivos and Herbolzheimer then extended this analysis to describe the sedimentation of dilute suspensions in narrow inclined channels, i.e. where their length in relation to the channel spacing is large. Again, based on the assumptions of laminar flow and small particle Reynolds number, expressions were derived for the clear liquid layer profile as well as the velocity fields in the clear liquid and suspension layers. The sludge layer was ignored and although this could lead to quantitative differences between theory and experiment, the qualitative features of flow should still apply. An unexpected outcome from the solution of the time-dependent equations was that the clear liquid layer formed beneath the downward facing surface attained a steady state profile only below a critical point - above this point
the thickness of the clear liquid layer increased with time until it occupied the entire channel spacing. They were able to show theoretically that, because of the transient behaviour, the Nakamura-Kuroda equations would overestimate the rate at which the top suspension/clear liquid interface settled with time. However, the Nakamura and Kuroda predictions for the volumetric settling rate would still hold under the conditions of the model. It was also shown that the discontinuity in the clear liquid layer profile could be suppressed in continuous settling systems if the feed and withdrawal arrangements were properly designed. They found that excellent agreement was obtained between the theoretical predictions and the batch sedimentation experiments.

Davis et al.\textsuperscript{60} presented a theory to describe the sedimentation of polydisperse suspensions consisting either of \(N\) distinct species of particles or of a continuum of particle sizes. Laminar flow and spherical particles of low Reynolds number were assumed. The hindered settling velocity of a particle was given by its Stokes settling velocity, \(v_0\), multiplied by a function of total local solids concentration, \(f(c)\). The derived equations took the form

\[
S(j)(t) = v_{0j} f(c) \left[ 1 + \frac{h(j)}{b} \sin \theta \right] \frac{b}{\cos \theta} \quad j = 1,2, \ldots N \quad (2.8)
\]

\[
S(v_{0},t) = v_{0j} f(c) \left[ 1 + \frac{h'(v_{0},t)}{b} \sin \theta \right] \frac{b}{\cos \theta} \quad (2.9)
\]

where \(S(j)(t)\) = dimensionless instantaneous volumetric rate at which the suspension is devoid of species \(j\)

\(\theta\) = angle of inclination to vertical

\(h(j)\) = vertical height of \(j\)th region

\(b\) = channel spacing.
2.1.2 Continuous Inclined Sedimentation Models

Very little work has been carried out to develop mathematical models of continuous inclined sedimentation. Most of the existing models are based on or related to an extension of the well established continuous vertical sedimentation models (i.e. the Yoshioka\textsuperscript{21} and Coe and Clavenger\textsuperscript{22} models). Mathematical models developed by Zahavi and Rubin\textsuperscript{23}, Graham and Lama\textsuperscript{24}, Jernquist\textsuperscript{25} and Obata and Watanabe\textsuperscript{26} are examples that fall into this category. Graham and Lama\textsuperscript{24} introduced a correction factor $F$ into the Nakamura-Kuroda equation and obtained a value for the capacity of a lamella thickener by means of a modified Coe and Clavenger technique. They used a system of calcium carbonate in water at feed concentrations from 15.1 to 54.2 g/l. The value for the calculated capacity varied by as much as 50\% from the measured value. They concluded that the calculation method was only suitable as a first approximation of settler capacity.

Zahavi and Rubin\textsuperscript{23} corrected the Yoshioka flux curve method which is used for conventional thickeners. However in some cases they obtained discrepancies of 55\%. Oliver\textsuperscript{15} also recommended the use of this method. He compared the behaviour of dispersed and flocculated systems in a continuous separator and found that the improvement factor for flocculated systems was less than that required for dispersed systems. He attributed this to floc breakage.

Jernquist\textsuperscript{25} determined the maximum capacity and concentration distribution based on the assumption that the settling rate was solely a function of concentration. The analysis was applicable to slurries where the concentration of particles was high. He applied the Kynch and Yoshioka theories in his calculations but gave no comparison with experimental data. Verhoff\textsuperscript{27} derived an equation for the optimization of high rate settlers which involved minimising a critical settling velocity. This velocity was a function of angle of inclination, settler plate length and rectangular area. The optimal critical velocity was then determined by differentiating the equation with
respect to certain variables. The resulting simultaneous equations were solved numerically and graphs were provided for design purposes.

A more fundamental dynamic flow model to describe sedimentation in a continuous system was first developed by Probstein, Yung and Hicks. The two-dimensional model assumed that the flow in any channel of the settler may be treated as comprising of three stratified 'fluid' layers, each of reasonably uniform density and viscosity, moving under the action of gravity; a clarified liquid, a feed suspension layer and a sludge layer (see Figure 2.4). They were able to justify using a constant value for the viscosity in the mud layer by assuming that the velocity in this layer would be low due to its high viscosity. Thus any non-uniformities in the viscosity to the order of a value of 2 would not cause any significant change in the sludge velocity. Further assumptions were that there was no net liquid flux across any plane normal to the channel walls and that the pressure was the same in each of the layers. These were combined with the conditions of no fluid slip at the wall and constancy of shear and velocity across the two stratified layer interfaces. This provided sufficient boundary conditions to describe the flows in the clarified and hindered layers. The mud was specified by assuming that the hindered particle settling rate in the vertical direction was constant and known. The mud flux was taken to be proportional to the component of this rate in the direction normal to the channel. Two significant sets of results emerged from their model:

i) mathematical expressions of scaling laws which were useful for design purposes

ii) for a given settler throughput in a cocurrent system two possible operating modes (subcritical and supercritical) existed. By definition, the subcritical mode (Figure 2.5(b)) occurred when the clear liquid layer thickness was less than half the channel spacing at the top of the settling channel and decreased gradually to a minimum at the base of the channel. In the supercritical mode (Figure 2.5(a)), on the other hand, the clear liquid layer thickness was greater than half the channel spacing
Countercurrent flow

Clarified layer

Cocurrent flow

Mass density

Sludge layer

(1) Clear liquid layer

(2) Suspension layer

FIGURE 2.4: THREE-LAYER MODEL OF PROBSTEIN, YUNG AND HICKS
FIGURE 2.5(a): THE SUPERCRITICAL MODE OF OPERATION

FIGURE 2.5(b): THE SUBCRITICAL MODE OF OPERATION

FIGURE 2.5: THE COCURRENT MODES OF OPERATION
at the top of the settler and gradually increased to a maximum at the settler base.

Probstein and Hicks\textsuperscript{29} found that the supercritical mode was inherently more stable and recommended that it should serve as the basis for a new type of lamella settler with a higher throughput than most commercial settlers which operate in the subcritical mode. Experimental work verified both sets of results.

In the recent work of Probstein and Leung\textsuperscript{30} the three layer model was generalised and applied to evaluate the performance of cocurrently operated lamella settlers and countercurrent tube settlers.

A more fundamental model by Acrivos and Herbolzheimer\textsuperscript{20} suggested that certain ad-hoc assumptions, made by Probstein and his co-workers regarding the existence of three steady-state stratified layers, were oversimplistic and hence only valid under certain operating conditions. Their model showed, for example, that in cases where the feed was introduced into the settler along its side, the feed and withdrawal locations must be chosen properly to enable the formation of steady-state stratified layers. Otherwise, transient behaviour would prevail. The subcritical and supercritical modes of operation were again verified theoretically.

2.2 SIZING LAMELLA SETTLERS

The sizing of lamella separators has been approached in three ways: empirical, semi-empirical and theoretical. These are discussed below.

2.2.1 Empirical Design Methods

Janerus\textsuperscript{31} described a method of sizing lamella settlers based entirely on experimental data from column settling tests. In order to simulate the shorter detention times achieved in inclined settlers, the settling tests were usually performed in a 500 ml graduated cylinder from which the top 100 ml of liquid was withdrawn after a certain time
to simulate a certain loading rate. The depth of the 100 ml layer was chosen to correspond to the plate distance intended for the design. From the settling tests, a relationship between the overflow clarity and the surface loading rate could be obtained. The necessary projected area was then determined for a desired clarity and flowrate. In general the projected area needed to be multiplied by a safety factor of 1.25 to 2. This allowed for non-ideal hydraulic conditions and any unexpected variations in the settling properties. Specifications such as plate inclination, length, width and the feed and withdrawal arrangements were generally specified independently based on the experience and recommendations of the manufacturers.

The empirical approach seems to suffer from several serious drawbacks:

i) the test samples and hydraulic conditions must be both reproducible and representative of the actual full size operation. In practice it is difficult to achieve this

ii) the specifications of most of the design parameters are oversimplistic and thus vulnerable to over or under design.

To alleviate the drawbacks outlined above several semi-empirical approaches have been developed which describe some of the ruling parameters on a theoretical basis.

2.2.2 Semi-Empirical Approach

Graham and Lama\textsuperscript{24} determined the design overflow from the sum of the overflow calculated for a vertical thickener (having the same free air/liquid interfacial area) by the Coe and Clavenger technique plus the additional overflow produced at the inclined surface. The latter was derived by the Nakamura-Kuroda equation

\[-\frac{dh}{dt} = Fv(1 + h/c_\alpha/b)\]  \hspace{1cm} (2.10)
where F was a correction factor. F was found to be a function of the solids concentration in the feed (ranging from 0.5 - 0.7). In their calculations they took a fixed value of 0.56. Their experiments were performed in an inclined thickener comprising two plane surfaces 44 ins by 96 ins at 2.3 ins separation, inclined at 50°. They used precipitated calcium carbonate in water at concentrations ranging from 15.1 to 54.2 g/l. The value for the calculated capacity varied by as much as 46% from the measured value and they concluded that the calculation method was only suitable as a first approximation.

Zahavi and Rubin\textsuperscript{23} modified the Yoshioka flux curve method commonly used to estimate the area of conventional thickeners by the addition of a term which accounted for the additional solids flux contributed by the inclined surfaces. However in some cases they obtained discrepancies of up to 55%.

2.2.3 Theoretical Approach

Two theoretical approaches have been taken in the development of design methods for lamella separators. Both are outlined below:

i) It was assumed that all particles settling in a channel behave independently of each other and thus possess unhindered trajectories of their own. The required surface area to achieve a desired particle removal efficiency was then determined from all the particle trajectories that start and end within the length of the lamella channel. In principle this assumption can only be justified in dilute suspensions where the particles experience unhindered settling behaviour;

ii) The particles were assumed to interact with each other to produce an overall settling behaviour. Hence the suspension was treated as a continuum under the action of gravity. This type of approach was adopted to handle hindered settling conditions.
In the former case the basic equations were formulated more than 70 years ago for ideal settling basins. A similar analysis has been applied to inclined plates. This considered a particle on a limiting trajectory entering the separating zone between two inclined plates at point A and being captured at point B (see Figure 2.6). By assuming plug flow conditions in the settling channel the residence time could be written as

\[ t_R = \frac{LWbn}{Q} \]  

(2.11)

where \( L \) = plate length, \( W \) = plate width, \( b \) = plate spacing, \( n \) = number of settling channels and \( Q \) = volumetric flow rate through the settler.

**FIGURE 2.6: LIMITING TRAJECTORY FOR SETTLING PARTICLES**
Another equation was then formulated for $t_R$ using a modified Nakamura-Kuroda equation.

\[
t_R = \frac{\text{vertical distance travelled by particle along trajectory}}{\text{enhanced settling velocity}}
\]

Thus

\[
t_R = \frac{b}{c\alpha} \quad \text{(2.12)}
\]

\[
t_R = \frac{b}{c\alpha} \quad \text{(2.13)}
\]

where $F$ was an empirical coefficient to account for inaccuracies in the Nakamura-Kuroda equation and $L\sin\alpha$ replaced $h$ in the original equation.

By equating the two residence times an expression relating the required lamella plate area to the desired separator throughput was obtained. Thus

\[
WLn = \frac{Q}{Fv[1 + L\sin\alpha\cos\alpha/b]} \cos\alpha \quad \text{(2.14)}
\]

The settling velocity $v$ and the improvement factor $F$ needed to be estimated or measured. In the literature $F$ was found to vary with suspension concentration and typically lay between 0.5 and 0.7 over a wide range of concentrations likely to be encountered in industrial applications.

Based on the residence time approach other more complicated models have been devised by researchers to account for different vessel shapes, the actual profile of the liquid flow and the thickness of sludge accumulated in the settling channel.

The earliest attempt to apply a theoretical analysis to inclined sedimentation under hindered settling conditions was by Jernquist. His method predicted the maximum capacity and concentration...
distribution for high feed concentrations. It was based on the assumptions that

i) the settling rate was solely a function of concentration

ii) the thickness of the clear liquid layer beneath the upper inclined surface and the thickness of the sludge layer on the lower lamella surface were negligible.

From material balances at steady state conditions he obtained expressions to describe the concentration distribution and solids flux along the entire length of the lamella thickener. The maximum thickener capacity could then be determined from graphs of solids flux versus solids concentration. He found that, under the conditions of the analysis, lamella thickening was a special case of vertical thickening.

Probstein et al\(^{28}\) developed a more general design procedure in which the capacity of the lamella separator was determined from mathematical expressions of the velocity fields in the different settling zones that exist in the settling channel. The throughput was shown to scale with a characteristic flux which was a maximum limiting value corresponding to the flux in a channel with two stratified layers moving in opposite directions each occupying half the channel height. Furthermore the channel length scaled with a characteristic length that was a measure of the solids settling capability within the channel.

2.3 PRACTICAL DESIGN CONSIDERATIONS

The following section discusses the various parameters which affect the capacity of a lamella separator as identified in the literature. Furthermore it outlines the design guidelines suggested for an optimum design of settler.
2.3.1 Flow Patterns

There are several designs of lamella separator available. These are discussed below. The factors which influence the selection of flow pattern are outlined.

2.3.1.1 Types of flow pattern

There are three main flow patterns under which lamella settlers can be operated:

i) Countercurrent flow in which the feed and sludge streams move in opposite directions (illustrated in Figure 2.7(a)). These settlers are available in several designs (shown in Figure 2.8). In the Lamella Gravity Settler (see Figure 2.9) the feed is introduced through a bottomless rectangular feed box which extends above the liquid level. Feed flows into the inclined chambers through the sides of the feed box and passes upwards until it leaves the settler through distribution orifices.

ii) Cocurrent flow in which the feed and sludge streams move in the same direction (illustrated in Figure 2.7(b)). A typical cocurrent sedimentation system is illustrated in Figure 2.10. The incoming feed is distributed evenly over the plates. It then travels downwards to the bottom of each plate where a return tube carries the effluent back to the top of the unit where it is discharged.

iii) Crosscurrent flow in which the direction of the feed stream is perpendicular to the sludge stream (illustrated in Figure 2.7(c)). A cross flow thickener has been developed by Rheax in which the plates are inclined in an herring bone pattern. The suspension proceeds horizontally across the plates while the sludge slides down them.
FIGURE 2.7(a): Countercurrent flow

FIGURE 2.7(b): Cocurrent flow

FIGURE 2.7(c): Crosscurrent flow

FIGURE 2.7: THE MAIN FLOW PATTERNS FOR THE LAMELLA SEPARATOR
FIGURE 2.8:  DESIGNS OF LAMELLA THICKENER
FIGURE 2.9: LAMELLA GRAVITY THICKENER
FIGURE 2.10: TYPICAL COCURRENT SEDIMENTATION SYSTEM
A more complex design is the dual flow\textsuperscript{32} classifier (illustrated in Figure 2.11) in which both cocurrent and countercurrent flow can be achieved in the same equipment. The manufacturers claim that this is advantageous when multiphase or heterogeneous systems are being separated. Figure 2.12 shows a separator unit in which clarification and thickening can be achieved in the same equipment by having two packs of lamella plates vertically above each other with the feed introduced between them. The plate separations can be different in the two sections and the lower pack may be vibrated to aid the compaction of the sludge.

\textbf{FIGURE 2.11: DUAL FLOW CLARIFIER}
2.3.1.2 Factors influencing the choice of flow pattern

The main criteria determining the choice of flow pattern are:

i) minimising the re-entrainment of particles into the clear liquid stream

ii) the rapid removal of particles from the plates.

The mathematical model of Probstein et al\textsuperscript{28,29} predicted that for applications where high quality supernatant was required the more stable cocurrent supercritical mode of operation was superior. However it is claimed there is a limit to the volumetric flow rate\textsuperscript{41} of sludge that can be handled.

Forsell and Hedstrom\textsuperscript{41} found that the cocurrent flow design was particularly suitable for light sludges with low yield stresses in which the sludge volume fraction was small. They thought that this was because the cocurrent flow of the suspension also provided a drag force down the slope in addition to the gravitational force which acts
to move the sludge layer. The gravitational force alone may be insufficient to cause any movement. An important application is floc separation in the treatment of surface water. Both flow patterns may be used in the treatment of heavy finely-dispersed sludges with low sludge volume fractions. However, countercurrent systems are generally used due to their simpler and less expensive design. An example is the clarification of circulating water used in wet scrubber plants.

Countercurrent flow settlers are nearly always used for suspensions with high volume fractions of solids. Examples include the separation of biological flocs in the activated sludge waste treatment process and the separation of metal hydroxide on the neutralisation of waste liquors from pickling plants and the galvanic industries. These recommendations are summarised below in Table 2.1:

TABLE 2.1: FACTORS INFLUENCING THE CHOICE OF FLOW PATTERN

<table>
<thead>
<tr>
<th>Sludge Volume Fraction</th>
<th>Type of Sludge</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Light, network-forming</td>
</tr>
<tr>
<td></td>
<td>Low yield stress</td>
</tr>
<tr>
<td></td>
<td>Heavy, finely-dispersed</td>
</tr>
<tr>
<td></td>
<td>High yield stress</td>
</tr>
<tr>
<td>Low</td>
<td>Cocurrent</td>
</tr>
<tr>
<td>High</td>
<td>Countercurrent</td>
</tr>
</tbody>
</table>

Schlitter and Markl\(^{37}\) described the operation of a cross flow thickener. They claimed that, because the particles settle in a direction that is normal to the flow of liquid, the settler can be operated satisfactorily at higher relative velocities than more conventional designs. Furthermore, since the drag of the liquid does not oppose the discharge motion of the sludge, then the angle of inclination of the plates can be selected independently.
2.3.2 Hydraulic Conditions

2.3.2.1 Laminar flow

It is important to establish laminar flow conditions in order to ensure that particles maintain a steady descent onto the lower inclined plate and are not intermittently swept upwards by turbulent currents. A low liquid Reynolds number, $R$, characterises laminar flow conditions in the separator where

$$ R = \frac{\rho V D}{\mu} $$

(2.15)

where $\rho = \text{liquid density}$, $V = \text{flow velocity}$, $\mu = \text{liquid viscosity}$, $D = \text{hydraulic diameter}$.

Hence, non-turbulent conditions can be achieved by reducing the hydraulic radius of the lamella channels. To assist in the development of laminar flow, adequate provision must be made to reduce the kinetic energy of the incoming feed stream to the separator. This is generally achieved by fixing an impingement plate to absorb the impetus of the feed stream just before it enters the lamella channels.

2.3.2.2 Even flow distribution

In order to use the plate area efficiently, the flow should be distributed evenly between the plate spacings as well as widthways across the plate spacings. Otherwise a bypass situation will develop in which some parts of the separator will become overloaded while others are underloaded.

2.3.2.2.1 Between plate spacings

If a flow distributor is not included in the design, the velocity through some spacings could be 100% larger than average which could mean that an extra 50 to 100% projected area is needed. Janerus
suggested two possible reasons for the maldistribution of flow:
i) the velocity in the feed and effluent chambers could be higher
than the velocities between the plate spacings. This means that
the dominating pressure drop is not through the plates;
ii) non-levelled equipment. Thus one flow path through the unit might
offer a larger static head than another path.

These problems can be remedied by including an orifice (see Figure 2.13) on the effluent side of each plate spacing which should be sized
to give a pressure drop of 2" to 3". This should\textsuperscript{38} create a
sufficient back pressure to force the bulk content to be distributed
evenly over the entire volume of the separator. It is inadvisable to
put a restriction at the feed entrance as the high inlet velocity
resulting would destroy flocs and prevent laminar flow over a large
region of the plate spacing.

Furthermore problems could arise from the plugging of the orifice. A
less desirable but possible alternative is to allow the effluent to
exit over a weir.

2.3.2.2.2 Across plate spacings

The settling channel is best utilised when the upward velocity is
constant across the entire width of the plate. Janerus recommended
that this could be achieved by:

i) keeping the entrance velocity low to avoid jet stream formation
ii) a length to width plate ratio of at least 5:1 to allow a stable
and even flow pattern to develop
iii) removing the liquid from the top of the plate as evenly as
possible. For example, a side effluent weir can create
undesirable flow patterns.
FIGURE 2.13: TYPICAL DESIGN ARRANGEMENT FOR ACHIEVING EVEN FLOW DISTRIBUTION

2.3.2.3 Environmental factors

Any significant variation in the temperature of the feed stream can generate thermal and density currents leading to the short circuiting of flow. Little\textsuperscript{45} observed this in a conventional clarifier when the temperatures of the feed stream and the bulk content differed by only 2°C. White et al\textsuperscript{46} found that when the temperature of the feed stream was higher than the bulk content by as little as 0.2°C, very poor distribution of flow occurred with practically all the flow passing up
the first tube. The most effective way of preventing this problem was to insulate the entire separator unit, which should be feasible because of its compactness.

2.3.3 Geometric Parameters

The design of the main body of the settler needs careful consideration in order to provide an optimum design. This includes the parameters of place spacing, width, inclination and length. Very little conclusive work has been carried out in this area.

2.3.3.1 Plate Spacing

The plate spacing\textsuperscript{31,37,47} should be as narrow as possible in order to achieve a maximum number of plates in a given settler volume. The more plates that there are in a given settler volume, then the greater the total projected area and hence the greater the settler throughput. However, the lower limit on the plate spacing is governed by potential clogging problems and the possibility of particle re-entrainment into the clear liquid stream. Frequent and severe clogging problems have been reported in most waste water applications but are almost non-existent in surface water treatment. In most lamella thickeners the plate spacing is about 5 cm.

2.3.3.2 Plate inclination

For a compact unit\textsuperscript{31,37} the angle of inclination should be as shallow as possible. Furthermore a shallow angle will prevent the sludge from flowing at too high a velocity which could result in the formation of eddies and the subsequent remixing of the sludge into the suspension layer. On the other hand, however, the angle must be steep enough to allow the sludge on the plates to be continuously and rapidly removed. For most applications the angle of inclination varies from 45° to 60°. Lower angles can be used satisfactorily in cocurrent operation. This was thought to be because the net forces moving the sludge layer downwards are the gravity force and the shearing force of the flowing
suspension. In countercurrent operation the shear force opposes the gravity force and thus hinders sludge transportation, whereas in cocurrent operation there is the cooperation of the two forces.

2.3.3.3 Plate length

With very long plates the velocity\(^{31}\) between them can be high enough to create turbulent flow. Janerus recommended that, for most applications, 3m plates would offer a compact enough solution without creating excessive velocities between the plates. Recently\(^{48}\), the existence of an optimum aspect ratio (vertical height of plate to plate spacing) has been found experimentally. If the settler length was increased beyond the optimum aspect ratio the increase in the maximum overflow achieved was small or non-existent due to flow instabilities which resulted in an uneconomic design of settler. It was found that for higher concentrations the optimum aspect ratio decreased. The values determined for the optimum aspect ratio are listed below:

<table>
<thead>
<tr>
<th>Feed Concentration</th>
<th>Optimum Aspect Ratio</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.5% v/v</td>
<td>25-30</td>
</tr>
<tr>
<td>2.0% v/v</td>
<td>15-22.5</td>
</tr>
</tbody>
</table>

Hedburg\(^{43}\), however, recommended a plate length to plate spacing ratio of 40:1.
2.3.3.4 Plate width

With wider plates the risk that jet streams will develop is greater. Janerus\textsuperscript{31} recommended a plate length to width ratio of 5:1 in order to avoid the potential problem.

2.3.4 Design of Settling Channels

The shape and configuration of settling channels are important considerations for achieving the optimum settling capacity. It has been suggested\textsuperscript{49} that the settling distance should be uniform so that most particles have the same settling time. Hence circular tubes were considered to be inefficient because the particles entering the top of the tube have a greater distance to settle than those entering at the sides. Furthermore circular tubes are less efficient because of the large amount of dead space between tubes in the array.

The Chevron Tube Settler module consists of an array of nested 24\" long extended polystyrene tubes with a cross-sectional chevron space (see Figure 2.14). The chevron design was developed by the Permutit Company and is claimed to give optimum settling characteristics because:

\begin{figure}[h]
\centering
\includegraphics[width=0.2\textwidth]{chevron_design.png}
\caption{Chevron Design for Settling Channel}
\end{figure}
i) the 1-in chevron configuration has the highest perimeter of any common shape for the same area;

ii) the settling distance for particles entering anywhere along the top of the tube is the same;

iii) the V-groove promotes optimum sludge compaction and flow.

The above claims were based on empirical findings which have to be experimentally verified. The reason being that there are a wide range of commercial units using different configurations which all claim to have advantages of their own. These include the honeycomb cross-section tubes, inclined parallel plates and the inclined corrugated plates\textsuperscript{50}.

2.3.5 Materials of Construction

In general, the lamella plates are constructed out of different types of plastics, grades of fibre glass-reinforced plastic and polyvinylchloride. The most popular metal is stainless steel\textsuperscript{31}.

The larger tanks are often made from carbon steel which has been epoxy painted or coated for chemical and physical protection. Other materials used include aluminium, stainless steel and rubber-lined carbon steel. Smaller tanks, on the other hand, tend to be made of fibre glass-reinforced plastic.

2.3.6 Feed Entry

At present\textsuperscript{31,38,47}, the feed is introduced into a feedbox from which it gains access to all plate channels through feedports which are located a short distance above the base of the plates. This eliminates the possibility of the re-entrainment of particles falling from the plates into the sludge collector. Moreover, the content in the feedbox will absorb the impetus of the incoming feed stream, thus helping to sustain laminar flow conditions within the plate channels.
Probstein and Hicks\textsuperscript{29} described two possible modes of cocurrent operation, subcritical and supercritical. The nature of the feed inlet dictated which mode was attained.

2.3.7 Design of Sludge Collector

The simplest means of sludge collection\textsuperscript{31} and withdrawal is a hopper. Care must be taken when designing the hopper. Shallow hoppers tend to rathole so that the velocity is much higher in the central parts of the hopper than at the walls. If this happens it is possible that dilute sludge can be discharged from the settler or that the relatively stagnant layers near the walls may eventually grow to fill the entire hopper thus rendering it inoperable. A steep hopper with a side angle of at least 55° is recommended.

One manufacturer included a low frequency vibrator in the sludge collector which aided the compression of the sludge and helped to compensate for the relatively short sludge detention time. This is particularly suitable\textsuperscript{41} for finely dispersed mineral sludges. In addition, the applied vibrations improved the flow characteristics of the sludges (which are mostly thixotropic) by lowering their apparent viscosities. However with sludges that form loose networks better results are obtained by gentle agitation using a rake mechanism.

2.3.8 Pretreatment of Suspensions

The settling characteristics of suspensions are often improved by a coagulation or flocculation step\textsuperscript{47,51} before they enter the settler. The amount of coagulant used needs to be carefully controlled to prevent either incurring excessive chemical costs, or greatly increasing the volume or mass of sludge for disposal. It is also counterproductive to use excessive coagulant because it can lead to charge reversal and stabilisation of a suspension. The coagulant is flash mixed, particularly important when using organic polyelectrolytes which are fast acting. A period of gentle mixing is then generally provided to promote the growth of flocs. The
flocculated suspension must flow gently into the sedimentation tanks to prevent floc breakage. Most commercial settler units provide special compartments for flocculation.

2.4 INDUSTRIAL APPLICATIONS

It has been estimated that there are approximately one thousand lamella separators in use worldwide, about half those being located in America[^31]. Their growth in popularity is mainly due to design improvements which have substantially improved their performance.

Lamella separators find wide applications where solid-liquid separation by pressure filtration is prohibited due to highly resistive, compressible filter cakes. They are especially useful where particle settling rates are so low that unacceptably large conventional settlers would be required.

2.4.1 Waste Water Treatment

The Water Research Centre conducted[^52] extensive studies on the insertion of inclined plates or tubes into sedimentation tanks for waste water treatment. They found that the performance of a module of sloping tubes installed in half a primary sludge tank was impaired when the bulk temperature of the tank was significantly lower than the incoming feed. Problems were also encountered in inclined tubes installed in the final settlement tanks of an activated sludge process. Periodic discharge of solids occurred and cleaning the tubes occupied as much as two days a week. The most advantageous application was in the uprating of humus tanks where a three times increase in capacity was obtained.

In a major project to use the Danube[^53] for hydroelectric power, cross flow lamella thickeners were used to recover fines from a sand classification plant. Due to strict water pollution regulations the water had to be returned in virtually its natural state. The plant was designed with a tank surface of 24 m² to handle 1150 m³/hour of
water which contained up to 100 tonnes/hour of less than 0.5 mm solids. It was claimed that the overflow contained less than 0.2 g/l of solids and that all material above 0.063 mm was removed.

York described the performance of an inclined plate separator for the removal of sludge from 40% phosphoric acid. The effectiveness of the lamella separator was compared to a conventional raked tank settler and a nozzle discharge disc centrifuge. He carried out pilot plant work using an inclined plate settler with a total projected area of 70 m² at an angle of 45°. The feed solids content was 3 wt% and he reported that an overflow concentration of between 12 to 15 wt% and an underflow concentration of 0.5 wt% were obtained.

Fischer outlined several applications of the lamella settler in the iron and steelmaking industries for the treatment of wet scrubber water. For most of these systems, the requirements were less than 100 mg/l suspended solids in the overflow, sludge concentrations of 15-20% and surface loadings approaching 1.0/ft². In blowdown streams or where metal hydroxides predominate, effluent solids concentrations of less than 20 mg/l and hydraulic loadings of up to 0.6/ft² were expected. The sludge underflow concentration with a Lamella Gravity Settler (LGS) type settler is found to be in the range of 0.5 to 2.0 wt% while that of a Lamella Gravity Settler/Thickener (LGST) type model ranges from 2 to 6 wt%.

Janerus and Lucas gave details of several hydroxide applications using the LGS or the LGST type settlers. In the treatment of waste water from a copper foil electroplating process the suspended solids content was reduced to less than 10 ppm, while rinse water from a zinc plating process containing 100 to 200 ppm was reduced to 26 ppm with the aid of a flocculating agent. In tertiary phosphorus removal (the phosphorus content of municipal waste water in some parts of the United States must be less than 1 ppm), a coagulant was used which precipitates out aluminium hydroxide.
2.4.2 Mining

There are many applications in the mining industries. Schlitter and Markl\textsuperscript{37} described a cross flow thickener used in the gold and copper mining industry in South Africa. They compared a 23m diameter conventional thickener with a settling area of 420 m\textsuperscript{2} to a cross flow thickener with a settling area of 1000 m\textsuperscript{2} in a volume of 100 m\textsuperscript{3} with the same duty of thickening flocculated hydrocyclone overflow. The lamella settler was found to be less sensitive to changes in the overflow than the conventional tank.

Cook and Childress\textsuperscript{47} pointed out that the use of thickeners has increased owing to stringent environmental regulations which limit the use of sludge ponds and to the economic incentive to clean more coal. The main areas of application lie in the clarification, thickening and fine classification of ores. Examples include the treatment of waste water in underground mines, the thickening of solids between milling systems and flotation plants, the thickening of tailings and concentrates and the improvement to water clarity in dressing plants.

2.4.3 Other Uses

Lamella thickeners are also being applied in oil-water separation\textsuperscript{50,57} and for dissolved air flotation\textsuperscript{32}. Advantages in the air flotation process include additional separation surface and better hydraulic control. Plate settlers using this principle have already been developed by CJBD and Anpress.

2.5 ADVANTAGES AND DISADVANTAGES

This section highlights the advantages and disadvantages of a lamella separator over a conventional vertical separator.
2.5.1 **Advantages**

Capital and operating costs are lower because not only can higher separating efficiencies be obtained, but the space required for an inclined settler can be as little as 10% of that needed for a conventional settler. Its compact size means that it can be installed close to the source of feed which can result in additional savings in pipe and pump insulation costs.

The lamella separator has a higher separating efficiency since laminar flow conditions can be achieved within the settling channels eliminating the effects of flow currents which may impede settling and cause short-circuiting. Lamella thickeners are now known to be capable of producing high sludge concentrations not previously achievable with the use of conventional thickeners.

Units are usually factory assembled which results in minimum on-site work. They tend to be made of low weight plastics and metals which makes it possible to install them on high locations in buildings and enables them to be moved easily from one location to another.

Maintenance problems should be low because there are fewer moving parts that require maintenance and that may malfunction. Sludge piping is easily accessible.

Lamella separators have a low sludge holding capacity and short detention time. This is an advantage if the sludge is valuable or if the settler needs to be emptied periodically since there will be little wastage of material.

It is possible to introduce lamella plates into existing clarifiers. This can be advantageous because additional land area is not required. The carry-over of suspended material in an existing clarifier can be reduced and the flocculant dosage decreased leading to savings in chemical costs. Alternatively the settler plates can be used for uprating the process.
Lamella settlers need not be affected by wind or thermal currents since they are easy to insulate and cover. They have much lower evaporation and heat loss compared to conventional settlers.

Thus, to summarise, the main advantages of lamella separators are:

a) Lower capital and operating costs  
b) Higher separating efficiencies  
c) Convenience of construction and installation  
d) Fewer maintenance problems  
e) Low sludge holding capacity  
f) Can uprate existing clarifiers  
g) Insensitivity to external conditions.

2.5.2 Disadvantages

There is a possibility that the overflow can be contaminated by entrained sediment. This can be caused by internal mixing in the settler when care is not taken in the design of the feed arrangements and in operating the unit.

The low sludge holding capacity can mean that separate storage facilities are required for the sludge.

The relatively short detention time for compression is a disadvantage where high sludge concentrations are created by long periods of compression. In practice this can be compensated for by the low amplitude vibration or agitation of the sludge.

Lamella settlers are particularly vulnerable to fouling problems because of the presence of a large number of plates and narrow plate spacings. Hence they are not recommended for applications with a large scaling potential, especially where the scale cannot easily be removed. Sticky solids also cause problems both with plate fouling and the periodic flushing of solids from the plates.
Thus, the main disadvantages of a lamella separator compared to a conventional settler are:

a) Possibility of contamination of overflow by entrained sediment
b) Separate storage facilities for the sludge may be required
c) Short detention time
d) Susceptibility to fouling problems
CHAPTER 3

DESIGN CONSTRAINTS

This section aims to outline the areas in which improvements can be made to the design of a lamella separator and the constraints which it is deemed are essential for the successful operation of the separator.

3.1 EXISTING DESIGN METHODS

The existing design methods are limited in their extent and accuracy, particularly in thickening applications. Extensive pilot plant trials are necessary to finalise design specifications since process engineers are not able to rely on available design methods. The more well tested design methods reported in the literature are only about 50% accurate in predicting the required separator capacities and hence large safety factors need to be incorporated.

Two common problems encountered are re-entrainment of particles into the clear liquid layer and the inability to achieve the designed level of sludge thickening. Poh\textsuperscript{48} considered that, although previous workers had attributed the deficiencies of the existing models largely to stability problems and mixing, the models themselves had inherent weaknesses due to the use of too many ad-hoc assumptions. Using an ideal system of glass beads dispersed in Reofos 65/Reomol DBP plasticiser liquid he was able to rectify some of these deficiencies. He found that the Nakamura-Kuroda equation was capable of accurately predicting the maximum overflow rate of a lamella separator provided that:

i) the requirements for achieving steady state conditions are met

ii) flow instability which leads to excessive particle re-entrainment into the overflow stream is reduced by the careful selection of the settler dimensions.
As can be seen from the above, there is clearly both a need and a potential to improve the design procedure for lamella separators.

3.2 RESEARCH OBJECTIVES

The following constraints were considered essential for the successful operation of a continuous separator:

i) Steady state constraint
To ensure the formation of steady state stratified viscous layers in the settling channel

ii) Laminar flow constraint
To minimise particle re-entrainment into the clear liquid layer

iii) Flow stability constraint
To ensure a continual and rapid removal of sludge from the lower inclined surface

iv) Sludge flow constraint

It is the intention of the following experimental study to determine whether the constraints outlined above can be successfully applied to non-ideal systems and to adapt, if necessary, existing design schemes to take account of non-ideality. In addition the design scheme will be extended to take into account the constraints of limited land area and limited capital expenditure. A detailed experimental investigation of the sludge flow constraint will be undertaken and a method proposed for determining the minimum angle at which the flow of sludge will occur.

Firstly, however, the constraints which are deemed necessary for the successful operation of the lamella separator will be discussed in detail below.
3.3 CONSTRAINTS

3.3.1 Steady State Constraint

Steady state is assumed to be inherently attainable in most existing design methods. For example, Probstein et al. developed a design method based on a dynamic flow model which assumed the existence of steady state stratified layers in the settling channels. However, steady state stratified layers do not occur in all cases as Acrivos and Herbolzheimer have since shown both theoretically and experimentally. They found that in a low aspect ratio separator, the assumption of steady state was valid. Using continuum mechanics an equation was developed which predicted that, at any fixed position, along the upper inclined surface, the clear liquid layer thickness, , increased linearly with time until it reached a steady state value whereupon it became independent of time. The equation for the steady state clear liquid thickness was given by:

\[(\delta)_{\text{steady state}} = [3\tan \theta]^{1/3}\]  

(3.1)

where

\[\delta = \Lambda^{1/3} \delta\]  

(3.2)

\[\theta = \text{settler angle to vertical}\]

\[\Lambda = \text{ratio of Sedimentation Grashof Number to Sedimentation Reynolds Number}\]

\[= gh^2 [p - \rho_f] c_0 / \mu V\]  

(3.3)

\[g = \text{acceleration due to gravity}\]

\[h = \text{initial vertical height of suspension}\]

\[\rho_p = \text{solids density}\]

\[\rho_f = \text{liquid density}\]

\[\mu = \text{liquid viscosity}\]

\[V = \text{vertical settling velocity}\]

+ A low aspect ratio settler is one in which the vertical height is of the same order of magnitude as the channel spacing.
All position coordinates were made dimensionless with \( h \) (the characteristic length of the macroscale motion which was taken to be the initial height of the suspension).

In the case of a high aspect ratio*, however, there were constraints on the dimensions and design of the separator which needed to be satisfied\(^{20} \) before steady state conditions could be achieved. The mathematical theory showed that the clear liquid layer that was formed along the length of the separator attained steady state only below a certain critical point. Above this point the thickness of the clear liquid layer increased rapidly with time until it occupied the entire channel spacing (see Figure 3.1).

The following equations were developed to predict both the thickness of the clear layer in the steady state section

* A high aspect ratio separator is one in which the vertical height is much greater than the channel spacing
\[
\delta_{\text{steady state}} = \frac{\hat{B}}{2} \left[ 1 - \sqrt{1 - \frac{4}{\hat{B}} (3 \tan \theta)^{1/3}} \right]
\]  
(3.4)

where \( \hat{B} = \Lambda^{1/3} \)  
(3.5)

\( B = \) dimensionless plate spacing

and the point of discontinuity

\[
\chi_c = \frac{\hat{B}^3}{192 \tan \theta}
\]  
(3.6)

Acrivos and Herbolzheimer\(^{19,20}\) also developed equations for the velocity field within the clear liquid layer. Agreement between their experimental and theoretical results was very good.

Poh\(^{48}\) was able to further verify these findings experimentally by measuring the steady state thickness of and the velocity field within the clear liquid layer in batch inclined sedimentation. He found that at low aspect ratios the agreement of the experimental with the predicted values was above 80\% for solids concentrations of up to 30\% v/v. At high aspect ratios, as had been predicted by the theory, it was not possible to maintain a steady state clear liquid/suspension layer interface along the entire length of the separator under all settler configurations. In certain cases a critical point of discontinuity was observed above which the interface was turbulent. The theory of Acrivos and Herbolzheimer\(^{20}\) was able to predict both the settler configurations at which this occurred and the position of the critical point. By conducting a series of experiments Poh was able to define a set of conditions under which the behaviour of inclined sedimentation could be accurately predicted by the Acrivos and Herbolzheimer models. These were

Low aspect ratio case:  
1.13 \( < h/b < 3.42 \)  
1\% v/v \( < c_0 < 30\% \) v/v  
20° \( < \theta < 70° \)  
7.61 \times 10^4 \( < \Lambda < 8.48 \times 10^7 \)  
0.17 \( < Re < 2.12 \)
High aspect ratio case: $41.31 < h/b < 75$

$1\% \text{ v/v} < c_0 < 2.5\% \text{ v/v}$

$20^\circ < \theta < 45^\circ$

$4.33 \times 10^8 < \Lambda < 5.47 \times 10^8$

$3.88 < Re < 10.76$

From a design point of view for a continuous separator the above results are important since

i) they establish the feasibility of operating a low aspect ratio separator on a continuous basis since steady state conditions can easily be achieved;

ii) for low aspect ratio separators they permit the use of existing design methods such as those proposed by Probstein et al where the ad hoc assumption of steady state layers is made;

iii) they make it feasible that steady state conditions can be achieved in a high aspect ratio separator providing the feed and withdrawal arrangements and/or settler dimensions are carefully chosen to suppress the transient behaviour.

Thus it should be possible, in principle, to attain steady state conditions for all values of aspect ratio. Examples of how such design constraints can be applied to the more common modes of operation are detailed below:

I. **Cocurrent flow**

In order to ensure that steady state conditions are obtained, it is necessary that

$$B^3 > 192 \times \text{tang}$$

(3.7)

In addition, two possible modes of operation can be used, the subcritical and supercritical modes. This is consistent with the
results of Probstein et al\textsuperscript{29}.

II. Middle feeding
Provided that a significant portion or all of the feed is introduced into the separator below the critical point of discontinuity, then steady state conditions are attainable.

III. Countercurrent flow
Steady state conditions can be achieved under all sets of operating conditions since all the feed is introduced below the point of discontinuity. No additional constraints on the separator dimensions are necessary.

In the following work the applicability of the above constraints to real systems will be examined experimentally in order to determine whether they can be used as guidelines for design purposes. Two systems have been selected, limestone of size ranges 5-10 $\mu$m and 45-53 $\mu$m both dispersed in water. In order to achieve steady state conditions in the cocurrent mode the theory predicts that the following constraints should be met. For limestone of size range 5-10 $\mu$m

$$b^3 > 8.53 \times 10^{-8} \times \tan \theta \times L \tag{3.8}$$

and for limestone of size range 45-53 $\mu$m,

$$b^3 > 4.33 \times 10^{-6} \times \tan \theta \times L \tag{3.9}$$

where: $b =$ plate spacing
$\theta =$ angle to vertical
$L =$ settler length
Taking the worst possible case, this gives the following conditions. For limestone of size range 45-53 μm,

\[ b \geq 0.02m \]  \hspace{1cm} (3.10)

and for limestone of size range 5-10 μm

\[ b \geq 0.005m \]  \hspace{1cm} (3.11)

3.3.2 Laminar Flow Constraint

It is important to operate under laminar flow conditions for two reasons:

i) It is a requirement for the formation of steady state stratified layers;

ii) It ensures that there are no turbulent currents generated within the separator which would intermittently sweep the settling particles upwards.

Laminar flow is characterised by a low liquid Reynolds number \( R \) for flow through the settler and can be achieved by reducing the hydraulic diameter of the settling channel. Thus

\[ R = \frac{\rho_f}{\mu} \frac{VD}{\rho_f} \]  \hspace{1cm} (3.12)

where

- \( R \) = liquid Reynolds number
- \( V \) = velocity of fluid flow
- \( \rho_f \) = density of fluid
- \( \mu \) = viscosity of fluid
- \( D \) = hydraulic diameter of settling channel

\[ D = 4 \times \frac{\text{cross-sectional area of settling channel}}{\text{wetted perimeter of settling channel}} \]
For a parallel plate lamella separator of channel width, \( W \), and plate spacing, \( b \), the hydraulic diameter is given by

\[
D = \frac{2Wb}{W+b}
\]  

(3.13)

Combining equations 3.12 and 3.13 gives

\[
R = \frac{2\rho_f WbV}{(W+b)\mu}
\]

(3.14)

For laminar flow conditions the liquid Reynolds number should be less than 2000. For safety, a lower limit of 500\(^4\) has been proposed. Hence

\[
\frac{2\rho_f WbV}{(W+b)\mu} < 500
\]

(3.15)

Now

\[
V = \frac{Q}{Wb}
\]

(3.16)

Inserting 3.16 into 3.15 and rearranging gives

\[
W > \frac{2Q\rho_f}{500\mu} - b
\]

(3.17)

The channel spacing, \( b \), is normally specified independently from equation 3.17 and is based on potential clogging problems. Thus the channel width must be equal to or greater than the right hand side of equation 3.17 above.
In order to maintain an even distribution of flow across the entire width of the channel, it is also necessary to impose an upper limit to the channel width. For this purpose it is generally accepted that the channel length to width ratio should be at least 5:1.

The limestone system of size range 5-10 μm is well within the above constraint with Reynolds numbers for flow ranging from 8-36. However, the limestone of size range 45-53 μm has Reynolds number values of between 615-1860 which is above the recommended value of 500. Nevertheless the Reynolds numbers are still less than 2000, the usual limit for laminar flow. These values were determined using equation 3.14 and substituting for \( V \) using equation 3.16. The value of \( Q \) used was that predicted for the overflow by the Nakamura-Kuroda equation (3.21). The underflow rate was assumed to be low and hence negligible.

3.3.3 Flow Stability Constraint

The contamination of the clear liquid layer with particles re-entrained from the suspension layer is one of the main factors causing substantial reductions in the overall separation efficiency of a lamella separator.

At present most lamella separators are operated in either the countercurrent or cocurrent subcritical modes. Probstein et al.\(^{29}\) have shown that the countercurrent and cocurrent subcritical modes of operation are more susceptible to particle re-entrainment than the cocurrent supercritical mode. This is because in the former modes there is an unfavourable velocity field at the clear liquid/suspension interface which results in part of the suspension layer being dragged along the direction of the clear liquid layer flow, as is illustrated in Figure 2.5(b). The problem is expected to be particularly severe with polydisperse systems since small particles in the unstable region of the suspension layer will be more susceptible to re-entrainment. To avoid this problem, they proposed that lamella separators be operated in the cocurrent supercritical mode. This has a
more favourable velocity field (see Figure 2.5(a)). The clear liquid layer adjacent to the clear/suspension layer interface is dragged downwards by the suspension layer due to its higher velocity. However, the main bulk of the clear liquid layer is moving upward at a relatively low velocity which has the positive effect of reducing the internal mixing and the subsequent carry-over of particles into the clear liquid layer.

Poh48 was able to confirm these results experimentally (see Figure 3.2(a)) and consistently obtained higher maximum overflows in the cocurrent supercritical mode. He observed that in the cocurrent subcritical and countercurrent modes an additional cause of particle re-entrainment was interfacial wave disturbance (see Figure 3.2(b)) which resulted in the mixing of the clear and suspension layers. The effect was, however, localised and occurred at the interface only. In the cocurrent supercritical mode the carry over of particles in the supernatant was due to a flow instability which affected the entire thickness of the suspension layer (see Figure 3.2(c) and 3.2(d)). It was initiated by interfacial waves which grew progressively as the suspension flowed down the separator until they reached breaking point whereupon particles were ejected into the clear liquid layer.

Research into the flow instability caused by wave disturbance has been carried out by Leung who concentrated his studies on the supercritical mode of operation. He attempted to characterise the unstable nature of the interface between the clear and suspension layers as a function of settler angle and feed rate. At low settler angles (α < 10° from the horizontal) he found that the instability was associated with an inflectional point in the flow due to shear, whilst at higher angles (10° < α < 60°) it was due to a gravity destabilising mechanism. The high angle case is of relevance to most industrial applications since angles of less than 10° are far below those needed to satisfy the sludge flow constraint. He was able to show both theoretically and experimentally that the critical flowrate at which the interfacial turbulence occurred (Q_{turb}) was dependent on the settler angle, channel spacing and density difference between the clear liquid layer
FIGURE 3.2(a): Re-entrainment of particles into the clear liquid layer due to unfavourable velocity profile (countercurrent flow)

FIGURE 3.2(b): Re-entrainment of particles due to the combined effects of an unfavourable velocity profile and interfacial instability (countercurrent flow)

FIGURE 3.2: INTERFACE WAVE DISTURBANCE
FIGURE 3.2(c): Formation of 'interfacial wave' due to flow instability (cocurrent supercritical mode)

FIGURE 3.2(d): Re-entrainment of particles into the clear liquid layer due to wave breakage brought about by flow instability (cocurrent-supercritical mode)

FIGURE 3.2: INTERFACIAL WAVE DISTURBANCE
and suspension layer. The effect of these factors is discussed below for high angle conditions only. It was found that

i) the difference in specific densities, $\Delta \rho g/\rho$, of the clear and suspension layers leads to a gravitational instability. The longitudinal component of the densimetric gravitational acceleration $(\Delta \rho g/\rho)\sin \alpha$, which is the driving force of the buoyancy flow is the chief cause of the instability, whilst the transverse component $(\Delta \rho g/\rho)\cos \alpha$, is stabilising. Thus it suggests that the critical flowrate $Q_{turb}$ is proportional to $\cot \alpha$.

ii) based on experiment and on a linear stability analysis for a two layer lamella system characterised by gravity waves, it was found that $Q_{turb}$ was proportional to $(\Delta \rho/\rho)^{-0.5}$ which reinforces the argument that the densimetric gravitational acceleration, on the whole, has a destabilising influence.

iii) $Q_{turb}$ varies with the channel spacing as $b^{0.5}$ such that a wider channel has a higher settler efficiency. However it was found that a limiting value of $Q_{turb}$ was reached for channel spacings greater than 10 cm. At large spacings the upper channel wall is so far away from the fluid interface that, so far as the interface is concerned, it can be considered to be at infinity.

Leung also determined the effect of settler variables on the operating efficiency for the subcritical and supercritical modes of operation. Since the maximum throughput, $Q_{\text{max}}$, in the supercritical mode is limited by flow instabilities, then it should vary as $Q_{turb}$ with $\cot \alpha$. However the settler capacity, $Q_{\text{theo}}$, depends on the effective settling area or $\cos \alpha$. Therefore the settler efficiency, defined as $Q_{\text{max}}/Q_{\text{theo}}$, should vary as $(\sin \alpha)^{-1}$. Experimental agreement with this result was excellent. In the subcritical mode there is an additional loss of efficiency due to particle re-entrainment associated with the unfavourable velocity profile that occurs. The re-entrainment becomes more pronounced at lower settler angles because of the increase in settler capacity which results in
the higher velocity of the clear liquid layer. Consequently the rate of increase in efficiency with decreasing angle is significantly less than \((\sin \alpha)^{-1}\).

It also follows that the maximum flowrate in supercritical operation should have the same density dependence as \(Q_{\text{turb}}\) and this is supported by the experimental results. Likewise, it would be expected that the efficiency in the subcritical mode would be less due to particle re-entrainment. Again this was confirmed experimentally, \(Q_{\text{max}}\) decreasing with increasing \(\Delta \rho / \rho\) to the \(-0.9\) power.

Experiments were not carried out to verify the effect of channel spacing. It has, however, been subsequently found\(^{48}\) that increasing the plate spacing from 1.5 to 3.4 cm in the countercurrent mode resulted in an increase in settler efficiency. For a feed concentration of 0.5% v/v the settler efficiency increased from a range of 65-90% to 80-100% and for a feed concentration of 2% v/v from 50-70% to 65-99%.

The effect of settler length on the flow instability was examined by Poh\(^{48}\). It was found that excessively long and narrow channels were susceptible to flow instabilities because any wave disturbance that was generated along the length of the interface had a chance to propagate and amplify to breaking point, ejecting particles into the clear liquid layer. Furthermore, for an ideal system it was shown experimentally that an optimum aspect ratio existed beyond which further increase in the channel length did not produce an increase in the maximum overflow. Three regions of separating efficiencies were identified (see Figure 3.3).

In region I there was excellent agreement between predicted and experimental overflows since, due to the relatively short channel length, the interfacial wave disturbance does not amplify to breaking point. The aspect ratios in this region were defined as \((h/b)_{\text{op}}\). For a plate spacing of 3.4 cm this region extended to settler lengths of up to 0.66m whereas for a plate spacing of 1.5 cm it reduced to under
FIGURE 3.3: THE THREE REGIONS OF SEPARATING EFFICIENCY

0.49m. The beginning of significant particle re-entrainment occurred in region II and hence the maximum separating efficiency was not achieved. In region III no significant increase in maximum overflow occurred with increasing length and hence this represented an upper limit to the separator length, \((h/b)_{ul}\). Given below are the values of \((h/b)_{ul}\) for a plate spacing of 3.4 cm.

<table>
<thead>
<tr>
<th>(c_0) (% v/v)</th>
<th>((h/b)_{ul})</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.5</td>
<td>25-30</td>
</tr>
<tr>
<td>2.0</td>
<td>15-22.5</td>
</tr>
</tbody>
</table>
The above values for aspect ratio are much lower than those used in most existing design methods where h/b is specified as ranging from 40-50. This implies that current industrial separators are operating well outside their optimum range.

In the present research work it is intended to examine real particulate systems to

i) determine whether an optimum and upper limit to the aspect ratio exists in order to verify its use as upper limit for design purposes

ii) determine experimentally the effect of plate spacing on the maximum overflow and settler efficiency

iii) verify the effect of settler angle on the maximum overflow and efficiency

iv) compare the efficiencies and maximum overflows of the countercurrent, cocurrent subcritical and cocurrent supercritical modes of operation

v) use the experimental results to determine the optimum settler configuration for two possible design constraints, that of limited land area and that of limited capital expenditure. Limitations on the land area may occur when a stack of lamella plates is to be inserted into an existing settling basin.

3.3.4 Sludge Flow Constraint

The sludge flow constraint must be satisfied in order to achieve an effective removal of the sludge collected on the lamella plates. If the angle is too low, it will result in poor sludge flow leading to a build-up along the entire length of the lamella plates. In extreme cases the sludge layer may grow to fill the entire channel. On the other hand, if the angle is too high the sludge will travel down the inclined plates at such a high velocity that it will be lifted off the plates and re-entrained into the suspension layer resulting in a reduction in settler efficiency.
It is apparent from the literature that this area of research has been severely neglected. In the current design procedure the practice is to overprovide the angle of inclination. This is undesirable because of the competing interest to obtain the greatest projected area for sedimentation. Hence there is an urgent need to be able to determine the minimum angle at which sludge flow will occur. Another neglected area is in the selection of settler material. The use of a material with a low coefficient of friction should make it possible for settlers to be operated at lower angles and reduce the possibility of particle build-up and adherence to the lamella plates.

In an attempt to improve the current situation, an experimental investigation is planned with the aims of

i) establishing a simple procedure to determine the minimum angle for sludge flow in a continuous separator
ii) establishing the mechanisms of sludge flow and determining the effect of relevant parameters on the mode of sludge flow
iii) examining the effects of different settler materials on the angle required for sludge flow
iv) determining the effects of the mode of sludge flow on the variation in the underflow solids concentration with time in a continuous system and whether the modes of sludge flow which occur in the batch tests can be directly related to the continuous system
v) determining the effect of the mode of settler operation in the continuous system on the variation in the underflow solids concentration with time.

Having achieved these objectives, it will be possible to put forward some useful and reliable design guidelines for the sludge flow constraint in a continuous system.
### 3.4 SIZING METHOD

A significant outcome of the theoretical development by Hill\textsuperscript{18} is that the sedimentation rate in an inclined vessel can be accurately predicted by the Nakamura-Kuroda equation, providing that $\Lambda$ is asymptotically large (i.e., $\Lambda \rightarrow \infty$). This was verified experimentally and theoretically by Acrivos and Herbolzheimer\textsuperscript{19,20}. Poh\textsuperscript{48} carried out a more extensive investigation into the applicability of the Nakamura-Kuroda equation in predicting the operating capacity of a continuous lamella separator. He showed, using an ideal system, that it was capable of accurately predicting the maximum overflow rate provided that:

1) the requirements for achieving the essential steady state conditions are met
2) the dimensions of the separator are carefully selected to minimise the adverse effects of flow instability which leads to particle re-entrainment in the overflow
3) the following conditions are met

\[ \Lambda = O(10^4) - O(10^7) \]
\[ \text{Re} = O(1) \]

The development of the general equation for predicting the overall capacity of the lamella separator will be discussed below. Consider (see Figure 3.4) a continuous separator with feedrate $Q_f$ at solids concentration $c_0$ from which an underflow $Q_u$ of solids concentration $c_u$ and a particle free overflow $Q_0$ are removed. The overall material balance is

\[ Q_f = Q_u + Q_0 \quad (3.18) \]

and the solids balance,

\[ Q_f c_0 = Q_u c_u \quad (3.19) \]
Now, from the Nakamura-Kuroda equation,

\[ Q_0 = \frac{bW}{\sin \alpha} [1 + \frac{h}{b} \cos \alpha] \quad (3.21) \]
Substituting 3.20 into 3.21 gives an equation for the feedrate in terms of separator dimensions and suspension properties. Thus

\[ Q_f = \frac{b v W}{s I n a} \left[ 1 + \frac{h}{b} \cos \alpha \right] \frac{1}{1 - c_o/c_u} \]  

(3.22)

Since for most industrial applications the ratio of the vertical height of the separator to the plate spacing is much greater than 1, equation 3.22 can be simplified to,

\[ Q_f = \frac{b v W}{s I n a} \frac{h}{b} \cos \alpha \frac{1}{1 - c_o/c_u} \]  

(3.23)

Replacing the vertical height of the separator, h, by the term (L\sin \alpha) which incorporates the settler length L gives,

\[ Q_f = \frac{L v W \cos \alpha}{1 - c_o/c_u} \]  

(3.24)

For the two particulate systems studied experimentally, the range of \( \Lambda \) and Re values were:

a) For limestone of size range 5-10 \( \mu m \)

\[ \Lambda = 3.6 \times 10^8 - 3.03 \times 10^9 \]

Re = 9 - 27

b) For limestone of size range 45-53 \( \mu m \)

\[ \Lambda = 2.7 \times 10^6 - 3.70 \times 10^7 \]

Re = 450 - 1700
3.5 APPLICABILITY OF THE CONSTRAINTS IN THE DEVELOPMENT OF THE DESIGN PROCEDURE

The constraints deemed necessary for the successful operation of a lamella separator have been outlined and discussed above. The applicability of these constraints to real particulate systems will be examined experimentally in subsequent chapters using limestone of size ranges 5-10 μm and 45-53 μm, both dispersed in water. The sludge flow constraint will be examined in more detail since this is an area that has been particularly neglected. The accuracy of the Nakamura-Kuroda equation in predicting the maximum overflow will be determined over a range of variables.

A design scheme has already been proposed by Poh\textsuperscript{48} in which the above mentioned constraints were imposed on the relevant design variables in order to suppress the various potential causes of non-idealities. This will be extended and improved with the aid of the findings from the experimental work on real systems.
### CHAPTER 4

**EXPERIMENTAL WORK**

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CHAPTER 4

EXPERIMENTAL WORK

4.1 INTRODUCTION

This chapter describes the experimental equipment and materials used and in addition outlines the experimental procedures followed. The aim of the experimental work is to provide data and information that can be used to improve the design of a lamella separator.

4.2 STUDY OF THE BEHAVIOUR OF SLUDGE FLOW ALONG THE LOWER INCLINED SURFACE OF THE SETTLER

4.2.1 Introduction

The aims of the following experimental work are to establish the sludge transport mechanisms and their dependence on the angle of inclination and materials of construction of the settler in order that a mathematical model of the sludge flow might be developed.

To this end, a qualitative and quantitative study of the sludge flow behaviour of several fully dispersed solid/liquid systems has been carried out in specially constructed batch rigs.

4.2.2 Test Materials

Details of the dispersed systems used are listed in Table 4.1.

The sludge flow behaviour of all of the materials listed in Table 4.1 was tested over an underlying surface material of perspex. In addition, two of the particulate systems, glass beads and limestone, both of size range 90-125 μm and dispersed in distilled water, were examined over surfaces of PVC, aluminium and PTFE. Bronze spheres were also examined over a surface of PVC.
FIGURE 4.1: MICROGRAPHS OF SOLIDS USED IN THE DIFFERENT FULLY DISPERSED SYSTEMS
TABLE 4.1: DETAILS OF DISPERSED SYSTEMS USED IN THE BATCH EXPERIMENTS

<table>
<thead>
<tr>
<th>Dispersed System</th>
<th>Physical Properties of Solids</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Liquid</td>
</tr>
<tr>
<td></td>
<td>Solid</td>
</tr>
<tr>
<td></td>
<td>Size Range</td>
</tr>
<tr>
<td></td>
<td>Specific Gravity</td>
</tr>
<tr>
<td></td>
<td>Shape</td>
</tr>
<tr>
<td>Distilled water</td>
<td>Glass beads</td>
</tr>
<tr>
<td></td>
<td>45- 53 μm</td>
</tr>
<tr>
<td></td>
<td>2.46</td>
</tr>
<tr>
<td></td>
<td>Spherical</td>
</tr>
<tr>
<td>&quot;</td>
<td>&quot;</td>
</tr>
<tr>
<td>&quot;</td>
<td>63- 75 μm</td>
</tr>
<tr>
<td>&quot;</td>
<td>&quot;</td>
</tr>
<tr>
<td>&quot;</td>
<td>90-125 μm</td>
</tr>
<tr>
<td>&quot;</td>
<td>&quot;</td>
</tr>
<tr>
<td>&quot;</td>
<td>150-180 μm</td>
</tr>
<tr>
<td>&quot;</td>
<td>&quot;</td>
</tr>
<tr>
<td>&quot;</td>
<td>200-250 μm</td>
</tr>
<tr>
<td>&quot;</td>
<td>&quot;</td>
</tr>
<tr>
<td>&quot;</td>
<td>355-420 μm</td>
</tr>
<tr>
<td>&quot;</td>
<td>&quot;</td>
</tr>
<tr>
<td>&quot;</td>
<td>Bronze spheres</td>
</tr>
<tr>
<td></td>
<td>90-125 μm</td>
</tr>
<tr>
<td></td>
<td>7.70</td>
</tr>
<tr>
<td>&quot;</td>
<td>Powdered glass</td>
</tr>
<tr>
<td></td>
<td>90-125 μm</td>
</tr>
<tr>
<td></td>
<td>2.21</td>
</tr>
<tr>
<td></td>
<td>Irregular, angular fragments</td>
</tr>
<tr>
<td>&quot;</td>
<td>Limestone</td>
</tr>
<tr>
<td></td>
<td>5- 10 μm</td>
</tr>
<tr>
<td></td>
<td>2.69</td>
</tr>
<tr>
<td></td>
<td>Irregular, granular</td>
</tr>
<tr>
<td>&quot;</td>
<td>&quot;</td>
</tr>
<tr>
<td>&quot;</td>
<td>45- 53 μm</td>
</tr>
<tr>
<td>&quot;</td>
<td>&quot;</td>
</tr>
<tr>
<td>&quot;</td>
<td>90-125 μm</td>
</tr>
<tr>
<td>&quot;</td>
<td>&quot;</td>
</tr>
<tr>
<td>&quot;</td>
<td>Ziron</td>
</tr>
<tr>
<td></td>
<td>&quot;</td>
</tr>
<tr>
<td></td>
<td>4.22</td>
</tr>
<tr>
<td>&quot;</td>
<td>&quot;</td>
</tr>
<tr>
<td>Reofos 65/</td>
<td>Glass beads</td>
</tr>
<tr>
<td></td>
<td>&quot;</td>
</tr>
<tr>
<td>&quot;</td>
<td>2.46</td>
</tr>
<tr>
<td>&quot;</td>
<td>Spherical</td>
</tr>
<tr>
<td>Reomol DBP</td>
<td>&quot;</td>
</tr>
</tbody>
</table>
The solids listed above were all incompressible particulates by nature and were sized using standard sieves.

The dispersed systems selected were designed to study the following parameters which were postulated to have significant effects on the sludge flow behaviour:

i) Shape and surface texture of solids (see micrographs in Figure 4.1);
ii) Density difference between solids and liquids;
iii) Size of solids;
iv) Roughness of surface material.

Water was used predominantly because of its relevance to most industrial applications. However a study was also made of the effects of liquid viscosity and surface tension. Nonidet was added to the distilled water in order to investigate the effects of surface tension, the viscosity of water not being significantly affected by the addition of Nonidet. The effect of viscosity was determined by comparing the sludge flow behaviour of glass spheres in a plasticisers mixture of Reofos 65/Reomol DBP to that in distilled water to which sufficient Nonidet had been added in order that the surface tensions of the two liquids were identical. The physical properties of the liquids at 25°C are given below:

**TABLE 4.2: PHYSICAL PROPERTIES OF THE LIQUIDS EMPLOYED IN THE BATCH EXPERIMENTS**

<table>
<thead>
<tr>
<th>Liquid</th>
<th>Density kg/m³</th>
<th>Viscosity cp</th>
<th>Surface Tension Dynes/cm²</th>
</tr>
</thead>
<tbody>
<tr>
<td>Distilled water</td>
<td>997.4</td>
<td>0.896</td>
<td>70.3</td>
</tr>
<tr>
<td>Distilled water + 1 drop Nonidet</td>
<td>&quot;</td>
<td>&quot;</td>
<td>43.9</td>
</tr>
<tr>
<td>Distilled water + 2 drops Nonidet</td>
<td>&quot;</td>
<td>&quot;</td>
<td>39.0</td>
</tr>
<tr>
<td>Distilled water + 3 drops Nonidet</td>
<td>&quot;</td>
<td>&quot;</td>
<td>36.1</td>
</tr>
<tr>
<td>Distilled water + 4 drops Nonidet</td>
<td>&quot;</td>
<td>&quot;</td>
<td>34.0</td>
</tr>
<tr>
<td>Reofos 65/Reomol DBP</td>
<td>1078</td>
<td>22.253</td>
<td>36.6</td>
</tr>
</tbody>
</table>
All the experiments were conducted in a constant temperature room maintained at 25°C ± 0.2°C.

4.2.3 Experimental Equipment

Two vessels were used to perform the sludge flow experiments. In the first it was only possible to observe the flow of sludge over a perspex surface whereas the second enabled the sludge flow over various surfaces to be studied. Details of the two vessels are given below.

4.2.3.1 Vessel with perspex base surface

The vessel was rectangular of internal dimensions 5 cm x 5 cm x 50 cm. To ensure that the visual examination of the sludge flow behaviour was possible, the vessel was constructed entirely of perspex as shown in Figure 4.2. The top 5 cm of the upper surface of the vessel was left open for the test materials to be added. The vessel was attached to a perspex base plate which was mounted on an optical stand. The base plate was constructed of perspex to enable the sludge flow to be viewed from below. The angle of inclination could be varied by means of the optical stand. In addition a fine control of the angle of inclination was provided in the form of a fine adjustment rod which was connected to the base plate. The angle of inclination was measured using a plumb line and protractor. A stirrer was used to create a homogeneous suspension within the vessel which, on settling, formed a uniform layer of sludge along the length of the lower inclined surface except for a short distance of about 10 cm from its base. This particle free area was achieved by positioning a stop about 10 cm from the end of the settler. Its function was to provide a free surface for sludge flow at the lower end of the tank.

4.2.3.2 Vessel with optional base material

The main body of the vessel was rectangular of internal dimensions 5 cm x 5 cm x 50 cm. A hopper was attached to the lower end of the
FIGURE 4.2: EXPERIMENTAL RIG WITH PERSPEX BASE FOR THE STUDY OF SLUDGE FLOW BEHAVIOUR
vessel (see Figure 4.3). This provided a free edge over which the sludge could flow and was an alternative to the stop employed in the vessel described previously. In addition, at the end of an experimental run the sludge could be easily removed from the base of the hopper by means of a valve.

The vessel was constructed of perspex to enable the sludge flow to be observed. The top plate of the vessel was removable and connected to the main body by means of a flange and gasket (as illustrated in Figure 4.3). Thus, by removing the top plate, sheets of different materials could be placed over the lower surface of the settler. This enabled a study of the effect of underlying surface material on sludge flow to be carried out. A stirrer was used as before to create an homogeneous suspension and a protractor and plumb line used to measure the angle of inclination.
FIGURE 4.3: EXPERIMENTAL RIG WITH OPTIONAL BASE MATERIAL TO STUDY SLUDGE FLOW BEHAVIOUR
4.2.4 Experimental Procedure

A summary of the steps in which the experiments were conducted is listed below:

4.2.4.1 Vessel with perspex base surface

i) The vessel was first filled with the desired quantity of liquid
ii) The stopper was then positioned at 10 cm from the base end of the settler before the desired quantity of solids was introduced
iii) Using the stirrer the entire mixture was agitated in order to create an homogeneous suspension
iv) The solids were then allowed to settle to form a sludge layer on the lower inclined surface. Care was taken to ensure that an even layer of sludge was formed over the entire length
v) The stopper and stirrer were then moved to the lower end of the vessel in order to create a free surface for sludge flow
vi) The vessel was then gradually and gently inclined until sludge movement first occurred. The nature of the sludge movement and the angle at which it occurred were recorded in detail
vii) After the initial sludge movement had ceased, step (vi), was repeated until the entire sludge layer was removed
viii) For each particulate system studied the entire procedure detailed above was repeated six times to obtain acceptable average results.

4.2.4.2 Vessel with optional base material

i) The vessel was filled with the desired quantity of liquid
ii) The required amount of solids was then introduced
iii) Using a stirrer, the entire mixture was agitated to produce an homogeneous suspension
iv) The solids in the suspension were then allowed to settle out to form a sludge layer over the entire inclined surface
v) Steps (vi) and (vii) of the previous procedure (Section 4.2.4.1) were then carried out
vi) The solids were removed from the hopper through the outlet at its base
vii) The solids were then reintroduced back into the vessel and steps (iii) to (vi) repeated six times to obtain acceptable average results.

4.3 EXPERIMENTAL STUDY OF THE OPERATION OF A CONTINUOUS LAMELLA SEPARATOR

The design, construction and operation of the rig used to examine continuous sedimentation is discussed below.

The aims of the following experimental work are firstly to determine the maximum overflows that can be achieved in continuous sedimentation over a range of settler variables and to compare with theoretical predictions, secondly to determine the minimum angle of operation of the settler and thirdly to examine the continuous flow of sludge down the lower inclined plane.

4.3.1 Test Materials

Two particulate systems were examined: limestone of size range 45-53 μm dispersed in water and limestone of size range 5-10 μm dispersed in water.

4.3.2 Lamella Separator

The continuous settling tests were performed in the lamella separator shown in Figure 4.4. It was constructed from clear perspex and consisted essentially of a main separator body, a clear liquid removal chamber and a sludge collector.
The entire separator was designed in modular form so that a wide range of settling channel aspect ratios (0-75) could be obtained with channel spacings of 1.5-8.0 cm and at angles of inclination of 0-90°. The internal dimensions of the rectangular modules that formed the main body of the separator are given below in Table 4.3:

TABLE 4.3: INTERNAL DIMENSIONS OF THE MAIN BODY OF THE SEPARATOR

<table>
<thead>
<tr>
<th>No. of rectangular modules available</th>
<th>Internal dimensions of module (cm)</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Length</td>
<td>Width</td>
</tr>
<tr>
<td>1</td>
<td>17</td>
<td>4</td>
</tr>
<tr>
<td>2</td>
<td>46</td>
<td>4</td>
</tr>
<tr>
<td>1</td>
<td>17</td>
<td>4</td>
</tr>
<tr>
<td>1</td>
<td>46</td>
<td>4</td>
</tr>
</tbody>
</table>

Each of the rectangular modules with a 3.4 cm channel spacing was designed so that a perspex spacer plate could be inserted to divide it into two compartments of equal dimensions. In this way the channel spacing could be changed from 1.5 to 3.4 cm. By using different modules for the feed and clear liquid overflow section, the separator could be adapted to operate in both cocurrent and countercurrent modes.

In cocurrent operation the feed was introduced at the top of the settling channel through a rectangular inlet whose depth could be varied by an adjustable perspex plate. This made it possible to test whether both the supercritical and subcritical modes of operation could be achieved and, this being the case, to study both these operating modes. Inserts could be placed in the inlet section between the adjustable plate and the lower surface of the settler to help maintain a uniform flow of feed.
In countercurrent operation the feed was introduced into the settler via inlets on its vertical sides near to the bottom of the separator in such a manner that the solids already settled on the lower inclined surface were not disturbed.

A magnetic stirrer was installed in the sludge collector to facilitate the removal of solids in the underflow system. The entire separator was supported between two rigid stands and could be positioned at any angle between the vertical and horizontal. A plumb line and protractor enabled the angle of inclination to be determined.

4.3.3 Continuous Flow Arrangement
The experimental rig used to study continuous inclined sedimentation is illustrated in Figure 4.5. Figure 4.6 shows diagrammatically the flow arrangement used in the continuous inclined sedimentation experiments.

A feed tank was provided in which an homogeneous suspension was created using a blade paddle stirrer. The latter was sufficiently powerful to remix the suspension even when the particles had settled out. The temperature in the feed tank was maintained at $25^\circ C \pm 0.1^\circ C$ by a thermostatically controlled heater used in conjunction with a cooler.

In a normal operation the feed was introduced into the separator from the feed tank. The overflow and underflow were recycled back to the feed tank via a mixing vessel. The reason for combining the overflow and underflow streams was to create higher velocities in the return tubing. This prevented any particles settling out on the way back up to the feed tank. A peristaltic pump (Heidolph Type SP) was used to pump the combined overflow/underflow mixture back to the feed tank. It was run at a constant speed which was greater than the total inflow into the mixing tank in order to prevent any accumulation therein. This did not hinder the mixing of the two streams but allowed an air/sludge mixture to be pumped up to the feed tank. The return pipe was carefully positioned in the feed tank to avoid entrainment of air into the system.
FIGURE 4.5: EXPERIMENTAL RIG FOR CONTINUOUS LAMELLA SEPARATOR
FIGURE 4.6: CONTINUOUS FLOW ARRANGEMENT
The underflow was pumped from the settler to the mixing tank by means of a peristaltic pump (Watson Marlow Type 601) whose capacity could be varied both by adjusting its pump speed and by using tubing of an appropriate size. The overflow flow rate was controlled by valve V3 and was measured by an Alcon rotameter. It was not necessary to provide a pump on the overflow stream because the head from the feed tank was sufficient to provide an adequate flow rate into the mixing tank. An additional benefit from this arrangement was that the overflow flow rate remained steady and was easy to control.

4.3.4 Sampling and Sample Measurements

Samples of feed, overflow and underflow were taken at points SP1, SP2 and SP3 respectively for the determination of particle size distribution and/or solids concentration.

i) Particle size measurement

A Malvern Particle Sizer was used to determine the particle size distribution of the feed and overflow. The accuracy of this instrument in determining the size distribution was checked by comparing its results to those attained by the Andreasan technique. For a particular percentage undersize the particle size given by the Andreasan Technique was between 9-16\% less than the value predicted by the Malvern Particle Sizer over the range 5-10 \(\mu m\). This was considered an acceptable discrepancy.

ii) Solids concentration

The underflow and feed concentration were determined by vacuum filtering a sample of approximately 10 cc volume. The filter was a cellulose nitrate membrane of pore size 0.8 \(\mu m\). The residue was dried in an oven for about 15 minutes then allowed to cool in a dessicator. The dried sample was weighed and the mass of solid determined. The volume concentration could now be computed from the respective masses of solid and water in the sample and from their respective densities.
Due to the low concentration of solids in the overflow the concentration in this stream could not be directly measured. The solution turbidity was determined instead using a Hach Turbidimeter. Solution turbidity has a linear relationship with concentration.

4.3.5 Experimental Procedures
4.3.5.1 Start-up

Either of two experimental procedures were followed depending on whether the experiment was a continuation of a previous run (already have solids in the feed tank) or a new run with fresh solids (see Figure 4.5).

4.3.5.1.1 From fresh material

1. The tank was filled with suspension liquid
2. Valve V2 was opened to let air out of the system
3. Valve V1 was then opened slowly to allow water to flow slowly into the separator
4. The underflow pump was activated for a short period to remove air from the underflow pipework
5. Valve V1 was fully opened and the separator filled with water
6. Valve V2 was closed when the separator was full
7. Valve V3 was opened fully in order to flush the air out of the overflow pipework
8. Valves V1 and V3 were then closed and any liquid in the mixing tank was pumped into the feed tank
9. The liquid in the feed tank was topped up to 10 litres with more suspension liquid
10. The required quantity of solids were then added to the feed tank to make up the desired concentration of suspension
11. The blade paddle stirrer, heater and cooler were then switched on and the contents of the feed tank allowed to reach 25°C
12. Valve V1 was then fully opened
13. The peristaltic pumps were switched on and set to the required flowrates
14. Valve V3, which controls the overflow, was gradually opened until the desired flowrate was obtained. This point was taken as the beginning of a run.

4.3.5.1.2 From suspension already in the feed tank

1. Fresh suspension liquid was gravity fed from a holding tank into the settler through the inlet below V1
2. Steps 6 to 11 were then followed as before
3. The concentration in the feed tank was checked. If necessary solids were added to make up the desired concentration of suspension
4. Steps 12 to 14 above were then followed.

4.3.5.2 Analysis of settler operation

Once start-up of the settler was completed the following analyses of settler operation were carried out.

4.3.5.2.1 Maximum overflow flowrate determination

The settler was allowed to run at the desired flowrate until steady-state was reached. Steady-state was taken to be the point at which the overflow turbidity and size distribution remained constant over a period of half an hour. A good estimate of the time at which the turbidity started to increase could be calculated from the throughput time (settler volume/overflow volume flowrate). The overflow was then tested to determine whether the settler was being operated at its maximum overflow flowrate.

The maximum overflow rate of a lamella separator is taken as the point at which solids first appear in the overflow as the feed flowrate is increased. If the feed is of a narrow size distribution the maximum overflow can be easily detected by a sudden increase in the turbidity of the overflow as the feedrate is gradually increased. However this is not the case for the particulate systems studied here: limestone
of size ranges 45-53 \( \mu m \) and 5-10 \( \mu m \). The limestone was sized using an Alpine Classifier. Even though the limestone was passed through the classifier numerous times it was not possible to totally remove the fines. Thus, the fines had to be taken into account when determining the maximum overflow. The methods used are discussed below.

i) **For limestone of size range 45-53 \( \mu m \)**

When the limestone was introduced into the continuous system and the system allowed to run for several hours, it was found that the proportion of fines present increased. This was most likely due to the grinding action of the peristaltic pump and to impact with the stirrer paddle. The percentage of fines was subsequently monitored over several days and was found to reach a constant value. A sampling method was then developed to take account of the fines present. A sample from the feed was taken at the same time as the turbidity of the overflow was measured. The time for the limestone of size range 45-53 \( \mu m \) to settle out of the feed sample was calculated using Stokes Law and the feed sample was allowed to stand for this calculated period of time. A sample of the supernatant was then pipetted off and its turbidity determined. This was called the background turbidity and was caused by the presence of fines. The difference between the overflow turbidity and the background turbidity was known as the excess turbidity. The excess turbidity will give a measure of the concentration of particles in size range 45-53 \( \mu m \) present in the overflow. It was determined over a range of overflow rates and was found to undergo a sudden increase as the flowrate was increased above a particular value. This was taken as the point of maximum overflow.

ii) **For limestone of size range 5-10 \( \mu m \)**

The percentage of fines initially decreased when the limestone was introduced into the settler. After a period of several weeks the concentration of fines remained constant. It was not practical to adopt the same method used for the 45-53 \( \mu m \) sized limestone because, due to the low settling velocity of the limestone, it would be necessary to wait several hours before determining the background and excess turbidity. Instead, the size range of the overflow was
monitored using a Malvern Size Analyser over a range of flowrates. As the overflow rate was increased the percentage of larger particles increased. This made it possible to determine the maximum overflow from the size analysis. The maximum overflow was subsequently taken as the point at which 10% of the particles in the overflow were greater than 5 μm. A value of 10% was taken in order to obtain consistent results because it was thought that, at percentages less than this, the particle size predicted was unreliable. Turbidity measurements were again taken in order to ascertain whether or not steady state had been reached.

4.3.5.2.2 Continuous operation of settler at maximum overflow rate

Subsequent to determining the maximum overflow flowrate all the solids in the settler were returned to the feed tank and the system restarted as before. The overflow was set at its maximum value. The underflow peristaltic pump was switched on and the pump speed adjusted until the underflow flowrate equalled the overflow flowrate in the case of 5-10 μm limestone and until the underflow rate was a third of the overflow rate in the case of 45-53 μm limestone. The settler was then kept running for about 4-5 hours. At intervals of between 15 to 30 minutes samples were taken from sampling points SP1, SP2 and SP3 and the feed and underflow concentration and overflow turbidity and size distribution determined.

4.3.5.2.3 Determination of minimum angle for sludge flow

This was determined at each settler length when steady state conditions had been achieved. The angle of inclination of the settler was gradually reduced until sludge flow ceased. This was determined by observation of the sludge flow and by monitoring the underflow concentration.
4.3.5.3 Shut-down

When a run was finished the following procedure took place:

1. Valve V1 was closed and the stirrer, heater and cooler switched off
2. The suspension in the separator was pumped by peristaltic pump into the mixing tank
3. The suspension in the feed tank was allowed to settle until enough clear liquid could be removed such that all the suspension could be contained in the feed tank.
## CHAPTER 5

**BEHAVIOUR OF SLUDGE FLOW ON LOWER INCLINED SURFACE**

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CHAPTER 5

BEHAVIOUR OF SLUDGE FLOW ON LOWER INCLINED SURFACE

5.1 INTRODUCTION

The following section discusses the results of the batch sludge flow tests. It identifies the sludge transport mechanisms and their dependence on the angle of inclination and materials of construction. The effect of particle, liquid and settler material parameters on the angle at which the predominant mode of sludge flow occurs are examined. This section aims to provide a basis for a mathematical model to predict the minimum angle of operation of a continuous lamella separator.

5.2 MECHANISMS OF SLUDGE FLOW

From the batch scale experiments it was found that the sludge flow was not continuous but occurred via a series of intermittent movements as the angle of inclination was progressively increased. The type of sludge flow movement that occurred was dependent on the thickness of the sludge layer, the nature of the solids, the inclination angle and the type of material used on the lower inclined surface.

Four distinct modes of sludge flow were identified and these were classified as layer movement, heap movement, fracture movement and bulk movement. The individual characteristics of each mode of flow are described below.

5.2.1 Layer Movement

This type of sludge flow entails the sliding of the top layers of solids over a relatively thin and almost stationary underlying layer (see Figure 5.1) which is typically 1-2 particle diameter(s) in thickness. By analogy, its gross behaviour resembles the flow of a
viscous fluid down an inclined surface. In addition the movement of the top layers appears to exhibit a flat vertical velocity profile (i.e. plug flow behaviour).

5.2.2 Heap Movement

Heap movement occurs when sludge flow takes place by the motion of small aggregates of solids over the entire inclined surface. Its appearance from a plan view is clearly illustrated in Figure 5.2.

5.2.3 Bulk Movement

In contrast to layer movement the entire thickness of the sludge moves en masse (as shown in Figure 5.3). Its flow behaviour resembles the sliding motion of a block down an inclined plate.

5.2.4 Fracture Flow

The prominent feature of this type of flow is that extensive fractures (see Figure 5.4) are formed within the sludge layer. Below these the sludge flows in the heap and/or layer modes of movement. Fractures are continuously being formed above the initial shear region. Its behaviour resembles that of powder flow. This mode of flow was only observed with limestone in the size range 5-10 μm.

Thus from a design point of view when sizing a lamella settler there are four possible flow conditions to consider which will give different rates of discharge.

Having described the various modes of sludge flow, reference is now made to Tables 5.1 and 5.2 which outline the overall sludge flow behaviour exhibited by each of the fully dispersed systems that have been tested. As can be seen the sequence of sludge movements that occur are clearly dependent on the thickness of the sludge layer and the nature of both the solids comprising the sludge and the surface over which the sludge is flowing.
<table>
<thead>
<tr>
<th>Dispersed System</th>
<th>Size Range of Solids (µm)</th>
<th>Sequence of Intermittent Sludge Layer Movements (with Progressive Increase in α)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Water</td>
<td>Glass beads 355-420</td>
<td>Bulk Movement → Heap Movement → Bulk Movement</td>
</tr>
<tr>
<td>Water</td>
<td>Glass beads 90-125</td>
<td>Heap Bulk Movement → Localised Bulk Movement → Heap Layer Movement → Bulk Movement</td>
</tr>
<tr>
<td>Water</td>
<td>Bronze spheres 90-125</td>
<td>Localised Heap Bulk Movement → Localised Bulk Layer Movement → Bulk Movement</td>
</tr>
<tr>
<td>Water</td>
<td>Powdered glass 90-125</td>
<td>Heap Movement → Localised Layer Movement → Heap Movement</td>
</tr>
<tr>
<td>Water</td>
<td>Zircon 90-125</td>
<td>Localised Layer Movement → Heap Movement</td>
</tr>
<tr>
<td>Water</td>
<td>Lime-stone 90-125</td>
<td>Localised Layer Movement → Heap Movement</td>
</tr>
<tr>
<td>Renosil 65/Reomol 0BP</td>
<td>Glass beads 90-125</td>
<td>Heap Movement → Bulk Movement → Layer Movement → Bulk Movement</td>
</tr>
<tr>
<td>Water</td>
<td>Lime-stone 45-53</td>
<td>Heap Movement → Layer Movement → Heap Movement</td>
</tr>
<tr>
<td>Water</td>
<td>Lime-stone 5-10</td>
<td>Heap Movement → Localised Layer Movement → Heap Movement → Fracture Movement → Heap Movement</td>
</tr>
<tr>
<td>Water</td>
<td>Glass beads 45-53</td>
<td>Heap Movement → Layer Movement → Heap Movement</td>
</tr>
</tbody>
</table>
### TABLE 5.2
SLUDGE FLOW BEHAVIOUR OVER DIFFERENT PLATE MATERIALS

<table>
<thead>
<tr>
<th>Underlying Surface Material</th>
<th>Solids</th>
<th>Sequence of intermittent sludge layer movements (with progressive increase in $\alpha$)</th>
<th>Initial concentration of the suspension used to form the sludge layer, $c_0$</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td>0.1% v/v</td>
</tr>
<tr>
<td>PTFE</td>
<td>Glass beads</td>
<td>Heap Movement</td>
<td>Layer Movement</td>
</tr>
<tr>
<td>PTFE</td>
<td>Ground glass</td>
<td>Heap Movement</td>
<td>Layer Movement</td>
</tr>
<tr>
<td>PVC</td>
<td>Glass beads</td>
<td>Heap Movement</td>
<td>Layer Movement</td>
</tr>
<tr>
<td>PVC</td>
<td>Ground glass</td>
<td>Heap Movement</td>
<td>Layer Movement</td>
</tr>
<tr>
<td>Aluminium</td>
<td>Glass beads</td>
<td>Heap Movement</td>
<td>Layer Movement</td>
</tr>
<tr>
<td>Aluminium</td>
<td>Ground glass</td>
<td>Heap Movement</td>
<td>Layer Movement</td>
</tr>
</tbody>
</table>
In some cases comparisons can be drawn between particulate systems in which two parameters occur which have opposing effects on the mode of sludge flow. Thus in addition to identifying the overall parameters dictating the mode of sludge flow, it is possible to obtain an indication of their relative influences, which are discussed below.

A comparison can be made of the relative effects of plate surface roughness to particle roughness. Bulk movement occurs in a system of ground glass (size range 90-125 μm) dispersed in distilled water when flowing over a surface of PTFE but not over surfaces of PVC, aluminium and perspex. As the particles are irregular, it would be expected that only heap and layer movement would occur. However PTFE has a low coefficient of friction which promotes bulk flow and this overrides the opposing effect of particle roughness. Glass spheres (size range 90-125 μm) dispersed in distilled water, however, exhibit bulk movement when flowing over surfaces of PTFE and perspex but not over PVC and aluminium. Hence the regularity in shape of the glass spheres is sufficient to permit bulk flow over perspex and PTFE but the roughness of aluminium and PVC override this tendency.

Density appears to have a strong effect on the mode of flow. Bronze spheres of size range 90-125 μm dispersed in distilled water and flowing over a surface of PVC will exhibit bulk flow whereas glass beads of the same size range will not. Thus the increased density is a more dominant effect than the relatively high roughness of the surface material.

The effect of reducing the particle size of spherical particles can be seen with the system of glass beads dispersed in distilled water flowing over a surface of perspex. Bulk flow occurred with size ranges 90-125 μm, 150-180 μm, 200-250 μm, and 355-420 μm but not with the lower size ranges 45-53 μm and 63-75 μm. Thus the tendency of spherical particles to exhibit bulk flow is overcome when the particle size is reduced sufficiently.
The effect which the above mentioned parameters have on the modes of sludge flow that occur is summarised in Table 5.3 below. For example, increasing the particle roughness will eventually result in bulk movement no longer taking place, only layer and heap movement.

### TABLE 5.3: TO ILLUSTRATE THE EFFECT OF CERTAIN PARAMETERS ON THE MODES OF SLUDGE FLOW WHICH OCCUR IN A PARTICULATE SYSTEM

<table>
<thead>
<tr>
<th>Fracture</th>
<th>Layer</th>
<th>Heap</th>
<th>Layer</th>
<th>Heap</th>
<th>Bulk</th>
</tr>
</thead>
<tbody>
<tr>
<td>→ Layer</td>
<td>→ Heap</td>
<td>→ Heap</td>
<td>→ Layer</td>
<td>→ Heap</td>
<td>→ Bulk</td>
</tr>
</tbody>
</table>

- Increasing particle roughness
- Increasing particle sphericity
- Increasing coefficient of friction — surface material
- Increasing density difference between solid and liquid
- Increasing particle size

The influence of the sludge layer thickness has been omitted from the above. This is because its sole effect appears to be in determining whether or not the sludge exhibits layer movement. As the thickness of the sludge layer is decreased a point is reached at which only heap and/or bulk movement will take place.

It should be noted that the initial sludge layer thickness is expressed in terms of the number of layers of particles via \( c_0 \) - the initial concentration of the suspension that is used to create the sludge layer. The estimated thicknesses of the initial sludge layer are given below for the size ranges of solids which were used in the experiments over the range of initial concentrations used. In determining these values two assumptions were made. Firstly, a porosity of 0.5 for the sludge layer was assumed and secondly a mean value of particle size was taken for each size range.
TABLE 5.4: ESTIMATED NUMBER OF PARTICLE LAYERS FORMED WITH EACH INITIAL SUSPENSION CONCENTRATION

<table>
<thead>
<tr>
<th>Initial Concentration of Suspension $c_0$ (% v/v)</th>
<th>Estimated sludge layer thickness (Number of particle layers)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Size range μm</td>
</tr>
<tr>
<td></td>
<td>5-10</td>
</tr>
<tr>
<td>0.1</td>
<td>10-11</td>
</tr>
<tr>
<td>0.2</td>
<td>21</td>
</tr>
<tr>
<td>0.3</td>
<td>32</td>
</tr>
<tr>
<td>0.4</td>
<td>42</td>
</tr>
<tr>
<td>0.5</td>
<td>53</td>
</tr>
<tr>
<td>1.0</td>
<td>106</td>
</tr>
<tr>
<td>1.5</td>
<td>160</td>
</tr>
</tbody>
</table>

For most practical applications the sludge layer is usually of the order of a few particle diameters in thickness. Referring to Tables 5.1 and 5.2 it can be seen that, for such sludges, layer movement will be the first particle motion to occur. Then, either at the same angle to the horizontal or as the angle is increased, heap and/or bulk movement takes place. Thus it would be expected that if the separator were operated at low angles to the horizontal then layer movement would be the predominant mode of sludge transport, whereas at steeper angles bulk and/or heap movement would take precedence. When the settler is operated in the cocurrent mode the particles entering in the feed stream already possess a momentum down the slope. This momentum is likely to make it possible to achieve the different modes of flow at lower angles than is possible in the countercurrent mode and as such should be taken into account in the settler design. In addition, it should enable the settler to be operated at lower angles in the cocurrent mode than can be achieved in the countercurrent mode.
5.3  **STUDY OF THE PARAMETERS AFFECTING THE ANGLE AT WHICH LAYER MOVEMENT OCCURS**

In addition to examining the various modes of flow a study was also made of the angles at which layer movement starts to occur. This will be discussed in detail below.

5.3.1  **Settler Material**

The effect of surface roughness on the angle of inclination required for layer movement was examined using four settler materials: perspex, PTFE, PVC and aluminium. These were studied in conjunction with the particulate systems of glass beads (size range 90-125 μm) and ground glass (size range 90-125 μm) which were both dispersed in distilled water. Bronze spheres dispersed in distilled water were studied over PVC and perspex. The experimental results are shown in Tables 5.5, 5.6 and 5.7 and Figures 5.5, 5.6 and 5.7.

**TABLE 5.5: EFFECT OF UNDERLYING SURFACE MATERIAL ON THE REQUIRED ANGLE FOR LAYER MOVEMENT**

<table>
<thead>
<tr>
<th>Initial concentration of the suspension* that is used to form the sludge layer $c_0$ (% v/v)</th>
<th>Required angle of inclination for layer movement, $\alpha^\circ$</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>PVC</td>
</tr>
<tr>
<td>0.5</td>
<td>38$^\circ$</td>
</tr>
<tr>
<td>1.0</td>
<td>36$^\circ$</td>
</tr>
<tr>
<td>1.5</td>
<td>33$^\circ$</td>
</tr>
</tbody>
</table>

* Ground glass of size range 90-125 μm dispersed in distilled water
Effect of underlying surface material on the required angle for layer movement

Ground glass (90-125μm) in distilled water

![Graph showing the effect of initial concentration of suspension on the required angle for layer movement for different surface materials, including PVC, Aluminium, PTFE, and Paraffin.](image)

FIGURE 5.5
It can be seen from the tables above that depending on the type of surface material used, the angle at which layer movement occurs can be reduced by as much as 16°. Thus careful selection of the settler material would make it possible to operate at lower angles. This could increase the total settling area available without incurring increased capital costs for additional plates provided that the surface material was not more expensive. For example:
Effect of underlying surface material on the required angle for layer movement

Glass beads (90–125 μm) in distilled water

Figure 5.6
Effect of underlying surface material on the required angle for layer movement

Bronze spheres (90–125μm) in distilled water

FIGURE 5.7
TABLE 5.8: PERCENTAGE INCREASE IN SETTLING AREA BY REDUCTION IN INCLINATION ANGLE TO HORIZONTAL

<table>
<thead>
<tr>
<th>Original Angle</th>
<th>Reduced by 13° to</th>
<th>% Increase in Settling Area</th>
</tr>
</thead>
<tbody>
<tr>
<td>35°</td>
<td>22°</td>
<td>13%</td>
</tr>
<tr>
<td>45°</td>
<td>32°</td>
<td>20%</td>
</tr>
<tr>
<td>55°</td>
<td>42°</td>
<td>30%</td>
</tr>
</tbody>
</table>

Clearly, the roughness of the surface material is a very important parameter and hence a mathematical model describing the sludge flow must take this factor into account.

5.3.2 Shape and Surface Texture of Solids

The effect of shape and surface texture of the solids is shown in Table 5.9, Figure 5.8 and by a comparison of Tables 5.5 and 5.6. All the solids have been sized to 90-125 μm using standard sieves and were dispersed in distilled water.

TABLE 5.9: EFFECT OF PARTICULATE SHAPE AND SURFACE TEXTURE ON THE REQUIRED ANGLE FOR LAYER MOVEMENT

<table>
<thead>
<tr>
<th>Initial Concentration of the suspension used to form the sludge layer $c_0$ (% v/v)</th>
<th>Required angle of inclination for layer movement*, $\alpha^0$</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Glass*</td>
</tr>
<tr>
<td></td>
<td>Beads</td>
</tr>
<tr>
<td>0.5</td>
<td>17°</td>
</tr>
<tr>
<td>1.0</td>
<td>17°</td>
</tr>
<tr>
<td>1.5</td>
<td>17°</td>
</tr>
</tbody>
</table>

* size of solids = 90-125 μm
+ underlying surface is perspex.
Effect of particulate shape and surface texture on the required angle for layer movement

Particle size: 90–125 μm
Underlying surface material: Perspex

FIGURE 5.8
The significant difference in the required angle of inclination between the irregularly shaped solids (zircon, powdered glass and limestone) and the spherical beads suggests that the surface texture and shape are important parameters. A good indication of the influence of this parameter can be obtained by comparing the results for the systems of glass beads and ground glass. For surface materials of aluminium and PVC the difference in required inclination angle of the two particulate systems is about 15° whilst for PTFE and perspex (which have a lower coefficient of frictional resistance than aluminium and PVC) the difference in angle reduces to under 10°. Furthermore reducing the roughness of the surface material appears to have a greater effect on the irregular particles.

5.3.3 Particle Size

The relevance of solids size as a parameter governing the layer movement is shown in Figures 5.9 and 5.10 and Tables 5.10 and 5.11. Six size ranges of glass beads and three of limestone have been studied. These are: for glass beads 45-53 µm, 63-75 µm, 90-125 µm, 150-180 µm, 200-250 µm and 355-420 µm and for limestone 5-10 µm, 45-53 µm and 90-125 µm.

### TABLE 5.10: EFFECT OF PARTICLE SIZE ON THE REQUIRED ANGLE OF INCLINATION FOR LAYER MOVEMENT

<table>
<thead>
<tr>
<th>Initial concentration of the suspension* used to form the sludge layer C₀ (% v/v)</th>
<th>Required angle of inclination of layer movement, α°</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Size range µm</td>
</tr>
<tr>
<td></td>
<td>45-53</td>
</tr>
<tr>
<td>0.5</td>
<td>34°</td>
</tr>
<tr>
<td>1.0</td>
<td>32°</td>
</tr>
<tr>
<td>1.5</td>
<td>31°</td>
</tr>
</tbody>
</table>

* Glass beads dispersed in distilled water
Effect of mean particle size on the required angle for layer movement

Glass beads in distilled water

**FIGURE 5.9**
TABLE 5.11: EFFECT OF PARTICLE SIZE ON THE REQUIRED ANGLE OF INCLINATION FOR LAYER MOVEMENT

| Initial concentration of the suspension* used to form the sludge layer $c_0$ (% v/v) | Required angle of inclination of layer movement, $\alpha^\circ$ |
|---|---|---|
|  | 5-10 | 45-53 | 90-125 |
| 0.5 | 42° | 36.5° | 31° |
| 1.0 | 40.5° | 36° | 30° |
| 1.5 | 40.5° | 36° | 29° |

* Limestone dispersed in distilled water

From Table 5.11 above it can be seen that, at any given concentration, as the size range of limestone is increased from 5-10 µm to 90-125 µm the angle required for layer movement is decreased. The overall reduction in angle is about 10°. A possible explanation is that as the particle size decreases the surface area of the particles per unit volume will increase. Thus there will be more points of contact between particles which will make it increasingly difficult for particles to move relative to one another. In addition there will be more points of contact between the particles and the underlying surface. Hence, the angle required for flow will be higher for smaller particles.

A similar trend follows when the particle size range of glass beads is increased from 45-53 µm to 90-125 µm but with an even larger decrease in the required angle for layer movement of about 15°. However for all size ranges above 90-125 µm that were examined the angle required for layer movement is greater than that at 90-125 µm. Furthermore, there appears to be little difference in required angle as the size range is increased from 150-180 µm to 355-420 µm. A plausible reason is that in the case of the smaller size range (90-125 µm) there are more overlying layers formed and the whole transport phenomenon resembles a layered chunk of particles sliding over a thin, slow
Effect of mean particle size on the required angle for layer movement

Limestone in distilled water

FIGURE 5.10
moving bottom layer. However, as the particle size is increased fewer overlying layers are formed and the phenomena resembles layered chunks of particles flowing over a relatively thick bottom layer. With increased concentration i.e. increased layers of particles, it would be expected that the difference in angle between 90-125 \( \mu \text{m} \) and the larger sizes would diminish and this is confirmed by the results.

5.3.4 Particle Density

The influence of particle density on the angle at which layer movement occurs was tested by comparing glass beads and bronze spheres of the same size range, 90-125 \( \mu \text{m} \), dispersed in distilled water. The results are shown in Figure 5.11 and Table 5.12.

**TABLE 5.12: EFFECT OF SOLIDS DENSITY ON THE REQUIRED ANGLE OF INCLINATION FOR LAYER MOVEMENT**

<table>
<thead>
<tr>
<th>Initial concentration of the suspension used to form the sludge layer ( c_0 ) (% V/V)</th>
<th>Required angle of inclination for layer movement, ( \alpha^\circ )</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Bronze Spheres* in distilled water</td>
</tr>
<tr>
<td></td>
<td>Perspex Surface</td>
</tr>
<tr>
<td>0.5</td>
<td>22°</td>
</tr>
<tr>
<td>1.0</td>
<td>19°</td>
</tr>
<tr>
<td>1.5</td>
<td>18.5°</td>
</tr>
</tbody>
</table>

* Size range of solids = 90-125 \( \mu \text{m} \)
Density of bronze spheres = 7700 kg/m\(^3\)
Density of glass beads = 2460 kg/m\(^3\)

An increase in density results in a greater gravitational force down the slope. If a simple frictional model adequately described the flow of sludge down the inclined plane then increasing the density would not alter the angle at which the sludge flow occurred provided that the drag force was negligible, i.e.

\[
ma = mg \sin \alpha - \mu mg \cos \alpha - \text{drag} \tag{5.1}
\]
Effect of solids density on the required angle for layer movement

Particle size: 90–125μm
Density of bronze spheres: 7.7g/cc
Density of glass beads: 2.46g/cc

FIGURE 5.11
where $m =$ mass of solids  
$a =$ acceleration  
$g =$ gravitational force  
$\mu_f =$ coefficient of frictional resistance  
$\alpha =$ angle of inclination to horizontal

However, from the results it can be seen that when the underlying surface is constructed of perspex, the angle required for layer movement is in fact higher for the system of bronze spheres even though bronze spheres have a density three times that of the glass beads. A possible reason for this is that because of their higher solids density the bronze spheres have a higher settling velocity and hence the initial layer formed is likely to be more compact. In addition, when the bronze spheres were observed under a microscope they appeared rougher than the glass beads. As was mentioned in an earlier section the particle roughness is itself an important parameter.

With a surface material of PVC, the difference between the angles for movement of the two particulate systems is negligible. Possibly this is because the increase in surface roughness has more effect on the glass bead system which has a a low particle-particle roughness compared to the bronze sphere system. This is borne out by the fact that bulk movement no longer occurs in the former case.

5.3.5 Liquid Viscosity

To investigate the effect of viscosity, systems of glass beads dispersed in two liquids of differing viscosities were compared. These were a plasticiser mixture of Reofos 65/Reomol DBP and distilled water to which Nonidet had been added. Nonidet was added to the distilled water in order that the surface tensions of the two liquids were identical and thus a direct comparison of the effect of viscosity could be made. The results are given in Table 5.13 and in Figure 5.12.
Effect of liquid viscosity on the required angle for layer movement

Glass beads (90–125 μm)
Density of Reofos 65/Reomol DBP: 1.078 g/cc
Density of distilled water: 0.9974 g/cc

FIGURE 5.12
TABLE 5.13: EFFECT OF LIQUID VISCOSITY ON THE REQUIRED ANGLE OF INCLINATION FOR LAYER MOVEMENT

<table>
<thead>
<tr>
<th>Initial concentration of the suspension used to form the sludge layer, (c_0) (% V/V)</th>
<th>Required angle of inclination for layer movement, (\alpha^0)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Glass beads* in Reofos 65/Reomol DBP</td>
</tr>
<tr>
<td>0.5</td>
<td>23°</td>
</tr>
<tr>
<td>1.0</td>
<td>23°</td>
</tr>
<tr>
<td>1.5</td>
<td>22°</td>
</tr>
</tbody>
</table>

* Size range of glass beads = 90-125 µm
Viscosity of Reofos 65/Reomol DBP at 25°C = 22.2528 cp
Viscosity of distilled water at 25°C = 0.896 cp

The required angles of inclination for layer movement using the more viscous plasticisers mixture are consistently 5-6 degrees higher than those using distilled water. Moreover, the rate of sludge discharge along the lower surface was observed to be very much slower in the former case. The plasticiser mixture has a viscosity which is about 25 times greater than that of water. This should have a significant effect on the flowability of the sludge as the plasticiser's mixture will exert a much stronger viscous resistance to sludge movement which will tend to reduce the sludge flowrate.

5.3.6 Surface Tension of Liquid

Nonidet is a dispersant which is expected to enhance the flow of sludge by improving the wetting effect on the particles and inclined surface by a reduction in surface tension (see Table 4.2). Four systems were tested, namely glass beads of size ranges 90-125 µm and 355-420 µm and bronze spheres and ground glass both of size range 90-125 µm. All of the above were dispersed in distilled water. For results refer to Table 5.14 and Figures 5.13, 5.14, 5.15 and 5.16.

The particulate system of bronze spheres was most significantly affected with overall reductions of between 3-5 degrees after three
Effect of surface tension on the required angle for layer movement

Glass beads (90–125μm)

FIGURE 5.13
Effect of surface tension on angle required for layer movement

Glass beads (355–420µm)

FIGURE 5.14
Effect of surface tension on the required angle for layer movement

Ground glass (90–125μm)

FIGURE 5.15
Effect of surface tension on the required angle for layer movement

Bronze spheres (90–125 \( \mu \)m)

FIGURE 5.16
drops of Nonidet had been added. The smaller size of glass spheres were slightly more affected than the larger size. The smallest reduction in the angle required for flow occurred in the ground glass system and was 1-2 degrees only.

TABLE 5.14: EFFECT OF SURFACE TENSION ON THE ANGLE REQUIRED FOR LAYER MOVEMENT

<table>
<thead>
<tr>
<th>Particulate System</th>
<th>Initial concentration of the suspension used to form the sludge, C₀ (% v/v)</th>
<th>Required angle of inclination for layer movement, θ °</th>
<th>Number of drops of Nonidet</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td>0</td>
</tr>
<tr>
<td>Glass Beads 90-125 μm</td>
<td>0.5</td>
<td>17.5</td>
<td>15</td>
</tr>
<tr>
<td></td>
<td>1.0</td>
<td>17</td>
<td>14.5</td>
</tr>
<tr>
<td></td>
<td>1.5</td>
<td>17</td>
<td>15</td>
</tr>
<tr>
<td>Glass Beads 355-420 μm</td>
<td>0.5</td>
<td>20</td>
<td>18.5</td>
</tr>
<tr>
<td></td>
<td>1.0</td>
<td>19</td>
<td>18.5</td>
</tr>
<tr>
<td></td>
<td>1.5</td>
<td>18.5</td>
<td>18</td>
</tr>
<tr>
<td>Ground Glass 90-125 μm</td>
<td>0.5</td>
<td>28</td>
<td>25.5</td>
</tr>
<tr>
<td></td>
<td>1.0</td>
<td>25</td>
<td>25.5</td>
</tr>
<tr>
<td></td>
<td>1.5</td>
<td>24</td>
<td>24.5</td>
</tr>
<tr>
<td>Bronze Spheres 90-125 μm</td>
<td>0.5</td>
<td>22</td>
<td>19</td>
</tr>
<tr>
<td></td>
<td>1.0</td>
<td>19</td>
<td>18</td>
</tr>
<tr>
<td></td>
<td>1.5</td>
<td>18.5</td>
<td>18</td>
</tr>
</tbody>
</table>

One would expect that the smaller size of glass beads would be more significantly affected by the addition of Nonidet than the larger size as the former have a higher surface area per unit volume available for wetting which will tend to make the effect of the Nonidet more pronounced. An exception to this occurs with the ground glass system which will have an even greater surface area to volume ratio than the glass beads of the same size range but is affected to a lesser extent. A possible reason could be that due to the roughness of the particles there are points at which the particles interlock (see Figure 5.17)
which inhibits the motion of particles relative to one another. An improvement in wettability would not be expected to significantly improve the flow of such particle contacts relative to one another.

FIGURE 5.17: INTERLOCKING LAYERS OF PARTICLES

5.3.7 Thickness of the Sludge Layer

As the thickness of the sludge layer is increased, the angle for movement tends to be reduced, but by a decreasing increment until it reaches a limiting value.

For very thin layers (of the order of 1-2 particle layers) only heap or bulk movement occurs and much higher angles are required for the motion of the sludge. The main controlling factor in this case will be the interaction between particles and the underlying surface. For thicker layers when layer movement occurs, there will be a combination of particle-particle interactions (as the overlying layers flow over the underlying layers) and particle-surface interactions (as slow motion of the lower layers occurs during layer movement). The latter will help to promote the flow of the overlying layers.
CHAPTER 6
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CHAPTER 6
OPERATING PERFORMANCE OF THE CONTINUOUS LAMELLA SEPARATOR

6.1 MAXIMUM HANDLING CAPACITY FOR CLEAR LIQUID OVERFLOW

6.1.1 Introduction

It is commonly found in the literature that theoretical predictions for the separator capacity are more than twice the actual capacity. Earlier workers\textsuperscript{28,29} claimed that the discrepancies were mainly due to stability problems and mixing within the settling channels. More recently Poh\textsuperscript{48} was able to show that there were other significant causes. These were that:

i) the operating requirements for achieving essential steady-state conditions were not satisfied;

ii) the theoretical models were applied beyond their limits of validity. This was because the models contained too many simplifying assumptions which both restrict and make it difficult to define their range of application.

He was able to verify the above by conducting his experiments under controlled conditions which surpressed the potential causes of discrepancy. He went on to show that, under these conditions, the general level of agreement between theory and experiment was much higher than that obtained by previous workers. Any deviations that occurred were then due to flow instability and mixing problems. Prior to Poh's work, Acrivos and Herbholzheimern\textsuperscript{19} had theoretically shown, using continuum mechanics, that the Nakamura-Kuroda equation could predict the settler overflow provided that the Sedimentation Reynolds Number was low and that $\Lambda$, the ratio of the Grashof Number to the Sedimentation Reynolds Number was high. Poh\textsuperscript{48} found experimentally that the Nakamura-Kuroda equation was accurate under the following conditions:

$$\Lambda = (0(10^4) - 0(10^3))$$
$$Re = (10(1) - 0(10))$$

He used a particulate system of glass beads of size range 90-125 um dispersed in a plasticising solution of Reofos
65/Reomol DBP which was well within these constraints. In most real systems, however, the dispersing liquid is water. This has a viscosity 22 times less than the Reofos/Reomol and consequently this makes it very difficult for the above-mentioned constraints to be met.

In the following chapter the operating performance of two particulate systems in a continuously operated lamella separator has been examined. One system is well outside the Reynolds Number constraint - limestone of size range 45-53 μm dispersed in water, and one is outside the range for A - limestone of size range 5-10 μm dispersed in water. Their maximum handling capacity in a continuous settler was determined experimentally over a range of settler angles, lengths, plate spacings and modes of operation. The concentration of the feed suspension in both cases was 0.5% v/v. The performance of the separator will be assessed by:

i) comparing the experimental maximum overflow to that predicted by the Namakura-Kuroda equation. The experimental value divided by the actual value will be defined as the efficiency

ii) determining the maximum overflow rates achievable in set floor areas over the range of variables studied i.e. plate length, angle and spacing and mode of settler operation

iii) comparing the maximum overflow rate achievable in a lamella settler to that in a conventional vertical settler of identical area.

The aims of the investigation are:

i) to provide a comparison between the actual maximum overflow and that predicted by the Nakamura-Kuroda equation to determine whether it is a valid means of predicting the overflow rate for either system

ii) to establish practical design guidelines which will enable the highest operating efficiencies to be achieved

iii) to determine whether the modes of behaviour observed in the ideal system occur in a real system
iv) to determine the optimum mode of operation when the floor space available is limited.

The results are discussed below.

6.1.2 Experimental Observations

This section describes the experimental observations made of the settling suspension. It also outlines the operational problems encountered and how they were overcome.

6.1.2.1 Limestone of size range 45-53 μm

The efficient operation of the lamella separator requires that the interface between the clear liquid and the suspension is stable. For this system the interface was observed to be unstable over the entire range of equipment variables. The turbulence was greatest at the steepest angles of inclination. This was thought to be caused by gravitational forces which will be more effective at the steeper angles. Increasing the length of settler also reduced the stability of this interface although, in countercurrent operation, its effect was less than that of varying the angle of inclination.

In the cocurrent mode of operation the clear liquid/suspension interface was more sensitive, there being appreciable turbulence whatever the flowrate, angle of inclination or length. The amount of interfacial turbulence was greater at steeper angles and longer separator lengths. It was not possible to run the separator at all at lengths 95 and 112 cm due to excessive carry over of the solids in the overflow.

An additional problem was encountered when operating in the cocurrent mode. Particles from the feed were directly carried over into the overflow on leaving the feed introduction section. This resulted in a high concentration of particles in the overflow. The feed stream entered the feed compartment by means of two inlets of internal
diameter 8 mm set at right angles to the direction of the main flow. It was observed that the feed was jetting into the feed chamber due to its high entry velocity. It is possible that this caused the particles to rebound off the sides of the settler and into the overflow stream hence giving rise to a high concentration of particles in the overflow. Two additional inlets were attached to this section in an attempt to reduce the entry velocity but this was not successful. An aluminium mesh was then placed over the inlet holes which reduced the jetting quite significantly but there was still only limited improvement in the re-entrainment of the solids. Next, feed inlets were fitted to the backplate of the feed section so that the feed entered parallel to the main flow. The fittings used were as large as could be accommodated and gave an estimated threefold reduction in the entry velocity. Furthermore an aluminium mesh was placed over the inlets to help promote the even distribution of the feed. An additional benefit of this system was that it provided a longer entry length over which the feed could settle into laminar flow. These modifications vastly improved the performance of the settler but were still not sufficient to enable it to be run satisfactorily in the cocurrent mode.

The Liquid Reynolds Number in the feed introduction section was determined and was found to be much higher than that in the systems of both Poh\textsuperscript{48} and Probstein\textsuperscript{29}, neither of whom had reported any re-entrainment. The Reynolds number in Poh's system was about 20 and in Probstein's about 230. However in the present system, at a feed flowrate of 0.5 litres/minute, the Reynolds Number was 330. In order to reduce the Reynolds Number the feed chamber was split into several sections using waterproofed honeycombed cardboard. This reduced the Reynolds Number to 70 and had the desired effect of eliminating the re-entrainment. The feed flowrate was then increased until re-entrainment again occurred which was at a Reynolds Number of 130. A series of experiments were then performed in which further inserts were added to each section and the flowrate increased until re-entrainment occurred. It was found that direct re-entrainment occurred when the Reynolds Number was greater than 130. This
emphasises the need for careful design of the feed section if re-entrainment is to be avoided and the settler operated efficiently.

6.1.2.2 Limestone of size range 5-10 μm

The interface between the clear and suspension layers was indistinct which led to difficulties in observing its behaviour. This was because it had not been possible to remove all the particles of under 5 μm in the cutting of the limestone size fraction using the Alpine Classifier. As a result, many of these fine particles were carried over from the suspension layer into the clear liquid stream and were subsequently removed in the overflow. This resulted in the clear liquid layer becoming cloudy and reduced the clarity of the interface between it and the suspension layer. To assist in the observation of this interface, it was illuminated by a laser beam.

In the countercurrent system, the interfacial turbulence increased both with increasing settler length and at steeper angles. This phenomena was also observed with the limestone of size range 45-53 μm although in the latter case the magnitude of the turbulence was greater.

In the cocurrent system no direct carry over of particles from the feed to the overflow took place as it did with the 45-53 μm limestone. This is as expected since the Liquid Reynolds Number in the feed introduction section was under 30 for the 5-10 μm system. However, because the particles were small, they were able to pass directly from the feed stream into the overflow through a small clearance in the top section between the edges of the adjustable plate and the settler wall. This problem could only be overcome by permanently sealing up the clearance using perspex glue. The top module was then adapted to permit two inlet/outlet configurations to be tested (the minimum required if both subcritical and supercritical conditions were to be attained) by making the top module reversible. This was achieved by attaching an additional overflow outlet on the lower surface of the existing module. To achieve the second condition the top module was
inverted with the exception of its upper flange plate which remained in the same position relative to the main body of the settler. With an inlet plate spacing of 1 cm and an outlet plate spacing of 2 cm the supercritical mode of operation was attained. When the top module was reversed with an inlet plate spacing of 2 cm and an outlet plate spacing of 1 cm, subcritical conditions were achieved.

On start-up in both the cocurrent systems, the incoming suspension flowed down the settling channel in a thin concentrated layer adjacent to the lower surface. Interfacial turbulence in the form of waves (see Figure 6.1) occurred between this layer and the clear layer above it. The magnitude of these turbulent waves increased both with increasing settler length and at steeper angles. As the concentrated suspension reached the lower end of the settler the turbulence had increased to such an extent that particles were ejected into the layer above, forming a suspension of low concentration which moved in the upwards direction.

In the subcritical mode of operation only, a low concentration suspension layer was also formed above the concentrated layer at the top end of the settler. This layer extended almost to the upper wall of the settler and moved in a downwards direction. It eventually met with the upward moving layer formed from ejected particles at the lower end of the settler. This caused the formation of circulatory currents at this point which gradually extended throughout the entire layer. In the supercritical mode some circulatory currents of a smaller magnitude were present in the less concentrated layer. These were caused by the movement in opposite directions of this layer and the concentrated suspension layer.

6.1.3 Maximum Overflow Rate

This section discusses the maximum overflow rates achieved in the continuous rig. It also describes the effect of settler variables and mode of operation on the maximum overflow. The maximum overflow rates obtained on the experimental rig ($Q_{\text{expt}}$) are compared to the values
FIGURE 6.1(a): Initiation of interfacial turbulence

FIGURE 6.1b: Ejection of particles from sludge layer

FIGURE 6.1: INTERFACIAL TURBULENCE WITH 5-10 μm LIMESTONE IN THE COCURRENT MODE
predicted for the maximum overflow rate by the Nakamura-Kuroda equation \( Q_{\text{theo}} \) where

\[
\text{Settler Efficiency} = \frac{Q_{\text{expt}}}{Q_{\text{theo}}} \times 100\%
\]

6.1.3.1 Limestone of size range 45-53 mm

When the experimental maximum overflow rates are compared to the Nakamura-Kuroda equation (see Tables A1 and A2) they are found to be significantly lower—between 36 to 49% for the countercurrent system and between 31 to 44% for cocurrent flow. The discrepancy between the actual and predicted values is not surprising since the values of the Sedimentation Reynolds Number are high and lie between 450 to 1700 outside the range recommended by Acrivos and Herbolzheimer. It would be expected that the discrepancy between the actual and predicted values would diminish as the Sedimentation Reynolds number decreased (i.e., at the shallower angles and shorter settler lengths. See Tables A1 and A2). This is the case when the settler length is reduced (see Figure 6.2) and when the angle of inclination to the horizontal is reduced at settler lengths 0.95 and 1.12m (see Figure 6.3) in the countercurrent mode and at lengths 0.49 and 0.66m in the cocurrent mode (see Figure 6.4). This finding is supported by the experimental observation that there was less interfacial turbulence between the clear and suspension layers as the settler length is reduced. Surprisingly however, when the angle to the horizontal is reduced at the 0.49m and 0.66m settler lengths in the countercurrent mode, there is a decrease in efficiency (see Figure 6.3).

For countercurrent flow, the maximum overflow rate (see Figure 6.5 and Table A1) for a given length of separator decreases steadily with increasing settler inclination angle to the horizontal except above 55° whereupon a much larger decrease occurs. A similar trend occurred in the cocurrent mode (see Figure 6.6 and Table A2). The low experimental maximum
Effect of settler length on the settler efficiency in the countercurrent mode

Limestone (45–53 μm) in water

FIGURE 6.2
Effect of settler inclination angle on the efficiency in the countercurrent mode

Limestone (45–53μm) in water

FIGURE 6.3
Effect of settler inclination angle on the settler efficiency in the cocurrent mode

Limestone (45–53μm) in water

FIGURE 6.4
Effect of settler inclination angle on the maximum overflow in the countercurrent mode

Limestone (45–53 μm) in water

FIGURE 6.5
Effect of settler inclination angle on the maximum overflow in the cocurrent mode

Limestone (45-53 µm) in water

FIGURE 6.6
overflow rates at steep angles are caused by the sludge being lifted off the lower inclined surface and re-entrained.

In the countercurrent mode, there is an almost linear relationship (see Figure 6.7) between settler length and maximum overflow, but with slightly higher values of maximum overflow at the shortest length. For cocurrent flow the operation of the settler at lengths above 0.66m was not possible due to excessive turbulence between the clear and suspension layers.

A plot of aspect ratio versus maximum overflow (Figure 6.8) did not yield the three regions observed by Poh. An upper limit to the overflow was not achieved with the countercurrent system, although it is possible that this could occur outside the range tested. However at the steeper angles of 45° and 55° to the horizontal, the overflow values start to level off with increasing aspect ratio.

It is useful to compare the performance of the countercurrent and cocurrent modes of operation. In general the cocurrent system was more susceptible to excessive turbulence than the countercurrent mode of operation. This rendered it inoperable at lengths above 0.66m. However with the settler lengths of 0.49 and 0.66m at angles of 35° and 40° the cocurrent system was more efficient (see Figure 6.9) than the countercurrent system. At the steeper angles, however, the highest overflow rates were achieved in the countercurrent system. This was because the solids sliding down the lower inclined surface in the cocurrent mode start to be lifted off the surface and re-entrained. In the countercurrent system this did not occur to such an extent because the solids travelled at a lower velocity down the plate.

6.1.3.2 Limestone of size range 5-10 μm

Efficiencies of between 37-61% for the countercurrent system, between 41.2-77.9% for the cocurrent subcritical system and between 62.8-81.6%
Effect of settler length on the maximum overflow in the countercurrent mode.

Limestone (45–53\(\mu m\)) in water

FIGURE 6.7
Effect of aspect ratio on the maximum overflow in the countercurrent mode

Limestone (45–53 μm) in water

FIGURE 6.8
Comparison of the maximum overflows in the cocurrent and countercurrent modes of operation

Limestone (45–53 μm) in water

FIGURE 6.9
for the cocurrent supercritical system were obtained (see Tables A3, A4 and A5) for a plate spacing of 3.4 cm. The efficiency values were much higher with the 5-10 µm limestone than with the 45-53 µm limestone. The improvement in efficiency occurred for two reasons. Firstly, the Sedimentation Reynolds Numbers are much lower with the smaller sized limestone and lie between the values of 6-21. These values are close to or within the range deemed by Poh to be necessary for the accurate prediction of the maximum overflow rate by the Nakamura-Kuroda equation. Secondly, due to the greater vertical settling velocity of the 45-53 µm limestone, its maximum overflow rates are much greater and hence the velocity within the overflow layer will be much higher than with 5-10 µm limestone. This results in an unfavourable velocity field in the vicinity of the clear/suspension layer interface since a greater portion of the suspension layer will be dragged along in the direction of the clear liquid stream thus increasing the number of particles being directly re-entrained into the overflow.

In the countercurrent and cocurrent subcritical systems the efficiency increased with decreasing settler length (see Figures 6.10 and 6.11) and decreasing angle to the horizontal (see Figures 6.12 and 6.13) and hence with decreasing Sedimentation Reynolds Number and decreasing values of A, the ratio of the Sedimentation Reynolds Number to the Grashof Number. This trend is expected since for Reynolds Number values under 10, Acrivos and Herbolzheimer were able to show theoretically that the Nakamura-Kuroda equation could predict the overflow rate accurately. However, although there was an improvement in efficiency, accurate predictions were not obtained, the highest efficiency being 77.9%. A possible reason is that A remains outside the limits determined by Poh. He predicted theoretically that A should lie in the range $0(10^4) - 0(10^7)$ whereas for this system it is in the range $0(10^8) - 0(10^9)$. Hence the steady state constraint cannot be completely satisfied. Other potential causes of discrepancies such as the laminar flow constraint and the constraint on the plate spacing have been fulfilled and hence could not have been responsible for the low efficiencies. In the cocurrent supercritical
Effect of settler length on the efficiency in the countercurrent mode with a plate spacing of 0.034 m

Limestone (5–10 μm) in water

FIGURE 6.10
Effect of settler length on the efficiency in the cocurrent subcritical mode with a plate spacing of 0.034m.

Limestone (5–10μm) in water

![Graph showing settler efficiency vs. settler length](image)

**FIGURE 6.11**
Effect of settler angle on the efficiency in the countercurrent mode with a plate spacing of 0.034m

Limestone (5–10μm) in water

FIGURE 6.12
Effect of settler angle on the efficiency in the cocurrent subcritical mode with a plate spacing of 0.034 m

Limestone (5–10 μm) in water

FIGURE 6.13
system there was little change in efficiency with either settler length, angle, Sedimentation Reynolds Number or $\Lambda$ (see Figures 6.14 and 6.15) which suggests that there is some fault in the predictive equation itself for this mode of operation.

Graphs were plotted of efficiency versus Sedimentation Reynolds Number (see Figures 6.16 and 6.17). At low Reynolds Numbers ($0(1)-0(10)$), as is the case with the 5-10 $\mu$m limestone, the efficiency decreases almost linearly with increasing Reynolds Number (see Figure 6.16). The efficiency of the cocurrent subcritical and the countercurrent modes significantly decreased with increasing Reynolds number, whilst the cocurrent supercritical mode decreased only slightly. Thus, as was found by Poh, the cocurrent supercritical mode is a more stable system. When the graphs of Reynolds Number versus efficiency were extrapolated back to zero Reynolds Number, 100% efficiency was still not reached. This suggests that the Nakamura-Kuroda equation is not capable of predicting the maximum overflow of this system. At higher Reynolds Numbers ($0(100)-0(1000)$), as obtained with the 45-53 $\mu$m limestone, the efficiency remained relatively constant at around 40% (see Figure 6.17). This result suggests that, for systems with a high Sedimentation Reynolds Number, very little improvement in efficiency can be achieved by altering the settler variables unless they give rise to a large decrease in the Sedimentation Reynolds Number. In order to increase the settler efficiency with the 5-10 $\mu$m limestone system, the sedimentation Reynolds Number and $\Lambda$ should be reduced in order to meet the constraints set on them. Reducing the angle to the horizontal or reducing the settler length will reduce the Sedimentation Reynolds Number and $\Lambda$. There is, however, a lower limitation set on the angle at which the settler can be operated due to the sludge flow constraint. Thus the controlling variable must be the settler length.

In order to fulfill both the constraint on $\Lambda$ and the Sedimentation Reynolds Number for the 5-10 $\mu$m limestone system with the settler angle at its minimum value of $35^0$ to the horizontal, it was calculated that the settler length would need to be less than 9 cm. This design
Effect of settler length on the efficiency in the cocurrent supercritical mode with a plate spacing of 0.034m.

Limestone (5–10 μm) in water

FIGURE 6.14
Effect of settler angle on the efficiency in the cocurrent supercritical mode with a plate spacing of 0.034m.

Limestone (5–10 μm) in water

FIGURE 6.15
Effect of Sedimentation Reynolds Number on the settler efficiency

Limestone (5–10μm) in water

FIGURE 6.16
Effect of Sedimentation Reynolds Number on the settler efficiency

Limestone (45–53μm) in water

FIGURE 6.17
gave total projected settling areas of only 0.0855 and 0.4423 m$^2$ in floor areas of 0.08 and 0.4 m$^2$ respectively. This is more than 5 times less than the total projected areas that can be achieved in equivalent floor areas with the settler lengths of 0.49-1.12m which were tested experimentally. Thus it is evident that, even if the plate efficiency were 100% at the settler length of 9 cm, the maximum overflow possible at this length in any floor area would be lower than what could be achieved with the longer settler lengths. However, when there are no limitations on land area, savings could be made on the cost of settler materials by operating at shorter lengths. This could be an important factor if the plates need to be made of an expensive material, for instance, in order to achieve good solids flow conditions.

The experimental results were consistent with Probstein's finding that the countercurrent and cocurrent subcritical systems were more susceptible to particle re-entrainment than the cocurrent supercritical system as shown not only by the higher efficiencies achieved in the latter mode (see Figure 6.16) but also by the experimental observation of the flow stability of the different layers within the settler.

Leung carried out research into flow stability and found that the settler efficiency decreased with increasing angle. This was confirmed by the present experimental results, although the drop in efficiency was not proportional to $1/\sin \alpha$ as he predicted. He further predicted that the separator efficiency would increase with plate spacing up to a value of 10 cm. In the present work three plate spacings have been tested experimentally, 1.5, 3.4 and 8 cm. In the countercurrent mode the efficiency passed through a maximum value at a plate spacing of 3.4 cm for a settler length of 0.49m and increased with plate spacing for a settler length of 0.66m (see Table A6 and Figure 6.18). Particularly unfavourable flow conditions were encountered in both cocurrent systems at a plate spacing of 8 cm and the efficiencies achieved at this plate spacing were much lower than at 3.4 cm.
Effect of the plate spacing on the settler efficiency in the countercurrent mode

Limestone (5–10μm) in water

FIGURE 6.18
The effect of settler variables on the maximum overflow is discussed below. For all modes the maximum overflow decreased with increasing angle to the horizontal (see Tables A3, A4, A5 and A7 and Figures 6.19, 6.20, 6.21 and 6.22). This was due to the increasing instability of the clear layer/suspension layer interface at steeper angles and the decrease in projected settling area.

When the settler length in the countercurrent mode was increased at plate spacings of both 1.5 and 3.4 cm the maximum overflow tended to increase but by decreasing increments to a limiting value. At an angle of 55° with a plate spacing of 3.4 cm and at angles of 45° and above for a plate spacing of 1.5 cm the maximum overflow actually decreased when the length of the settler was increased from 0.95 to 1.12m (see Figures 6.23 and 6.24). This was reflected by a sharp drop in efficiency at these angles. A possible reason for the decrease in overflow rate is that the interface between the clear and suspension layers was observed to become more unstable at longer settler lengths thus leading to increased particle re-entrainment. The interface was particularly unstable with the 1.5 cm plate spacing which accounts for the decrease in overflow at longer lengths occurring over a wider range of angles. The narrower plate spacing does not alter the value of either $A$ or the Sedimentation Reynolds Number so this cannot be the cause of the increased instability. However the flowrate of the suspension between the plates will be greater for the narrower plate spacing. This could result in more unfavourable flow conditions within the suspension layer leading to a greater part of the suspension layer being dragged along with the clear liquid stream, and the subsequent re-entrainment of particles into the clear layer.

In the cocurrent subcritical mode a similar result (see Table A4 and Figure 6.25) to the countercurrent mode was obtained. In this mode the settler could be operated at the lower angle of 35°. No levelling off of the maximum overflow rate occurred at this angle. This was because flow conditions were more stable due to the lower values of the Sedimentation Reynolds Number and $A$. 
Effect of settler angle on the maximum overflow in the countercurrent mode with a plate spacing of 0.034m

Limestone (5-10μm) in water

FIGURE 6.19
Effect of settler angle on the maximum overflow in the countercurrent mode with a plate spacing of 0.015m

Limestone (5-10μm) in water

FIGURE 6.20
Effect of settler angle on the maximum overflow in the cocurrent subcritical mode with a plate spacing of 0.034m

Limestone (5-10μm) in water

FIGURE 6.21
Effect of settler angle on the maximum overflow in the cocurrent supercritical mode with a plate spacing of 0.034m

Limestone (5–10μm) in water

![Graph showing the effect of settler angle on overflow](image-url)

FIGURE 6.22
Effect of settler length on the maximum overflow in the countercurrent mode with a plate spacing of 0.034 m

Limestone (5-10 μm) in water

FIGURE 6.23
Effect of settler length on the maximum overflow in the countercurrent mode with a plate spacing of 0.015m

Limestone (5–10μm) in water

FIGURE 6.24
Effect of settler length on the maximum overflow in the cocurrent subcritical mode with a plate spacing of 0.034 m

Limestone (5–10 μm) in water

**FIGURE 6.25**

[Graph showing the effect of settler length on maximum overflow]
In the cocurrent supercritical mode (see Figure 6.26 and Table A5) the maximum overflow increased with length almost linearly. In this mode the majority of the limestone within the settler was present in a thin concentrated layer which flowed down the settler adjacent to the lower inclined plate. Above this layer there was a suspension layer of low concentration. Due to its low concentration, any interfacial instabilities between this layer and the clear layer above it would result in only low concentrations of limestone entering the clear layer. Hence the effect of flow instability would be less than in the cocurrent subcritical and countercurrent modes in which there is a high concentration suspension layer adjacent to the clear layer. Furthermore the flow was generally more stable in the cocurrent supercritical mode.

The effect of channel spacing on the separator performance is shown in Figure 6.27 and Table A6. In the countercurrent mode for a settler length of 0.66m the maximum overflow increased with plate spacing whereas at a settler length of 0.49m the overflow passes through a maximum at a plate spacing of 3.4 cm. In the cocurrent modes the maximum overflow decreased when the plate spacing was increased from 3.4 to 8 cm. It was expected that the maximum overflow would increase with plate spacing for all settler lengths. Leung predicted that the efficiency would increase up to a plate spacing of 10 cm above which the upper channel wall was far enough removed from the fluid interface that it could be considered at infinity.

A plot of maximum overflow versus aspect ratio did not yield the 3 regions of separation efficiency as identified by Poh (see Figures 6.28, 6.29, 6.30 and 6.31). No region (I) of excellent agreement with the Nakamura-Kuroda equation was apparent in any of the modes tested. However a region (II) of particle re-entrainment in which the optimum overflow rates were not achievable was obtained in all cases. Except for the cocurrent supercritical mode at all angles and the cocurrent subcritical mode at 35° there was a region (III) in which further increases in the aspect ratio produced no significant increase in the achievable overflow rate. Region (III) marked the uppermost limit on
Effect of settler length on the maximum overflow in the cocurrent supercritical mode with a plate spacing of 0.034 m.

Limestone (5–10 \( \mu \)m) in water

FIGURE 6.26
Effect of plate spacing on the maximum overflow in the countercurrent mode

Limestone (5–10μm) in water

FIGURE 6.27
Effect of aspect ratio on the maximum overflow in the countercurrent mode with a plate spacing of 0.034m

Limestone (5–10μm) in water

FIGURE 6.28
Effect of aspect ratio on the maximum overflow in the countercurrent mode with a plate spacing of 0.015m

Limestone (5–10µm) in water

![Graph showing the effect of aspect ratio on the maximum overflow in the countercurrent mode with different settler angles. The graph plots maximum overflow against aspect ratio, with settler angles ranging from 0 to 60 degrees.](image)

FIGURE 6.29
Effect of aspect ratio on the maximum overflow in the cocurrent subcritical mode with a plate spacing of 0.034m

Limestone (5–10μm) in water

FIGURE 6.30
Effect of aspect ratio on the maximum overflow in the cocurrent supercritical mode

Limestone (5–10μm) in water

FIGURE 6.31
the channel length that ought to be used for design purposes. Exceeding this length will be uneconomic because of the diminishing return on the overflow rate. The uppermost limit was defined as \((h/b)_{ul}\). The estimated values of \((h/b)_{ul}\) as obtained from Figures 6.28, 6.29, 6.30 and 6.31 for each mode are summarised below.

TABLE 6.1: RANGE OF \((h/b)_{ul}\) FOR VARIOUS MODES OF OPERATION WITH 5-10 μm LIMESTONE

<table>
<thead>
<tr>
<th>Mode</th>
<th>((h/b)_{ul})</th>
</tr>
</thead>
<tbody>
<tr>
<td>Countercurrent, 3.4 cm plate spacing</td>
<td>23-25</td>
</tr>
<tr>
<td>Countercurrent, 1.5 cm plate spacing</td>
<td>45-61</td>
</tr>
<tr>
<td>Cocurrent subcritical</td>
<td>19-24</td>
</tr>
<tr>
<td>Cocurrent supercritical</td>
<td>Value not reached (&gt; 27)</td>
</tr>
</tbody>
</table>

From the above results it can be seen that in a real system the upper limit on the aspect ratio depends on the mode of settler operation. This is in contrast to Poh's finding in an ideal system. In addition, the upper limit in the cocurrent supercritical system was not reached even though aspect ratios up to a value of 27 were tested. Hence for this mode \((h/b)_{ul}\) will be higher than that achieved in the ideal system tested by Poh.

With the 3.4 cm plate spacing, the value of \((h/b)_{ul}\) was far below the aspect ratio recommended in the present design for a lamella settler where \((h/b)_{design}\) is specified as ranging from 40-50. At a plate spacing of 1.5 cm, the value of \((h/b)_{ul}\) was much higher than that at 3.4 cm. Thus, it would seem that in a real system \((h/b)_{ul}\) is dependent on the channel spacing. Further real systems need to be tested but the results suggest that it would be better to specify an upper limit on \(h\) instead of \(h/b\). The upper values for \(h\) in each mode are given below and it can be seen that they lie within a close range.
A surprising result is that with the 45-53 μm limestone system, \((h/b)_{ul}\) was not reached over the range of aspect ratios tested which were identical to those of the 5-10 μm limestone. It would have been expected that, since the Sedimentation Reynolds Numbers were much higher and therefore instability more inherent, \((h/b)_{ul}\) would become correspondingly lower. However with the 45-53 μm limestone system the efficiency was affected to a much lesser degree by Sedimentation Reynolds Number than was the 5-10 μm limestone system. A general guideline appears to be that if Re is high then the efficiency will be low and will not be markedly reduced by increasing the aspect ratio. For lower Reynolds Numbers much higher efficiencies can be achieved but the overflow is more sensitive to aspect ratio.

In the 45-53 μm limestone system, the cocurrent mode was much more susceptible to excessive turbulence than the countercurrent mode. The reverse is the case with the 5-10 μm limestone system. Furthermore, no direct carry over of particles into the overflow occurred with the 5-10 μm limestone. The range of settler variables were identical and the constraints on the plate spacing and laminar flow were satisfied in both cases.
6.2  **EFFECT OF LIMITATIONS IMPOSED ON THE DESIGN**

This section considers cases in which there are limitations imposed on the design of the settler which will ultimately affect the optimum design.

6.2.1  **Cost of Settler Plates is the Limiting Factor**

If the lamella plates are expensive and the limiting factor is the capital cost, then the least costly design will be the one with the highest single plate efficiency.

6.2.1.1  **Limestone of size range 45-53 μm**

From Figure 6.2 it can be seen that the efficiency of the settler increased with decreasing settler plate length at settler angles of 45° and 55° and remained constant at 35°. Thus it would be advisable to design a settler with the shortest plates possible.

6.2.1.2  **Limestone of size range 5-10 μm**

The highest single plate efficiencies occurred when operating in the cocurrent supercritical mode close to the minimum angle of operation at 35°. The length of the settler plates can be set at 1.12m, the maximum tested, since the efficiency of the settler did not vary over the range of lengths tested at this angle. Although the cocurrent mode is more efficient and hence requires fewer plates, the countercurrent settler is a simpler design to construct. Hence a countercurrent module could cost less than a cocurrent module irrespective of the number of plates. The highest plate efficiencies in the countercurrent settler are also at angles close to the minimum angle of operation of 40° but at the shortest length tested, 0.49m.
6.2.2 Limited Land Area Available

Another possible design situation is when the floor space available is limited or expensive. The design will thus require a compact separator which can achieve high overflow rates. An example of this is the insertion of a stack of lamella plates into an existing settling basin.

Two floor areas have been considered, 0.08 m$^2$ and 0.4 m$^2$. The number of 0.04m wide plates (the same width as the experimental settler) that could be accommodated in these areas was then determined over the range of plate spacings, angles and lengths and the modes of settler operation that were studied experimentally. The overflow rates achievable in each floor area over this range of settler variables could then be determined using the experimental values obtained for the maximum overflow rates. The calculation procedure is described below.

From Figure 6.32 it can be seen that

$$ A = W[(N - 1)X + Y] $$

(6.1)
where:  \( A = \) floor area  
\( W = \) plate width  
\( N = \) number of plates  
\( \alpha = \) angle to horizontal

and

\[
X = \frac{b}{\sin \alpha} \quad \quad (6.2)
\]
\[
Y = L \cos \alpha \quad \quad (6.3)
\]

where:  \( b = \) plate spacing  
\( L = \) plate length

Substituting for \( X \) and \( Y \) in equation 6.1 and rearranging gives,

\[
N = \frac{(A - WL\cos \alpha) \sin \alpha}{Wb} + 1 \quad \quad (6.4)
\]

Now

Maximum overflow rate \( \text{Experimental maximum overflow} \)

in floor area, \( A \) = \((N-1)\) rate for a single plate \( \quad (6.5)\)

The maximum overflow rates achieved are also compared to those in a vertical separator of identical floor area.

6.2.2.1 Limestone of size range 45-53 \( \mu \)m

The countercurrent mode of operation will be discussed first. With the larger floor space (see Figure 6.33 and Table A8) of 0.4 m\(^2\), it can be seen that, the longer the settler length, the higher the overflow rate that is achievable. Similar results were obtained with floor areas above 0.4 m\(^2\). With a floor space of 0.08 m\(^2\) (see Figure 6.34 and Table A9), however, increasing the settler length beyond 0.95m results in little or no increase in the overflow rate. This is
Effect of settler length on the maximum overflow in the countercurrent mode with a floor area of 0.4 m²

Limestone (45–53 μm) in water

FIGURE 6.33
Effect of settler length on the maximum overflow in the countercurrent mode with a floor area of 0.08 m$^2$

Limestone (45–53 μm) in water

FIGURE 6.34
due to the area occupied by the end plate which will be a dead space (see Figure 6.35 below).

As the available floor area decreases, the percentage area occupied by the end plate will increase. This will result in the number of possible plates and hence the projected settling area being reduced by a greater factor than the reduction in floor area. In addition an increase in plate length will significantly increase the percentage area occupied by the dead space when the floor area is small. Consider the table below. For the smaller floor area, increasing the plate length by 18% from 0.95 to 1.12m results in an increase in projected area of 4.8% whilst the experimental overflow per unit area is reduced by 4.2%. Thus, overall, little benefit can be expected from increasing the plate length. However, for the larger floor area, the increase in projected area is 16.2% while the experimental overflow per unit area decreases by 4.2%. In this case, it is advantageous to increase the plate length.
### TABLE 6.3: EFFECT OF FLOOR AREA ON PROJECTED AREA

<table>
<thead>
<tr>
<th>Angle of Inclination = 35°</th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Floor Area</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>0.08 m²</td>
<td>0.4 m²</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>Settler Length</td>
<td>Settler Length</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>0.95m</td>
<td>1.12m</td>
<td>0.95m</td>
<td>1.12m</td>
<td></td>
</tr>
<tr>
<td>Actual projected area*</td>
<td>0.560</td>
<td>0.587</td>
<td>4.33</td>
<td>5.03</td>
<td></td>
</tr>
<tr>
<td>Experimental overflow per unit area m/min</td>
<td>0.0498</td>
<td>0.0477</td>
<td>0.0498</td>
<td>0.0477</td>
<td></td>
</tr>
<tr>
<td>% increase in projected area</td>
<td>4.8%</td>
<td></td>
<td>16.2%</td>
<td></td>
<td></td>
</tr>
<tr>
<td>% increase in overflow</td>
<td>-4.2%</td>
<td></td>
<td>-4.2%</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

* Actual projected area = (N-1)WLcosα  (6.6)
+ Experimental overflow per unit area = \( \frac{Q_{\text{expt}}}{WL\cos \alpha} \)  (6.7)

where:  
N = number of plates  
W = plate width  
L = plate length  
\( \alpha \) = plate angle of inclination to horizontal  
\( Q_{\text{expt}} \) = experimental maximum overflow determined for a single plate

The angle of inclination has less effect on the overflow rate achieved. The largest percentage changes in overflow (see Figures 6.36 and 6.37) occur in both floor areas at settler lengths 0.49 and 0.66m where the overflow increases with angle to the horizontal up to 55°. Above 55° the maximum overflow decreases. The overflow also increases with angle for settler lengths of 0.95 and 1.12m in the smaller floor area. With the 0.95 and 1.12m lengths and the floor
Effect of settler inclination angle on the maximum overflow in the countercurrent mode with a floor area of 0.08m²

Limestone (45-53μm) in water

FIGURE 6.36
Effect of settler inclination angle on the maximum overflow in the countercurrent mode with a floor space of 0.4 m

Limestone (45–53 μm) in water

![Graph showing the effect of settler inclination angle on the maximum overflow in the countercurrent mode with different floor spaces.](image)

**FIGURE 6.37**
area 0.4 m² however, the overflow passes through a maximum value between 40° to 45° although the actual variation is very small. The reason for this can be seen by examining the projected area and the overflow rates per unit area (see Table A8) obtained from the continuous rig. As can be seen from Figure 6.38 the overflow per unit area increased with increasing angle at settler lengths 0.49 and 0.66m, whereas at settler lengths 0.95 and 1.12m it passed through a minimum and showed a slight overall decrease. With the 0.08 m² floor area and at settler lengths 0.49 and 0.66m the projected area passes through a maximum at 55° with increasing angle (see Figure 6.39). However the increase in overflow per unit area is proportionally greater and so the maximum overflow increased with increasing angle. At settler lengths 0.95 and 1.12m the projected area passes through a maximum at 50° whereas the overflow per unit area passes through a minimum at 50° and this results in an increase in the maximum overflow with increasing angle. With the 0.4 m² floor area the projected area passes through a maximum at 45° (see Figure 6.40) for each settler length. Hence, due to the relatively large increase in overflow per unit area with increasing angle up to 55° at lengths 0.49 and 0.66m, the actual overflow increased. At lengths 0.95 and 1.12m however, because the overflow per unit area showed an overall decrease, the achievable overflow passed through a maximum.

Thus, for the countercurrent system and larger floor areas, the highest overflow rates can be achieved with longer lengths and angles of between 40° to 45° to the horizontal. With smaller floor areas where the projected area of the end plate becomes significant an angle of 55° at a length of 1.12m is desirable. These are similar to the settler configurations that the Nakamura-Kuroda equation predicted for the highest overflow rate that can be achieved with each floor area (see Tables A8 and A9).

It is also useful to compare the maximum overflow rates in a lamella separator to those in a vertical settler of identical floor area (see Tables A8 and A9). For the floor area 0.08 m², it is possible to obtain overflow rates in the lamella separator of up to 3.9 times
Effect of settler inclination angle on the overflow rate per unit area

Limestone (45–53μm) in water

FIGURE 6.38
Effect of settler inclination angle on the projected area in the countercurrent mode with a floor area of 0.08m²

Limestone (45–53μm) in water

FIGURE 6.39
Effect of settler inclination angle on the projected area in the countercurrent mode for a floor area of 0.4 m$^2$

Limestone (45–53 μm) in water

FIGURE 6.40
those in a vertical settler whilst for the 0.4 m² floor area overflow rates 5.6 times greater can be achieved. This clearly illustrates the benefit of a lamella separator over a vertical settler particularly when the floor space is limited or expensive.

For the cocurrent system, an increase in overflow was achieved with increasing settler length. In contrast to the countercurrent system, however, the overflow rate passed through a maximum value at 35° to 40° and through a minimum at 50°-55° (see Figures 6.41 and 6.42 and Tables A10 and A11) with increasing angle to the horizontal. The latter was because the experimental overflow per unit area as determined in the continuous rig, passed through a minimum with increasing angle to the horizontal whilst the projected area passed through a maximum at 45° to 50°. The highest settling rate was achieved with a settler length of 0.66m at an angle of 40°. However, because the settler could not be operated in this mode at lengths above 0.66m the highest rate was much less than that attained in the countercurrent mode. In addition, at both lengths 0.49 and 0.66m, it was possible to achieve higher overflow rates in the countercurrent mode. Nonetheless, the overflow rates obtained in the cocurrent mode were up to 2.8 times greater than those in a vertical settler of the same floor area (see Tables A10 and A11).

6.2.2.2 Limestone of size range 5-10 μm

The countercurrent mode with a plate spacing of 1.5 cm gave rise to the highest overflow rates (see Tables A12-A19) even though for a single lamella channel of this spacing the lowest efficiencies and the lowest overflow per unit projected area were obtained. This was because the total projected area available for settling with 1.5 cm plate spacing is about twice that for a 3.4 cm plate spacing. Thus, when settler compactness is the important factor, the projected area must be taken into consideration.

The effect of settler length on the maximum overflow in each land area will now be considered. With the smaller floor area of 0.08 m² the
Effect of settler inclination angle on the maximum overflow in the cocurrent mode with a floor area of 0.08 m$^2$

Limestone (45–53 μm) in water

FIGURE 6.41
Effect of settler inclination angle on the maximum overflow in the cocurrent mode with a floor area of 0.4m²

Limestone (45–53μm) in water

![Graph showing the effect of settler inclination angle on maximum overflow](image)

**FIGURE 6.42**
maximum overflow passed through a maximum value under (see Figures 6.43, 6.44 and 6.45 and Tables A12-A14) the following conditions:

- at all settler angles and plate spacings in the countercurrent mode
- at settler angles above 35° to the horizontal in the cocurrent subcritical mode

The maximum value occurred at a settler length of 0.95m in all cases. The maximum overflow increased with settler length (see Figure 6.45 and 6.46 and tables A14 and A15) in the following cases:

- at all settler angles in the cocurrent supercritical mode
- at an angle of 35° to the horizontal in the cocurrent subcritical mode.

With the larger floor area of 0.4 m², the maximum overflow passed through a maximum (see Figures 6.47, 6.48 and 6.49 and Tables A16-A18) with varying settler length in only a few cases:

- in the countercurrent mode with a plate spacing of 1.5 cm at angles above 40°
- in the countercurrent mode with a plate spacing of 3.4 cm at an angle of 55°
- in the cocurrent subcritical mode at an angle of 55°.

In all other cases (see Figures 6.47, 6.48, 6.49 and 6.50 and Tables A16-A19) the maximum overflow increased with settler length.

From the above results it can be seen that the overflow passes through a maximum with increasing settler length occurred under the settler configurations which gave rise to the lowest settler efficiencies, i.e. at the steeper angles, at narrower plate spacings and in the countercurrent and cocurrent subcritical modes. This is because the settler efficiencies decrease with increasing settler length for both plate spacings in the countercurrent mode and at all angles except 35° to the horizontal in the cocurrent subcritical mode. Although the
Effect of settler length in the countercurrent mode on the maximum overflow for a floor space of 0.08m²

Plate spacing = 0.034m
Limestone (5-10μm) in water

FIGURE 6.43
Effect of settler length in the countercurrent mode on the maximum overflow for a floor space of 0.08m²

Plate spacing = 0.015m
Limestone (5–10μm) in water

FIGURE 6.44
Effect of settler length in the cocurrent subcritical mode on the maximum overflow for a floor space of 0.08 m$^2$.

Limestone (5–10 $\mu$m) in water

**FIGURE 6.45**
Effect of settler length in the cocurrent supercritical mode on the maximum overflow for a floor space of 0.08m²

Limestone (5–10μm) in water

FIGURE 6.46
Effect of settler length in the countercurrent mode on the maximum overflow for a floor area of 0.4 m²

Plate spacing = 0.034 m
Limestone (5–10 μm) in water

FIGURE 6.47
Effect of settler length in the countercurrent mode on the maximum overflow for a floor space of 0.4m²

Plate spacing = 0.015m
Limestone (5–10μm) in water

FIGURE 6.48
Effect of settler length in the cocurrent subcritical mode on the maximum overflow for a floor space of 0.4m²

Limestone (5–10μm) in water

FIGURE 6.49
Effect of settler length in the cocurrent supercritical mode on the maximum overflow for a floor space of 0.4m²

Limestone (5-10 μm) in water

FIGURE 6.50
total projected area of the settler in both floor areas increases with settler length, it is insufficient to compensate for the reduced settler efficiency particularly in the smaller floor area, 0.08 m². As with the 45-53 μm limestone, the projected area of the end plate is an important factor (see Figure 6.35). This area is a dead space and in the case of the smaller floor area occupies a significant percentage of the total settling area. The percentage of dead space increases with increasing settler length. This reduces the remaining available settling area and causes the overflow to pass through a maximum over a greater range of settler configurations. With the larger floor area, the projected area of the end plate is insignificant compared to the total projected area and hence the overflow only passes through a maximum value in those cases where the settler efficiency decreases markedly with increasing settler length. In the cocurrent supercritical mode at all angles and the cocurrent subcritical mode at 35°, the efficiencies change little with settler length and hence the maximum overflow increased with settler length. Thus, if the settler efficiency decreases with plate length the overflow will tend to pass through a maximum. However, if the settler efficiency is not significantly affected by increasing settler length, then the maximum overflow will increase with length.

No trends were apparent when graphs of the maximum overflow versus settler angle were plotted. In the countercurrent mode with a 1.5 cm plate spacing the maximum value decreased with increasing settler angle to the horizontal for both floor areas (see Figures 6.51 and 6.52). This was because the largest decrease in settler efficiency with increasing angle occurred in this mode. In the cocurrent subcritical mode with both floor areas (see Figures 6.53 and 6.54) and the countercurrent mode with a plate spacing of 3.4 cm and a floor area of 0.08 m² (see Figure 6.55), the maximum overflow passes through a maximum at 45–50°. In the countercurrent mode with a plate spacing of 3.4 cm and a floor area of 0.4 m² (see Figure 6.56) the overflow decreases slightly with increasing angle to the horizontal. In both the cocurrent subcritical mode and countercurrent mode with a plate spacing of 3.4 cm the efficiency decreased with increasing angle to
Effect of settler angle in the countercurrent mode on the maximum overflow for a floor space of 0.08m²

Plate spacing = 0.015m
Limestone (5-10μm) in water

FIGURE 6.51
Effect of settler angle in the countercurrent mode on the maximum overflow for a floor area of 0.4m²

Plate spacing = 0.015m
Limestone (5-10μm) in water

![Graph showing the effect of settler angle on maximum overflow]

**FIGURE 6.52**
Effect of settler angle in the cocurrent subcritical mode on the maximum overflow for a floor space of 0.08m²

Limestone (5-10μm) in water

FIGURE 6.53
Effect of settler angle in the cocurrent subcritical mode on the maximum overflow for a floor space of 0.4m$^2$.

Limestone (5–10μm) in water

FIGURE 6.54
Effect of settler angle in the countercurrent mode on the maximum overflow for a floor space of 0.08m²

Plate spacing = 0.034m
Limestone (5-10μm) in water

FIGURE 6.55
Effect of settler angle in the countercurrent mode on the maximum overflow for a floor space of 0.4 m²

Plate spacing = 0.034 m
Limestone (5–10 μm) in water

FIGURE 6.56
In the cocurrent supercritical mode with a floor spacing of 0.08 m², the maximum overflow passes through a maximum at 50° for settler lengths of 0.66, 0.95 and 1.12m (see Figure 6.57), but shows little change with angle at 0.49m. With a floor spacing of 0.4 m² at all settler lengths, little change in overflow occurs with angle (see Figure 6.58). This is because the settler efficiency in this mode is not affected by angle. Thus the controlling factor on the settler overflow will be the projected settling area available (see Figures 6.39 and 6.40).

As a guideline, if the settler efficiency is markedly reduced (more than 10% change over 15°) by increasing the settler angle then the highest overflow rates will be achieved at the shallower angles. However if the efficiency increases slightly or not at all with decreasing angle, then the optimum angle can be predicted by calculating the settler configuration with the maximum projected area.

To summarise, the highest overflow rates were achieved in the countercurrent mode with a plate spacing of 1.5 cm, a settler length of 0.95m and an angle of 40°. The cocurrent supercritical mode gave the next highest overflow rates at a settler length of 1.12m and an angle of 45°.

The maximum overflow rates achieved in a lamella separator were also compared to those in a vertical separator of identical floor area. The results are summarised below:
Effect of settler angle in the cocurrent supercritical mode on the maximum overflow for a floor space of 0.08m²

Limestone (5-10µm) in water

FIGURE 6.57
Effect of settler angle in the cocurrent supercritical mode on the maximum overflow for a floor space of 0.4m².

Limestone (5-10 μm) in water

FIGURE 6.58
TABLE 6.4: COMPARISON OF THE MAXIMUM OVERFLOW RATES ACHIEVABLE IN A LAMELLA SEPARATOR TO THOSE IN A CONVENTIONAL VERTICAL SEPARATOR OF IDENTICAL FLOOR AREA

<table>
<thead>
<tr>
<th>Mode</th>
<th>Floor Area</th>
<th>Lamella overflow/vertical overflow</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.08 m²</td>
<td></td>
</tr>
<tr>
<td>Countercurrent</td>
<td>0.4 m²</td>
<td>4.82-7.85</td>
</tr>
<tr>
<td>1.5 cm plate spacing</td>
<td></td>
<td>5.45-12.37</td>
</tr>
<tr>
<td>Countercurrent</td>
<td>0.08 m²</td>
<td>3.33-4.33</td>
</tr>
<tr>
<td>3.4 cm plate spacing</td>
<td>0.4 m²</td>
<td>3.77-6.57</td>
</tr>
<tr>
<td>Cocurrent</td>
<td>0.08 m²</td>
<td>2.96-5.99</td>
</tr>
<tr>
<td>subcritical</td>
<td>0.4 m²</td>
<td>3.35-8.85</td>
</tr>
<tr>
<td>Cocurrent</td>
<td>0.08 m²</td>
<td>4.30-5.99</td>
</tr>
<tr>
<td>supercritical</td>
<td>0.4 m²</td>
<td>4.93-11.21</td>
</tr>
</tbody>
</table>

The advantage of a lamella separator over a vertical separator can be clearly seen from the above particularly with the larger floor area in the countercurrent mode with a 1.5 cm plate spacing and in the cocurrent supercritical mode.

6.3 SLUDGE FLOW PERFORMANCE IN THE CONTINUOUS RIG

6.3.1 Introduction

The ability to predict the minimum angle is necessary in order to prevent sludge build-up from occurring. In the following study the minimum angle for sludge flow will be determined experimentally on the continuously operated rig. In some cases, it may not be advantageous to operate at the minimum angle. Steeper angles of operation may be required for a consistent sludge concentration. Thus an additional aim of the study is to examine the characteristics of the sludge flow at various settler angles, lengths, plate spacings and modes of operation and to determine whether the consistency of the sludge concentration is affected. The results of the batch solids tests will also be compared to the continuous tests to find out whether similar modes of flow occur and hence to confirm the applicability of the
batch tests in predicting the mode of solids flow in the continuous mode.

6.3.2 Sludge Flow Performance of Limestone of Size Range 45-53 μm

In the countercurrent mode, the sludge did not flow at angles below 30°. Between 30° to 35° the flow was intermittent. This is undesirable as blockage problems can occur when there is a sudden surge of sludge entering the underflow pipework. The shallowest angle at which a steady flow of sludge occurred was 35°. There was also an upper limit to the angle of inclination of 60° to the horizontal. Above this angle the velocity of the sludge down the slope was so high that it was lifted off the surface and re-entrained.

Two modes of sludge movement were observed - layer and heap (these are described on pp 98-99). Heap flow took place when there was either less hold-up or less sludge settling on the lower surface and hence was observed at steeper angles where the velocity of the sludge will be greater and the hold-up on the surface less. At settler lengths 0.49 and 1.12m, heap movement occurred towards the top end of the lower inclined plate and layer movement at the lower end. At the longer length this is due to the increased re-entrainment of particles from the suspension layer into the clear layer which means that, in order to maintain a low concentration of particles in the overflow, the settler cannot be operated at its optimum for sludge removal and thus most of the solids have settled out in the lower regions of the settler. The amount of solids settling out at the top of the settler was therefore small and not sufficient for layer movement to occur. At the shorter length, however, there is a lower throughput of suspension and hence there will be less solids on the lower inclined surface. Only at the lower end of the settler is there a sufficient amount of solids for layer movement to occur. The onset of heap movement took place at a steeper angle for length 0.95m than for length 0.66m. This was because the throughput of solids at each angle was greater for the longer length and thus enabled layer movement to take place at higher angles.
The applicability of the batch flow tests in predicting not only the minimum angle for sludge flow but also the mode of sludge flow, is apparent from the above.

To recapitulate, in the batch flow tests, heap movement was identified as occurring with thin layers of sludge and at steeper angles than layer movement. The angles at which the various modes of sludge flow occur in the continuous rig is summarised below:

**Table 6.5: Modes of Sludge Flow occurring in Continuous Countercurrent Operation**

<table>
<thead>
<tr>
<th>Settler Length (m)</th>
<th>Settler Angle of Inclination to the Horizontal</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.49</td>
<td>30°</td>
</tr>
<tr>
<td></td>
<td>Heap Movement + Layer Movement</td>
</tr>
<tr>
<td>0.66</td>
<td>&quot; Layer Movement</td>
</tr>
<tr>
<td></td>
<td>Heap Movement</td>
</tr>
<tr>
<td>0.95</td>
<td>&quot; Layer Movement</td>
</tr>
<tr>
<td></td>
<td>Heap Movement</td>
</tr>
<tr>
<td>1.12</td>
<td>&quot; Build-up + Heap movement + Layer Movement</td>
</tr>
<tr>
<td></td>
<td>Heap Movement + Layer Movement</td>
</tr>
</tbody>
</table>

In the cocurrent mode of operation, a very slow moving flow of sludge with a large hold-up occurred between 30° to just below 35°. A steady flow occurred at just below 35°. The limestone started to be lifted off the lower surface at angles above 50° to the horizontal which was 10° below the angle it occurred in the countercurrent mode. This is because the solids enter the main settler section with an initial velocity down the slope and hence travel down the slope at much higher velocities than in the countercurrent mode. At steeper angles the mode of sludge flow was heap movement and at shallower angles it was
layer movement. Heap movement occurred at lower angles with the shorter settler length.

The feed and sludge concentrations were monitored every 15 minutes over a period of two hours. The results are given in Tables A20-A37. In order to determine the degree of variation of sludge concentration with time, a best line was fitted to the data obtained at steady state and the standard deviation of the sludge concentration about the regression line calculated. This gives a numerical value to the consistency of the sludge flow. The results are given below in Table 6.6.

TABLE 6.6: STANDARD DEVIATION OF SLUDGE CONCENTRATION WITH TIME FOR LIMESTONE OF SIZE RANGE 45-53 μm

<table>
<thead>
<tr>
<th>Mode of Operation</th>
<th>Settler Length m</th>
<th>Settler Angle to Horizontal</th>
<th>Standard Deviation of Sludge Concentration (% v/v)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Countercurrent</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>&quot;</td>
<td>0.49</td>
<td>35°</td>
<td>0.064</td>
</tr>
<tr>
<td>&quot;</td>
<td>0.49</td>
<td>45°</td>
<td>0.075</td>
</tr>
<tr>
<td>&quot;</td>
<td>0.49</td>
<td>55°</td>
<td>0.075</td>
</tr>
<tr>
<td>&quot;</td>
<td>0.66</td>
<td>35°</td>
<td>0.066</td>
</tr>
<tr>
<td>&quot;</td>
<td>0.66</td>
<td>45°</td>
<td>0.084</td>
</tr>
<tr>
<td>&quot;</td>
<td>0.66</td>
<td>55°</td>
<td>0.055</td>
</tr>
<tr>
<td>&quot;</td>
<td>0.95</td>
<td>35°</td>
<td>0.050</td>
</tr>
<tr>
<td>&quot;</td>
<td>0.95</td>
<td>45°</td>
<td>0.024</td>
</tr>
<tr>
<td>&quot;</td>
<td>0.95</td>
<td>55°</td>
<td>0.094</td>
</tr>
<tr>
<td>&quot;</td>
<td>1.12</td>
<td>35°</td>
<td>0.089</td>
</tr>
<tr>
<td>&quot;</td>
<td>1.12</td>
<td>45°</td>
<td>0.360</td>
</tr>
<tr>
<td>&quot;</td>
<td>1.12</td>
<td>55°</td>
<td>0.131</td>
</tr>
<tr>
<td>Cocurrent</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>&quot;</td>
<td>0.49</td>
<td>35°</td>
<td>0.040</td>
</tr>
<tr>
<td>&quot;</td>
<td>0.49</td>
<td>45°</td>
<td>0.084</td>
</tr>
<tr>
<td>&quot;</td>
<td>0.49</td>
<td>55°</td>
<td>0.136</td>
</tr>
<tr>
<td>&quot;</td>
<td>0.66</td>
<td>35°</td>
<td>0.025</td>
</tr>
<tr>
<td>&quot;</td>
<td>0.66</td>
<td>45°</td>
<td>0.039</td>
</tr>
<tr>
<td>&quot;</td>
<td>0.66</td>
<td>55°</td>
<td>0.097</td>
</tr>
</tbody>
</table>
At the length 0.49m there was very little difference in the degree of sludge concentration variation. As the angle was increased, the fluctuations in concentration increased slightly at the higher angles. This is in accordance with the experimental observation that the mode of sludge flow remained the same at all angles. At the length 0.66m, it can be seen that there are larger deviations in the sludge flow at 45° than at 35° or 55°. At 45° the onset of heap flow occurred. Below this angle the mode of sludge flow was layer movement. Instabilities in the flow at 45° would be expected until heap movement was properly established as at the steeper angle of 55°. At 0.95m settler length the onset of heap flow was at 55° and this is again reflected by the sudden increase in the standard deviation of the sludge concentration at this angle. At a length of 1.12m build-up was observed to be occurring at the top of the settler at angles below 45°. This stationary layer started to flow at 45° causing instabilities in the flow as indicated by the high standard deviation value. The flow then becomes steadier with the establishment of heap movement as the angle is increased.

As a design guideline, there appears to be an optimum range of settler lengths over which steady sludge flow can be achieved. Larger deviations in the flow occurred at the 0.49 and 1.12m settler lengths. Steady flow is attained when the mode of sludge flow is well established which is either close to the minimum angle where layer movement occurs or at higher angles with heap movement. Care must be taken when operating at the higher angles however since problems can be encountered with solids being lifted off the lower surface which can result in a reduction in efficiency. The ability to predict the minimum angle for flow makes it desirable to operate at the lower angles.

In the cocurrent mode, at both lengths studied, the flow became more unstable at the steeper angles because the solids were being lifted off the lower surface. The flow became more stable at the longer length. This was because there was a greater distance over which the flow could become well established and because there were more solids
travelling down the slope. If the two modes of operation are compared it can be seen that the flow of sludge was more stable in the cocurrent mode at low angles. At steeper angles, because of particle re-entrainment in the cocurrent mode, the countercurrent mode gave a more consistent flow of sludge.

6.3.3 Sludge Flow Behaviour of Limestone Size Range 5-10 μm

In the countercurrent mode, the minimum angle at which satisfactory sludge flow occurred was 40°. The sludge and feed concentration was monitored over a period of 4 hours at an angle of 35°. The sludge concentration remained almost equal to the feed concentration and both concentrations decreased continuously with time indicating continuous build-up and no flow.

At 40° and above, a stationary layer of particles built up on the lower surface of the settler. The thickness of the layer decreased with increasing distance up the settler. At the lower end there was a shear zone from which clumps of particles were continually being sheared. The shear zone was the shape of an inverted 'V' and gradually moved up the lower surface of the settler throughout the test period. Since the sludge build-up did not become excessive, it is thought that some sludge flow occurred on the upper surface of the stationary layer although it was not possible to observe it. After the clumps of particles had sheared off they slid down the lower surface in parallel lines. They were joined by particles which were settling out in the lower regions of the settler. The additional particles were sufficient to cause the mode of sludge flow to develop into heap movement. The region of heap flow was not observed at the length 0.49m and angle 55°, however. This was because the solids throughput was too low. A similar trend in the mode of sludge flow took place with a plate spacing of 1.5 cm.

There are some marked similarities in the modes of sludge flow occurring in the batch and continuous rigs (see Figure 6.59).
At the top end of the built-up layer it is thought that the sludge flow is analogous to that observed in the batch rig at initial sludge concentrations of 0.3-0.5 \% v/v where localised layer movement occurred with the underlying layers remaining stationary. Within the shear zone a critical thickness is reached at which the downwards gravitational force is sufficient for fracture flow to take place. As clumps of particles are sheared off from the sludge layer the sludge above them reaches the critical thickness due to continual settling out of particles. Thus the shear zone gradually moves up the settler. In the batch rig fracture flow only took place at the initial sludge concentrations of 1.0 and 1.5 \% v/v. Below the fracture region in the continuous rig the mode of flow occurring is similar to that observed in the batch rig at initial sludge concentrations of 0.1-0.2 \% v/v. The conditions in the batch rig are slightly different because the sludge layer is of constant thickness. This is why the fracture region in the continuous rig is small. It also explains why the modes of flow occurring below the fracture region in the batch and
continuous rigs are different. In the continuous rig the mass of particles flowing below the fracture region is much less than in the batch rig because not only is the shear zone small but there is also a free surface below it.

In the cocurrent mode the minimum angle at which the sludge would flow was 35°. This was 5° less than the minimum angle in the countercurrent mode. Particles that enter the main body of the settler in the cocurrent mode already possess a velocity in the downwards direction parallel to the lower inclined plate and this provides them with sufficient momentum to flow down the inclined slope at lower angles than in the countercurrent mode.

The sludge flow at 35° was very slow and occurred by the movement of large heaps of solids. These were 'V' shaped and in some cases extended over the width of the plate. In addition there was a large hold-up of solids. This type of flow was observed in the batch rig below the shear zone at the higher initial concentration of 1.0 % v/v. Over the range of angles from 40° to 55° the sludge flowed in heap movement and the hold-up on the lower surface was small. The velocity of the heaps down the slope increased as the inclination angle became steeper. This is analogous to the mode of flow observed in the batch rig at steeper angles.

At start-up in the cocurrent mode it was possible to observe the interface between the sludge and suspension layers. Interfacial waves developed as the sludge moved down the inclined plane (see Figure 6.1). The initiation of the interfacial waves took place closer to the feed inlet as both the settler inclination angle and settler length was increased. At the longer settler lengths the feedrates were higher and thus the particles entering the settler had a higher initial velocity. Hence, they more rapidly attained a velocity at which they lifted off the lower surface. Likewise as the angle of inclination was increased the particles flowed down the lower inclined surface at higher velocities due to the increased gravitational force and hence lifted off the surface at shorter distances from the feed.
inlet. At all angles and lengths the waves grew progressively along the direction of the flow of sludge until the particles were finally ejected into the suspension layer. The magnitude of the waves increased with increasing angle which resulted in particle re-entrainment after the sludge had travelled over much shorter distances.

The feed and sludge concentrations were determined every 20 minutes over a period of about 4 hours. The sludge concentration increased with time up to about 60 minutes. Beyond 60 minutes steady state was reached and the sludge concentration fluctuated about a mean value. The results are given in Tables A38-A68 over a range of angles, settler lengths, plate spacings and modes of flow. As with the larger sized limestone, a best line was fitted to the steady state sludge concentration values and the standard deviations determined. These are given in Table 6.7 overleaf.

The standard deviations of the sludge concentration attained in the countercurrent mode at a plate spacing of 0.034m passed through a minimum with increasing angle at settler lengths 0.66, 0.95 and 1.12m (see Figure 6.60). The largest standard deviations occurred at 40° to the horizontal. This is the minimum angle at which flow would occur and thus the shearing off of clumps of particles is more sporadic.

The standard deviation was particularly high at settler lengths 0.95 and 1.12m and an angle of 40° because the throughput of solids was greatest at these lengths. At the shorter length of 0.49m, however, the standard deviation remained constant with increasing angle until 55° where it underwent a large increase. The reason for the large increase in standard deviation at 55° was because the lowest throughput of sludge occurred with this configuration and the amount of solids settling out was insufficient for a continuous heap flow to develop at the lower end of the settler. Large deviations were also obtained at 55° at length 1.12m. This was because 55° was the steepest angle examined and hence the solids were travelling at their highest velocity. In addition there is a large downwards gravitational
TABLE 6.7: STANDARD DEVIATION OF SLUDGE CONCENTRATION WITH TIME FOR LIMESTONE OF SIZE RANGE 5-10 μm

<table>
<thead>
<tr>
<th>Mode of Operation</th>
<th>Plate Spacing m</th>
<th>Settler Length m</th>
<th>Angle to Horizontal</th>
<th>Standard Deviation of Sludge Concentration % V/V</th>
</tr>
</thead>
<tbody>
<tr>
<td>Counter-</td>
<td>0.034</td>
<td>1.12</td>
<td>40°</td>
<td>0.09895</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>45°</td>
<td>0.01772</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>50°</td>
<td>0.02759</td>
</tr>
<tr>
<td></td>
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<td></td>
<td>55°</td>
<td>0.05743</td>
</tr>
<tr>
<td></td>
<td></td>
<td>0.95</td>
<td>40°</td>
<td>0.1029</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>45°</td>
<td>0.02971</td>
</tr>
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<td></td>
<td></td>
<td>50°</td>
<td>0.02266</td>
</tr>
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<td></td>
<td>55°</td>
<td>0.02575</td>
</tr>
<tr>
<td></td>
<td></td>
<td>0.66</td>
<td>40°</td>
<td>0.04192</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>45°</td>
<td>0.02042</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>50°</td>
<td>0.02725</td>
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<td></td>
<td></td>
<td>55°</td>
<td>0.03396</td>
</tr>
<tr>
<td></td>
<td></td>
<td>0.49</td>
<td>40°</td>
<td>0.02410</td>
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<td>45°</td>
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<td>0.02432</td>
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<td>55°</td>
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<td></td>
<td></td>
<td></td>
<td>55°</td>
<td>0.03460</td>
</tr>
<tr>
<td>Cocurrent</td>
<td>0.034</td>
<td>0.66</td>
<td>35°</td>
<td>0.04273</td>
</tr>
<tr>
<td>sub-critical</td>
<td></td>
<td></td>
<td>40°</td>
<td>0.03288</td>
</tr>
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<td></td>
<td></td>
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<td>45°</td>
<td>0.02372</td>
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<td>0.03268</td>
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<td></td>
<td></td>
<td>55°</td>
<td>0.03564</td>
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<tr>
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<td>0.034</td>
<td>0.66</td>
<td>35°</td>
<td>0.05307</td>
</tr>
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<td>super-critical</td>
<td></td>
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<td>40°</td>
<td>0.02608</td>
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<td></td>
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<td>50°</td>
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<tr>
<td></td>
<td></td>
<td></td>
<td>55°</td>
<td>0.04371</td>
</tr>
</tbody>
</table>
Effect of settler angle in the countercurrent mode on the consistency of the solids concentration in the underflow stream.

Plate spacing = 0.034m
Limestone (5-10μm) in water

![Graph showing the effect of settler angle on solids concentration](image)

FIGURE 6.60
force because the solids throughput is high which also increases the velocity of the sludge. The resulting high velocity of the sludge leads to severe solids re-entrainment.

At angles $45^\circ$ and $50^\circ$ low standard deviations were achieved over the range of settler lengths. This was because problems with re-entrainment and the sporadic flow of particles were minimised.

The sludge flow concentrations were also determined with a settler length of 0.66m in the countercurrent mode at a plate spacing of 0.015m and in the cocurrent subcritical and supercritical modes at a plate spacing of 0.034m. This enabled the effect of the mode of settler operation on the sludge flow to be determined.

In both the cocurrent modes examined, the standard deviation was at a minimum between $40^\circ$ and $50^\circ$. At $35^\circ$ the sludge flowed in large heaps and the hold-up was high. This was responsible for the larger standard deviation at this angle. The higher standard deviation at $55^\circ$ was caused by interfacial turbulence between the sludge and suspension layers which was more severe at steeper angles. This resulted in increased re-entrainment of the particles and hence deviations in the sludge concentration. The higher velocity of the sludge on entering the hopper at this angle could have caused some turbulence which could also have resulted in the increased re-entrainment of particles from the sludge.

At a plate spacing of 0.015m in the countercurrent mode, a high standard deviation value was attained at $45^\circ$. At other angles the standard deviation values were close to those in the other modes.

Figure 6.61 compares the standard deviation in sludge concentration achieved in each mode of operation over a range of angles. It can be seem that there is very little difference between the various modes of operation. The largest differences occur at $45^\circ$ where the standard deviation varies from 0.02-0.045 % v/v. However, when considered as a percentage of the sludge concentration the variation is only 2.5-5.5%.
Effect of the mode of settler operation on the consistency of the solids concentration in the underflow stream

Limestone (5–10μm) in water

FIGURE 6.61
This result is important because it implies that the mode of settler operation has little effect on the sludge flow and hence that the drag force exerted by the suspension layer on the sludge layer is negligible compared to the gravitational force. Thus the main factors for maintaining a consistent sludge concentration are the angle of operation and the length of the settler.
# Chapter 7
SLUDGE FLOW MODEL AND OVERALL DESIGN METHOD

### 7.1 Introduction

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<tbody>
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<td>7.1 INTRODUCTION</td>
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</tbody>
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### 7.2 Prediction of the Minimum Angle of Operation

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<th>Page No</th>
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### 7.3 Proposed Design Scheme

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7.1 **INTRODUCTION**

At present most design procedures are based on assumptions of ideal settling conditions. However non-idealities occur because the necessary requirements for achieving ideal conditions are not met and hence correction factors need to be incorporated into the design. Since the correction factors are generally arbitrary in nature they tend to be excessively large and this results in an uneconomic design that is far below the optimum.

Recently, due to an improved understanding of flow stability and steady state conditions in the settling channel, a design scheme was proposed in which constraints were imposed on relevant design variables in order to suppress the various potential causes of non-idealities. This section aims to extend and improve this design procedure.

A summary of the ideal state constraints and the design variables which they influence is listed below:

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As can be seen from the above table, some of the design variables come under the influence of several design constraints. In these cases the constraint which is most dependent on the variable will have the overriding influence.

The proposed design procedure will also be extended to take into account two potential design situations. These are:

1. when the cost of settler plates and hence plate efficiency is the most important factor;
2. when land area is limited and hence settler compactness is the most important factor.

The angle of inclination is one of the major design variables. The sludge flow constraint is the most dependent on the angle. If the angle is set too low build-up of solids on the lower inclined surface can occur and in extreme cases this can lead to the partial blockage of the settling channel. Thus, it is important that the inclination angle is sufficiently high to satisfy the sludge flow constraint. Based on the results obtained from the batch flow tests a method is proposed for actually determining the minimum angle at which sludge flow will occur. This method can then be incorporated as an initial stage in the design procedure of lamella separators.

7.2 PREDICTION OF THE MINIMUM ANGLE OF OPERATION

As the angle of inclination to the horizontal is reduced in a lamella settler, a point is reached at which the sludge that has settled on the lower plates starts to build up and can eventually fill the entire settling channel. The angle at which this occurs must be just less than the minimum angle of operation of the settler (i.e. the point at which the sludge just starts to flow).

In the following section a method of determining this angle is proposed. Two separate cases will be considered - cocurrent and countercurrent.
From the results of the batch tests on sludge flow it can be seen that it should be possible to describe the sludge flow by a simple frictional model. Justifications for this are discussed below.

i) Mode of flow. Layer movement has been identified as the main type of flow occurring. The angle at which layer movement occurs has been measured. This could be used as the coefficient of friction in the model. Layer movement has been observed as a block of particles sliding over a slow moving underlying layer of particles.

ii) The angle for layer movement appears to reach a limiting value with increasing initial sludge layer thickness. Using a frictional model it would be expected that no change in angle would result from increasing either the layer thickness or its overall weight provided that the drag force was negligible.

iii) The angle at which flow occurs depends on the roughness of the underlying surface material and on the roughness and irregularity of the particles.

iv) The density of the particles does not significantly affect the angle at which flow occurs.

7.2.1 Countercurrent Model

It is assumed that, when the settling particles alight on the lower inclined surface, their component of velocity parallel to the surface is negligible. Thus, at the minimum angle of operation where the sludge is just about to flow, the situation is analogous to that in the batch flow tests.

An additional assumption is that, close to the minimum angle, layer movement is the predominant mode of sludge transport. This assumption is justified for two reasons:
i) In the batch flow tests it was found that layer movement occurred at the lower angles and preceded heap and bulk movement.

ii) Layer movement has been observed in the continuous rig at low angles of operation, both in the 45-53 μm limestone system discussed in Chapter 6 and in work carried out by P.H. Poh. With the limestone of size range 5-10 μm, fracture flow was observed. Layer movement occurred below the fracture point.

In order to use a frictional model a coefficient of frictional resistance, \( \mu_f = \tan\phi \), is required where \( \phi \) is the angle to the horizontal at which motion of the solid on the inclined slope occurs. Referring back to the batch sludge tests it was found that as the thickness of the sludge layer was increased, the angle at which layer movement occurred decreased to a limiting value. It is proposed that \( \mu_f \) is determined from the limiting value at which layer movement occurs since this will be the shallowest angle at which flow can possibly take place.

### 7.2.1.1 Forces acting on the sludge

Consider the typical velocity profile within a settler operating in the countercurrent mode (see Figure 7.1).

From the diagram it can be seen that the velocity of both the suspension and sludge layers is in the downwards direction at the interface separating them and so the drag force on the sludge will be exerted in the downwards direction. Thus, parallel to the inclined plate the force balance is

\[
F_D + F_G \sin\alpha = F_F
\]  

(7.1)

where: \( F_D \) = drag force exerted on sludge layer by suspension layer per unit area
\( F_G \) = gravitational force per unit area
\( F_F \) = frictional force per unit area

This is illustrated diagrammatically in Figure 7.2.

**FIGURE 7.1: TYPICAL VELOCITY PROFILE FOR A SETTLER OPERATING IN THE COUNTERCURRENT MODE**

**FIGURE 7.2: FORCES ACTING ON THE SLUDGE LAYER**
The mathematical development of each force is discussed below.

7.2.1.1.1 Gravitational force

Assuming the effective mass of the sludge layer per unit area to be \( m \), then the gravitational force down the slope can be represented by the following equation:

\[
F_G \sin \alpha = mgs\sin \alpha \tag{7.2}
\]

where: \( g = \) acceleration due to gravity  
\( \alpha = \) angle of inclination to horizontal.

The effective mass per unit area is defined by the equation:

\[
m = \varepsilon T (\rho_p - \rho_f) \tag{7.3}
\]

where: \( \varepsilon = \) volume concentration of solids in the sludge layer  
\( T = \) thickness of the sludge layer  
\( \rho_p - \rho_f = \) effective density  
\( \rho_f = \) density of liquid  
\( \rho_p = \) density of solids

7.2.1.1.2 Frictional force

The standard form of equation for the frictional force between an object and a surface was used in this study. This states that the frictional force is proportional to the normal force which, in this case, is the component of the gravitational force perpendicular to the surface. The constant of proportionality used is the tangent of the angle at which layer movement starts to occur (\( \tan \phi \)).
Thus:

\[ F_F = F_G \cos \alpha \times \tan \phi = mg \cos \alpha \tan \phi \]  

(7.4)

7.2.1.1.3 Drag force

It was assumed that, at the point when the sludge is just about to move, the situation at the sludge/suspension interface is analogous to fluid flow at a wall. The suspension was assumed to behave like a Newtonian fluid with the conditions of no fluid slip at the suspension/sludge interface and constancy of shear.

Thus, the drag force at the wall is

\[ F_D = \frac{\mu du}{dH} \]  

(7.5)

where:  
- \( \mu \) = viscosity of suspension  
- \( u \) = velocity within the suspension  
- \( H \) = width of clear + suspension layer (see Figure 7.3)

The velocity profile was determined from an adaptation of Probstein's work. He developed a mathematical model for lamella separators based on the assumption that the flow in any channel could be treated as a two-dimensional stratified viscous channel flow. Further assumptions that he made were:

1) Uniform density and viscosity within each layer though not necessarily the same as in any other layer  
2) Fluids behave like Newtonian fluids of constant viscosity  
3) No net liquid flux across any plane parallel to the channel walls  
4) Pressure drop is taken to be the same in each of the layers  
5) No fluid slip at wall  
6) Constancy of shear and velocity across the two stratified layer interfaces.
From the assumption of steady state channel flow under the action of gravity, the momentum equation can be written:

\[ 0 = -\frac{\partial p}{\partial x} + \rho g \sin \alpha + \frac{d}{dy} \left( \mu \frac{du}{dy} \right) \quad (7.6) \]

\[ 0 = -\frac{\partial p}{\partial y} + \rho g \cos \alpha \quad (7.7) \]

where:  
- \( x \) = coordinate in direction parallel to settler length  
- \( y \) = coordinate in direction perpendicular to settler length  

(see Figure 7.3)  
- \( p \) = pressure  
- \( \frac{\partial p}{\partial x}, \frac{\partial p}{\partial y} \) = pressure gradient  
- \( \rho \) = density in any layer

![Diagram of coordinate system showing variables used for determining the drag force on the sludge layer](image-url)
From assumption (iv) let:

\[ C_0 = \left( \frac{dP}{dx} \right)_1 = \left( \frac{dP}{dx} \right)_2 \]  

(7.8)

where: \( C_0 \) = constant

numeral 1 represents the clear layer and numeral 2 the hindered layer.

From assumption (i) together with the assumption that the viscosity is the same in the clear and suspension layer, the momentum equation (7.6) can be integrated. This gives

\[ u_1 u_1 = (C_0 - \rho_1 g \sin \alpha) y^2/2 + C_1 y + C_2 \]  

(7.9)

and

\[ u_1 u_2 = (C_0 - \rho_2 g \sin \alpha) y^2/2 + C_3 y + C_4 \]  

(7.10)

where: \( u_1 \) = velocity in clear layer

\( u_2 \) = velocity in suspension layer

\( \rho_1 \) = density of clear layer

\( \rho_2 \) = density of suspension layer

\( C_1, C_2, C_3, C_4 \) = constants of integration

Five boundary conditions define the conditions of no fluid slip at the channel wall or sludge/suspension interface and the conditions of continuous velocity and shear across the stratified layer interfaces.

Thus referring to Figure 7.3:

\[ y = 0 \quad u_1 = 0 \]  

(7.11)
where: \( n \) = thickness of clear liquid layer;  
\( H = \) thickness of clear layer + hindered layer

It is assumed that the sludge on the lower surface (region 3) is just on the point of moving. Thus:

\[ u_3 = 0 \] (7.16)

The constants in equations 7.9 and 7.10 can now be solved. From equation 7.11 we have

\[ C_2 = 0 \] (7.17)

while from equations 7.12 and 7.13

\[ C_4 = \frac{n^2}{2} (\rho_2 - \rho_1)g \sin \alpha \] (7.18)

From the assumption of no liquid flux across any plane where \( x = \) constant we obtain the equation
Using equation 7.19 plus conditions 7.13 and 7.14, we find that:

\[ C_1 = -\frac{n}{H^2} (\rho_2 - \rho_1) g \sin \alpha (H-n)^2 \]  
(7.20)

\[ C_3 = \frac{n^2}{H^2} (\rho_2 - \rho_1)(2H-n)g \sin \alpha \]  
(7.21)

and

\[ C_0 = \rho_2 g \sin \alpha - \frac{n^2}{H^3} (3H-2n)(\rho_2 - \rho_1) \]  
(7.22)

Thus the drag equation of the sludge/suspension interface can be obtained by inserting the values determined for \( C_0, C_3 \) and \( C_4 \) in equation 7.10 and differentiating which gives

\[ \mu_1 \frac{du_2}{dy} = n^2 \frac{(H-n)}{H^2} (\rho_2 - \rho_1)g \sin \alpha \]  
(7.23)

Now \( H = b - T \)  
(7.24)

where: \( b = \) plate spacing  
\( T = \) thickness of sludge layer

Equation 7.23 becomes

\[ \mu_1 \frac{du_2}{dy} = n^2 \frac{(b-T-n)}{(b-T)^2} (\rho_2 - \rho_1)g \sin \alpha \]  
(7.25)
7.2.1.2 Development of model

Substituting expressions for drag (7.25), gravitational (7.2) and frictional forces (7.4) in equation 7.1 and rearranging gives

\[
\tan \alpha = \tan \phi \left[ 1 + \frac{K_1 (b-T) n^2}{(b-T)^2 T} \right]
\]  
\[ (7.26) \]

where

\[
K_1 = \frac{(\rho_2 - \rho_1)}{\varepsilon (\rho_s - \rho_f)}
\]  
\[ (7.27) \]

\[
= \frac{c_0}{\varepsilon}
\]  
\[ (7.28) \]

c_0 = initial concentration of suspension.

Two methods can be adopted in solving the above equation. These will be discussed below.

7.2.1.2.1 Method 1

Letting,

\[
n = b - T - \Delta
\]  
\[ (7.29) \]

in equation 7.26,

where \( \Delta \) = thickness of suspension layer
gives,

\[
\tan \alpha = \tan \phi \left[ 1 + \frac{K_1 \Delta (b-T-\Delta)^2}{T(b-T)^2} \right]
\]  
\[ (7.30) \]
The maximum overflow, \( Q_1 \) will be equal to the flowrate in the clear liquid stream. Thus,

\[
Q_1 = \int_0^\eta u_1 \, dy
\]  
(7.31)

where: \( Q_1 = \) maximum overflow per unit breadth.

Substituting for \( u_1 \) using equation 7.9 gives

\[
Q_1 = \frac{1}{\mu_1} \int_0^\eta [(C_0 - \rho_1 g \sin \alpha) y^2/2 + C_1 y + C_2] \, dy
\]  
(7.32)

Replacing \( C_0, C_1 \) and \( C_2 \) using expressions 7.22, 7.20 and 7.17 respectively and integrating gives,

\[
Q_1 = \frac{1}{\mu_1} (\rho_2 - \rho_1) g \sin \alpha n^3 (n-H)^3/3H^3
\]  
(7.33)

Replacing \( n \) and \( H \) with \( b, t \) and \( \Delta \) via equations 7.24 and 7.29 gives

\[
Q_1 = \frac{1}{\mu_1} (\rho_2 - \rho_1) g \sin \alpha (b-T-\Delta)^3/3(b-T)^3
\]  
(7.34)

Providing steady-state conditions are met and dimensions of the settler suitably chosen then the maximum overflow can be determined from the Nakamura-Kuroda equation. Thus,

\[
Q_1 = \frac{bv}{\sin \alpha} \left( 1 + \frac{h}{b} \cos \alpha \right)
\]  
(7.35)
where: \( v \) = vertical settling velocity of suspension  
and \( h = L \sin \alpha \)

where: \( L \) = length of settler.

Since in most applications \( h/b \) is much greater than 1 then equation 7.35 can be simplified to

\[
Q_1 = v L \cos \alpha \quad (7.37)
\]

Equating the two expressions 7.34 and 7.37 for maximum overflow gives

\[
\tan \alpha = \frac{3vL\mu v (b-T)^3}{g(\rho_2-\rho_1)\Delta^3(b-T-\Delta)^3} \quad (7.38)
\]

We now have two expressions for \( \tan \alpha \), 7.38 and 7.30. Equating them gives

\[
\frac{3vL\mu v (b-T)^3}{g(\rho_2-\rho_1)\Delta^3(b-T-\Delta)^3} = \frac{\tan \phi}{1 + K_1 \left[ \frac{\Delta(b-T-\Delta)^2}{(b-T-\Delta)^2} \right]} \quad (7.39)
\]

7.2.1.2 Method 2

Acrivos and Herbolzheimer derived an expression for \( n \), the thickness of the clear liquid layer where

\[
n = K_2 \left( \frac{1}{3} L \tan \theta \right)^{1/3} \quad (7.40)
\]

and,

\[
K_2 = 3\nu/g c_0 (\rho_p - \rho_f) \quad (7.41)
\]
Substituting 7.40 in 7.26 gives

\[ 1 + \frac{K_1 K_2^{2/3} L^{2/3}}{T(b-T)} \tan \theta - \frac{K_1 K_2 L \tan \theta}{T(b-T)^2} - \tan \phi \tan \theta = 0 \quad (7.42) \]

where: \( \theta = 90 - \alpha \) \quad (7.43)

7.2.2 Comparison of Theoretical Predictions with Experimental Findings

Equation 7.39 in Method 1 will now be solved for a feed suspension of limestone of size range 5-10 \( \mu \text{m} \) dispersed in water where

\[
\begin{align*}
v &= 2.289 \times 10^5 \text{ m/s} \\
\mu_1 &= 1 \times 10^{-3} \text{ kg/ms} \\
g &= 9.81 \text{ m/s}^2 \\
\rho_2 &= 1008.4 \text{ kg/m}^3 \\
\rho_1 &= 997.4 \text{ kg/m}^3 \\
K_1 &= c_0/\epsilon \\
c_0 &= 0.005 (\text{v/v}) \\
\epsilon &= 0.35 \\
K_1 &= 0.0143
\end{align*}
\]

Substituting these values in 7.39 gives

\[ 8.333 \times 10^{-10} \frac{L(b-T)^3}{(b-T-\Delta)^3} \Delta^3 = \frac{\tan \phi}{1 + 0.01429 \frac{\Delta(b-T-\Delta)^2}{(b-T)^2 T}} \quad (7.44) \]

This can be rearranged to the form

\[ A + B\Delta + C\Delta^2 + D\Delta^3 + E\Delta^4 + F\Delta^5 + G\Delta^6 = 0 \quad (7.45) \]
where: \( A = \frac{8.333 \times 10^{-10}}{\tan \phi} (b-T)^2 \); \( B = \frac{1.191 \times 10^{-11}}{T \tan \phi} (b-T)^3 \); 

\( C = \frac{-2.382 \times 10^{-11}}{T \tan \phi} (b-T)^2 \); \( D = \left[ \frac{1.191 \times 10^{-11}}{T \tan \phi} (b-T)^2 \right]^{-1} \)

\( E = 3(b-T)^2; \quad F = 3(b-T); \quad G = 1 \)

Equation 7.45 can be solved for \( \Delta \) using the Newton-Raphson method of iteration for a range of values of \( b, L \) and \( T \). Having obtained the value of \( \Delta \), the angle at which flow will occur, \( \alpha \), can now be determined from either of equations 7.30 or 7.38. \( \alpha \) was determined over the following range of values:

\[ \phi = 40^\circ \Rightarrow \tan \phi = 0.84 \]

\[ b = 0.015 - 0.034 \text{m} \]

\[ T = 0.0001 - 0.02 \text{m} \]

\[ L = 0.49 - 1.12 \text{m} \]

Over the entire range of settler variables above, the calculated value of \( \alpha \) was 40\(^\circ\). Since \( \alpha \) remains constant it can be concluded that the drag of the liquid on the sludge layer has a negligible effect. The experimental value for the minimum angle for sludge flow, obtained on the continuous rig, was in agreement with the predicted value of 40\(^\circ\).

Method 2 will now be used to calculate the minimum angle for flow with the 5-10 \( \mu \text{m} \) limestone system where:

\[ K_1 = 0.01 \]
The thickness of the sludge layer used was that at which, in the batch tests, the angle for flow reached its limiting value. For the limestone system this occurred at an initial concentration of 0.4% v/v. The porosity of the sludge layer was assumed to be 0.5 which gave,

$$T = 0.0004\text{m}$$

Substituting these values in equation 7.42 and with

$$b = 0.034\text{m}$$

gives

$$1 + 6.581 \times 10^{-4} L^{2/3} \tan\theta^{2/3} - 1.842 \times 10^{-5} L \tan\theta - 0.84 \tan\theta = 0$$  \hspace{1cm} (7.46)

Equation 7.46 was solved for a range of values of L. As in method 1, over the entire range of settler variables the angle for flow calculated was 40°.

Both methods were also used to calculate the minimum angle for flow with a system of limestone of size range 45-53 \(\mu\text{m}\) dispersed in water. The limiting angle \(\phi\), at which layer movement occurred, as determined from the batch results, was 36°. The value of the minimum angle calculated was also 36°. Again, the drag force of the liquid appears to have no effect on the angle at which sludge flow occurs. In the continuous rig the minimum angle for flow in the countercurrent system using 45-53 \(\mu\text{m}\) limestone was 35°. This is very close to the predicted value.
7.2.3 Cocurrent Model

In the cocurrent system the particles that enter the settler already possess a velocity in the direction down the slope. Thus the limiting angle of friction used in the countercurrent model is not applicable. It is necessary to determine the angle of sliding friction which will be the same or lower. However, attempts to determine this angle in the batch rig were not successful since reproducible conditions could not be obtained.

The experimental results show that for limestone of size range 45-53 μm, the minimum angle at which sludge flow occurred was the same for both cocurrent and countercurrent modes of operation, whilst for limestone of size range 5-10 μm the minimum angle for sludge flow was 5° lower in the cocurrent system. Since the minimum angle for sludge flow in the countercurrent mode does not appear to be much higher than that in the cocurrent mode according to the experimental results, it would seem reasonable to accept the countercurrent value of minimum angle for the cocurrent mode until a predictive method to determine it is established.

7.3 PROPOSED DESIGN SCHEME

The design scheme outlined below attempts to extend and improve the scheme proposed by Poh. He aimed to suppress the various potential non-idealities as a result of his research findings regarding flow stability and steady state conditions. The prediction of the minimum angle at which sludge will flow is an important step towards reducing non-ideal conditions by reducing the probability of sludge build-up occurring.

The proposed design scheme is divided into two sections: the first outlines the proposed design procedure when the cost of the settler plates and hence plate efficiency is the most important design factor. In the second section the proposed design procedure when the land area is limited is described.
When the cost of the settler plates is the main design criterion the angle of inclination should be as low as possible in order to achieve high settler efficiencies and maximum overflows. However care must be taken when operating at low angles that sludge build-up does not occur on the lower surface of the settler. Thus as a first step in the design procedure the minimum angle of operation must be determined as outlined in the previous sections above. If consistency of the sludge flow is required, then the angle selected should be about 5° above the minimum value. In order to achieve lower angles the type of settler material used should be considered carefully particularly with solids which do not flow easily. For example, in the case of ground glass the difference in the minimum angle for flow over surfaces of PTFE and PVC was more than 10°.

Having provided a suitable angle of inclination and settler material, it is important that a physical constraint is imposed on the channel spacing to ensure that the shear bulk of the sludge being discharged does not create clogging problems. Typical channel spacings range between 5-10 cm. However it was found in the present work that increasing the channel spacing to 8 cm did not lead to an increase in efficiency under all conditions. Thus an upper limit of 8 cm is proposed although further work is necessary to confirm this.

The channel spacing, settler length and inclination angle are under the influence of the steady state and flow stability criterion. However because other constraints are more dependent on the channel spacing and inclination angle this leaves the variable of channel length to satisfy the steady state and flow stability constraints. It has been established in Chapter 6 that for a given angle of inclination and channel spacing there exists a corresponding limiting channel length within which the formation of steady state stratified layers is possible. It is therefore important that the desired channel length is always less than the limiting value. In addition, the flow stability constraint also has a limit on the channel length in order to minimise the re-entrainment of particles into the clear liquid stream. Thus it is proposed that the design of the settler is
based on the more limiting of the two constraints. Since the settler efficiency decreases with increasing plate length, the settler should be designed with as short plates as possible. There is a limit on the minimum value for the settler length, however. Janerus recommended that the length to width ratio should be at least 5:1 to ensure laminar flow conditions. Once the length has been selected the settler width should be determined as described below. If the length to width ratio is less than 5:1 then a higher value for settler length must be selected.

The laminar flow constraint is applied to give the required channel width. The latter should be such that the resulting hydraulic diameter gives a Reynolds number of no greater than 500. To complete the design the required number of channels is specified.

By taking steps to prevent the creation of non-ideal conditions in this way, it is expected that substantial improvements to the overall design can be achieved.

The case of limited land area will now be considered. It has been found that narrower plate spacings give rise to higher maximum overflows in a limited land area due to the potential for a high total projected area available for settling. To reduce the effects of flow instability which will occur with a narrow plate spacing the inclination angle should be low. Thus it will be necessary to determine the minimum angle for flow as a first step. Since the channel spacing is narrow it is important that the physical constraint imposed on the channel spacing is determined accurately to ensure that the sludge being discharged does not block the settler.

As before the channel length is the remaining variable to satisfy the flow stability and steady state criterion. However in this case the settler length should be as close to the limiting value as possible since the projected area increases with settler length. Once the settler length has been determined the channel width can be calculated from the laminar flow constraint and the guideline that the length to
width ratio should be at least 5:1 as was explained in the first section.

Figure 7.4 is a flow chart illustrating the steps taken and the alternative choices available in the proposed design scheme.
FIGURE 7.4: PROPOSED DESIGN SCHEME FOR LAMELLA SEPARATOR
1. A mathematical model was developed which predicted the minimum angle at which solids would flow down the lower inclined surface of a continuous lamella separator operating in the countercurrent mode. The model was based on a frictional analysis, the coefficient of frictional resistance being determined from simple batch tests. There was very good agreement between predicted and experimental results, although it should be noted that extensive tests have not been carried out. The ability to predict the minimum angle should help to ensure that solids are continuously removed and that problems with the build-up of sludge in the settler do not occur.

2. It has been found both from the literature available and from the current experimental results, that the minimum angle required for solids flow in the cocurrent mode was less than or equal to that in the countercurrent mode. Thus, although a model has not been developed to predict the minimum angle for solids flow in the cocurrent mode, the use of the value determined for the countercurrent mode will ensure that solids are continuously removed in the cocurrent mode.

3. The type of material of construction used for the plates significantly affected the minimum angle at which solids flow occurred in the batch tests. The selection of plate material is particularly important for solids with poor flowability properties to enable the settler to be operated at lower angles.

4. The angle at which solids flow occurred was dependent on several parameters. These were:

- particle size
- particle shape and surface texture
- liquid viscosity
- liquid surface tension
- type of settler plate material

5. Four modes of sludge were identified. These occurred under similar conditions in both continuous and batch tests which was further confirmation of the validity of the batch tests in predicting the minimum angle for flow.

6. The consistency of the underflow solids concentration in the continuous mode was neither affected by the mode of settler operation nor by the mode of sludge flow in the systems tested. The solids concentration was most consistent when the mode of flow was well established. It was also dependent on settler length and angle. For both limestone systems in the cocurrent mode at angles of $55^\circ$ to the horizontal and above, lift-off of particles from the lower surface of the settler caused deviations in the solids concentration. Hence there was also an upper limit to the angle at which the settler can be operated.

7. The cocurrent supercritical mode of operation was shown to give rise to higher plate efficiencies and to be a more stable system than both the cocurrent subcritical and countercurrent modes.

8. The value of the Liquid Reynolds Number at the feed inlet in the cocurrent mode was critical if the problem of direct re-entrainment of particles from the feed into the overflow was to be avoided. As a guideline it was found that the Reynolds Number should be under 130 to avoid direct re-entrainment.

9. In the systems tested the optimum settler configuration was shown to be dependent on the most important design criteria:

   - if the land area available was limited, then a settler of narrow plate spacing operating at an angle close to the minimum and at a length of no more than 0.95m gave the higher overflow
rates. This is because the low single plate efficiency of such a settler is compensated for by the high overall projected area.

- if capital is limited and hence single plate efficiency the most important factor, then the cocurrent supercritical mode at angles close to the minimum is the optimum design. However, if a countercurrent model is cheaper and the preferred mode then the settler plates should be of a short length and set at an angle close to the minimum value in order to provide an optimum design.

10. For every settler configuration tested, the overflow rates achievable in a lamella separator were higher than those that would be achieved in a vertical separator of the same floor area. In some cases overflow rates of greater than 10 times were obtained.

11. The Nakamura-Kuroda equation was not able to predict the overflow accurately over the range of conditions tested due to flow instabilities. However with 5-10 μm limestone in the cocurrent supercritical mode between 70-80% agreement was obtained. In the cocurrent subcritical and countercurrent modes of operation it was shown that the predictability of the Nakamura-Kuroda equation could be significantly improved by carefully selecting the settler dimensions to reduce the effects of flow instability and to reduce the Sedimentation Reynolds Number. With 45-53 μm limestone varying the settler dimensions did not result in significant improvements.

12. An optimum ratio (h/b)_{ul} was shown to exist with the countercurrent and cocurrent subcritical modes. It was dependent on plate spacing. The values are given below:
<table>
<thead>
<tr>
<th>Mode</th>
<th>((h/b)_{ul})</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cocurrent subcritical, plate spacing 3.4 cm</td>
<td>19-24</td>
</tr>
<tr>
<td>Countercurrent, plate spacing 3.4 cm</td>
<td>23-25</td>
</tr>
<tr>
<td>Countercurrent, plate spacing 1.5 cm</td>
<td>45-61</td>
</tr>
</tbody>
</table>

Over the range of variables tested with 5-10 μm limestone in the cocurrent supercritical mode and with 45-53 μm limestone in the cocurrent and countercurrent modes, an optimum aspect ratio was not reached, suggesting that \((h/b)_{ul}\) is also dependent on mode of operation and particle size.
CHAPTER 9
RECOMMENDATIONS FOR FUTURE WORK

1. The capability of the procedure proposed to predict the minimum angle at which sludge flow will occur in the countercurrent systems should be tested using particulate systems other than limestone and including flocculating systems and sludges of deformable particles.

2. A procedure to predict the minimum angle for sludge flow in the cocurrent mode should be developed. In addition a procedure to determine the maximum angle of operation at which lift off of particles from the lower surface occurs, should be developed.

3. The sludge flow behaviour of a wider range of particulate systems should be examined in order to determine whether there are any general guidelines for predicting the minimum angle at which sludge flow occurs.

4. Sludge flow should be examined over a wider range of settler materials. Careful selection of settler material can result in improved solids flowability and also lower plate angles of inclination.

5. The possibility of adding a dispersant to improve solids flow ability should be examined.

6. The proposed design guidelines should be tested further by experiments with other fully dispersed systems. In particular, real systems which fulfill the constraints deemed necessary for the accurate prediction of the Nakamura-Kuroda equation should be examined.

7. Flocculated systems should be examined in the continuous rig.
8. The settling efficiency of lamella settlers with shorter plate lengths and a wider range of plate spacings should be determined.

9. The existence of an optimum aspect ratio should be confirmed with other particulate systems in the continuous rig.
APPENDICES

A. OPERATING PERFORMANCE OF THE CONTINUOUS LAMELLA SEPARATOR

A.1 Continuous operation of the limestone of size range 45-53 \( \mu m \) system

Tables A1-A2:
Maximum overflow rates for different modes of operation.

A.2 Continuous operation of the limestone of size range 5-10 \( \mu m \) system

Tables A3-A7:
Maximum overflow rates for different modes of operation.

A.3 Effect of limitations imposed on the settler design with limestone of size range 45-53 \( \mu m \)

Tables A8-A11:
Maximum overflow rates in different floor areas for different modes of operation.

A.4 Effect of limitations imposed on the settler design with limestone of size range 5-10 \( \mu m \)

Tables A12-A19:
Maximum overflow rates in different floor areas for different modes of operation.

A.5 Sludge flow performance of limestone of size range 45-53 \( \mu m \)

Tables A20-A37:
Solids concentration in feed and underflow as a function of operating time for different modes of operation.

A.6 Sludge flow behaviour of limestone of size range 5-10 \( \mu m \)

Tables A58-A68:
Solids concentration in feed and underflow as a function of operating time for different modes of operation.
<table>
<thead>
<tr>
<th>Settler Length (m)</th>
<th>Angle of Inclination to Horizontal (°)</th>
<th>Aspect Ratio A = h/b</th>
<th>R</th>
<th>Maximum Overflow Rate Q₀ (l/min)</th>
<th>Settler Efficiency (Q₀)expt / (Q₀)theo x100%</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td>Predicted (Q₀)theo</td>
<td>Experimental (Q₀)expt</td>
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<td>1699</td>
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</table>

Note: Maximum Overflow Rate for Countercurrent Flow

Value range: 45-53 μm
Number of settling channels = 1
Channel spacing, b = 3.4 cm
Channel width, w = 4 cm

The values were calculated using the Wespe equation (3.21).
### TABLE A2
MAXIMUM OVERFLOW RATE FOR COCURRENT MODE

<table>
<thead>
<tr>
<th>Settler Length</th>
<th>Angle of Inclination to Horizontal</th>
<th>Aspect Ratio h/b</th>
<th>A</th>
<th>R</th>
<th>Maximum Overflow Q_o 1/min</th>
<th>Settler Efficiency (Q_o^expt/100)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
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<td></td>
<td></td>
<td></td>
<td>Predicted* (Q_o^theo)</td>
<td>Experimental (Q_o^expt)</td>
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<td>0.9</td>
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<td>0.8</td>
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<td></td>
<td>1.76</td>
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* Using the Nakamura-Kuroda equation (3.21)
### TABLE A3
MAXIMUM OVERFLOW RATE FOR COUNTERCURRENT FLOW

Particle size range = 5-10 μm
Number of settling channels = 1
Channel spacing, b = 3.4 cm
Channel width, w = 4 cm

<table>
<thead>
<tr>
<th>Settler Length m</th>
<th>Angle of Inclination to Horizontal α°</th>
<th>Aspect Ratio h/b</th>
<th>A (10^8)</th>
<th>R</th>
<th>Maximum Overflow (Q_o) cc/min</th>
<th>Settler Efficiency ((Q_o)_{expt} \times 100%)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>Predicted* ((Q_o)_{theo})</td>
<td>Experimental ((Q_o)_{expt})</td>
</tr>
<tr>
<td>0.49</td>
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<td>21.00</td>
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</table>

* Using the Nakamura-Kuroda equation (3.21)
### TABLE A4

**MAXIMUM OVERFLOW RATE FOR COCURRENT SUBCRITICAL FLOW**

- Particle size range = 5-10 μm
- Number of settling channels = 1
- Channel spacing, b = 3.4 cm
- Channel width, W = 4 cm

<table>
<thead>
<tr>
<th>Settler Length (m)</th>
<th>Angle of Inclination to Horizontal (°)</th>
<th>Aspect Ratio (h/b)</th>
<th>A (mm)</th>
<th>R (mm)</th>
<th>Maximum Overflow Rate (Qo) cc/min</th>
<th>Settler Efficiency (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>Predicted (Qo)theo</td>
<td>Experimental (Qo)expt</td>
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</tr>
<tr>
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TABLE A5
MAXIMUM OVERFLOW RATE FOR COCURRENT SUPERCritical FLOW

<table>
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<th>Settler Length m</th>
<th>Angle of Inclination to Horizontal α°</th>
<th>Aspect Ratio h/b</th>
<th>A</th>
<th>R</th>
<th>Maximum Overflow Q₀ cc/min</th>
<th>Settler Efficiency (Q₀&lt;sub&gt;expt&lt;/sub&gt;/Q₀&lt;sub&gt;theo&lt;/sub&gt;) x 100%</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>Predicted* (Q₀&lt;sub&gt;theo&lt;/sub&gt;)</td>
<td>Experimental (Q₀&lt;sub&gt;expt&lt;/sub&gt;)</td>
</tr>
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<td>0.49</td>
<td>35</td>
<td>8.27</td>
<td>2.84 x 10⁶</td>
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* Using the Nakamura-Kuroda equation (3.21)
**TABLE A7**

**MAXIMUM OVERFLOW RATE FOR COUNTERCURRENT FLOW**

Particle size range = 5-10 μm  
Number of settling channels = 1  
Channel spacing, b = 1.6 cm  
Channel width, W = 4 cm

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*Using the Nakamura-Kuroda equation (3.21)
**Table A8**

**Maximum Overflow Rate for Countercurrent Flow in a Floor Area of 0.4 m²**

- Particle size range = 45-53 μm
- Channel spacing, b = 3.4 cm
- Channel width, W = 4 cm
- Plate thickness = 0.4 cm

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*Using the Nakamura-Kuroda equation (3.21) and equation 6.5*
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*Using the Hofmeister-Kopač equation (3.21) and equation 5.5.
TABLE A10
MAXIMUM OVERFLOW RATE FOR COCURRENT FLOW IN A FLOOR AREA OF 0.08 m²

Particle size range = 45-53 μm
Channel spacing, b = 3.4 cm
Channel width, w = 4 cm
Plate thickness = 0.4 cm

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<th>Total Projected area m²</th>
<th>Maximum Overflow 1/min</th>
<th>Inclined Vertical Overflow</th>
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* Using the Nakamura-Kuroda equation (3.21) and equation 6.5
TABLE A11
MAXIMUM OVERFLOW RATE FOR COCURRENT FLOW IN A FLOOR AREA OF 0.4 m²

Particle size range = 45-53 μm
Channel spacing, b = 3.4 cm
Channel width, w = 4 cm
Plate thickness = 0.4 cm

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<th>Total No. of Plates Accommodated</th>
<th>Total Projected Area m²</th>
<th>Maximum Overflow 1/min</th>
<th>Inclined Vertical Overflow</th>
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* Using the Nakamura-Kuroda equation (3.21) and equation 6.5
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<th>Angle of Inclination to Horizontal $\alpha^0$</th>
<th>Overflow per Unit Projected Area m/min</th>
<th>Total No. of Plates Accommodated</th>
<th>Total Projected Area m²</th>
<th>Maximum Overflow cc/min Predicted*</th>
<th>Experimental</th>
<th>Inclined Vertical Overflow</th>
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*Using the Nakamura-Kuroda equation (3.21) and equation 6.5
TABLE A13
MAXIMUM OVERFLOW RATE FOR COUNTERCURRENT FLOW IN A FLOOR AREA OF 0.08 m²

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<th>Total No. of Plates Accommodated</th>
<th>Total Projected Area m²</th>
<th>Maximum Overflow cc/min</th>
<th>Inclined Vertical Overflow</th>
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* Using the Nakamura-Kuroda equation (3.21) and equation 6.5

Particle size range = 5-10 μm
Channel spacing, b = 1.5 cm
Channel width, W = 4 cm
Plate thickness = 0.4 cm
<table>
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<th>Angle of Inclination to Horizontal $\alpha^\circ$</th>
<th>Overflow per Unit Predicted Area m/m/min</th>
<th>Total No. of Plates Accommodated</th>
<th>Total Projected Area m$^2$</th>
<th>Maximum Overflow cc/min</th>
<th>Inclined Vertical Overflow</th>
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<td>Predicted*</td>
<td>Experimental</td>
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* Using the Nakamura-Kuroda equation (3.21) and equation 6.5
### TABLE A15

**MAXIMUM OVERFLOW RATE FOR COCURRENT SUPERCritical FLOW IN A FLOOR AREA OF 0.08 m²**

- Particle size range = 5-10 μm
- Channel spacing, b = 3.4 cm
- Channel width, W = 4 cm
- Plate thickness = 0.4 cm

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<tr>
<th>Settler Length m</th>
<th>Angle of Inclination to Horizontal α°</th>
<th>Overflow per Unit Projected Area m/min</th>
<th>Total No. of Plates Accommodated</th>
<th>Total Projected Area m²</th>
<th>Maximum Overflow cc/min Predicted*</th>
<th>Experimental</th>
<th>Inclined Vertical Overflow</th>
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<td>0.49</td>
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* Using the Nakamura-Kuroda equation (3.91) and equation 6.5.
### Table A16

Maximum Overflow Rate for Countercurrent Flow in a Floor Area of 0.4 m²

- Particle size range = 5-10 μm
- Channel spacing, \( b \) = 3.4 cm
- Channel width, \( W \) = 4 cm
- Plate thickness = 0.4 cm

| Settler Length \( m \) | Angle of Inclination to Horizontal \( \alpha^0 \) | Overflow per Unit Projected Area \( \text{m} / \text{min} \) | Total No. of Plates Accommodated | Total Projected Area \( \text{m}^2 \) | Maximum Overflow \( \text{cc/min} \) Predicted* | Experimental | Inclined | Vertical Overflow |
|-----------------------|-----------------------------------------------|-------------------------------------------------|-----------------------------------|---------------------------------|-----------------|---------------|-------------|
| 0.49                  | 40                                            | \( 9.59 \times 10^{-4} \)                         | 163                               | 2.45                            | 3831            | 2347         | 4.27        | 4.24        |
|                       | 45                                            | \( 9.38 \times 10^{-4} \)                         | 179                               | 2.48                            | 3884            | 2327         | 4.32        | 4.12        |
|                       | 50                                            | \( 9.21 \times 10^{-4} \)                         | 195                               | 2.45                            | 3842            | 2262         | 4.12        | 4.12        |
|                       | 55                                            | \( 8.81 \times 10^{-4} \)                         | 209                               | 2.35                            | 3699            | 2069         | 3.77        | 3.77        |
| 0.66                  | 40                                            | \( 8.65 \times 10^{-4} \)                         | 160                               | 3.24                            | 4912            | 2860         | 5.10        | 5.10        |
|                       | 45                                            | \( 8.30 \times 10^{-4} \)                         | 177                               | 3.30                            | 5069            | 2744         | 4.99        | 4.99        |
|                       | 50                                            | \( 8.25 \times 10^{-4} \)                         | 193                               | 3.28                            | 4960            | 2702         | 4.92        | 4.92        |
|                       | 55                                            | \( 8.38 \times 10^{-4} \)                         | 207                               | 3.15                            | 4782            | 2629         | 4.79        | 4.79        |
| 0.95                  | 40                                            | \( 7.39 \times 10^{-4} \)                         | 157                               | 4.57                            | 6735            | 3376         | 6.15        | 6.14        |
|                       | 45                                            | \( 7.26 \times 10^{-4} \)                         | 173                               | 4.65                            | 6834            | 3374         | 6.15        | 6.14        |
|                       | 50                                            | \( 6.96 \times 10^{-4} \)                         | 189                               | 4.62                            | 6804            | 3213         | 5.85        | 5.85        |
|                       | 55                                            | \( 6.69 \times 10^{-4} \)                         | 204                               | 4.45                            | 6569            | 2958         | 5.38        | 5.38        |
| 1.12                  | 40                                            | \( 6.70 \times 10^{-4} \)                         | 154                               | 5.29                            | 7700            | 3542         | 6.45        | 6.45        |
|                       | 45                                            | \( 6.66 \times 10^{-4} \)                         | 171                               | 5.42                            | 7883            | 3608         | 6.57        | 6.57        |
|                       | 50                                            | \( 6.11 \times 10^{-4} \)                         | 187                               | 5.39                            | 7854            | 3291         | 5.99        | 5.99        |
|                       | 55                                            | \( 5.45 \times 10^{-4} \)                         | 201                               | 5.16                            | 7558            | 2814         | 5.12        | 5.12        |

*Using the Nakamura-Kuroda equation (3.21) and equation 6.5
<table>
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<tr>
<th>Settler Length m</th>
<th>Angle of Inclination to Horizontal $\alpha^\circ$</th>
<th>Overflow per Unit Projected Area m/min</th>
<th>Total No. of Plates Accommodated</th>
<th>Total Projected Area $m^2$</th>
<th>Maximum Overflow cc/min</th>
<th>Inclined Vertical Overflow</th>
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*Using the Nakamura-Kuroda equation (3.21) and equation 6.5
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<th>Angle of Inclination to Horizontal (°)</th>
<th>Overflow per Unit Projected Area (m/min)</th>
<th>Total No. of Plates Accommodated</th>
<th>Total Projected Area (m²)</th>
<th>Maximum Overflow (cc/min) Predicted</th>
<th>Experimental</th>
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*Using the Nakamura-Kuroda equation (3.21) and equation 6.5
TABLE A19

MAXIMUM OVERFLOW RATE FOR COCURRENT SUPERCritical FLOW IN A FLOOR AREA OF 0.4 m²

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<th>Overflow per Unit Projected Area m/min</th>
<th>Total No. of Plates Accommodated</th>
<th>Total Projected Area m²</th>
<th>Maximum Overflow cc/min</th>
<th>Inclined Vertical Overflow</th>
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*Using the Nakamura-Kuroda equation (3.21) and equation 6.5
TABLE A20

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

Particle size range = 45-53 \textmu m
Channel length = 0.49 m
Inclination angle to horizontal = 35^\circ
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

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<th>Maximum Overflow Rate ((Q_o)_{expt}) l/min</th>
<th>Underflow Rate ((Q_u)_{expt}) l/min</th>
<th>Concentration of Solids in Feed, (c_0) % v/v</th>
<th>Concentration of Solids in Underflow, (c_u) % v/v</th>
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<tr>
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<td>0.26</td>
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<td>0.26</td>
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</table>
TABLE A21

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

Particle size range = 45-53 μm
Channel length = 0.49 m
Inclination angle to horizontal = 45°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate $(Q_o)_{\text{expt}}$ l/min</th>
<th>Underflow Rate $(Q_u)_{\text{expt}}$ l/min</th>
<th>Concentration of Solids in Feed, $c_o$ % v/v</th>
<th>Concentration of Solids in Underflow, $c_u$ % v/v</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
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<td>0.435</td>
<td>0.0</td>
</tr>
<tr>
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<td>0.326</td>
<td>1.669</td>
</tr>
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<td>0.25</td>
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<tr>
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<td>1.503</td>
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<td>0.25</td>
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TABLE A22

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTCURRENT MODE

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{\text{expt}}) l/min</th>
<th>Underflow Rate ((Q_u)_{\text{expt}}) l/min</th>
<th>Concentration of Solids in Feed, (c_o) % v/v</th>
<th>Concentration of Solids in Underflow, (c_u) % v/v</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>0.7</td>
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<td>0.5</td>
<td>0.0</td>
</tr>
<tr>
<td>15</td>
<td>0.7</td>
<td>0.23</td>
<td>0.393</td>
<td>1.921</td>
</tr>
<tr>
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<td>0.23</td>
<td>0.459</td>
<td>1.860</td>
</tr>
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<td>0.23</td>
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<td>0.23</td>
<td>0.436</td>
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</tr>
<tr>
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<td>0.23</td>
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<tr>
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<td>0.7</td>
<td>0.23</td>
<td>0.387</td>
<td>1.768</td>
</tr>
</tbody>
</table>

Particle size range = 45-53 μm
Channel length = 0.49 m
Inclination angle to horizontal = 55°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1
TABLE A23

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ( (Q_o)_{\text{expt}} ) l/min</th>
<th>Underflow Overflow Rate ( (Q_u)_{\text{expt}} ) l/min</th>
<th>Concentration of Solids in Feed, ( c_o ) % V/V</th>
<th>Concentration of Solids in Underflow, ( c_u ) % V/V</th>
</tr>
</thead>
<tbody>
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<td>0.542</td>
<td>0.0</td>
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<td>0.33</td>
<td>0.455</td>
<td>1.861</td>
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<td>0.33</td>
<td>0.444</td>
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<td>0.430</td>
<td>1.652</td>
</tr>
<tr>
<td>75</td>
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<td>0.33</td>
<td>0.423</td>
<td>1.641</td>
</tr>
<tr>
<td>90</td>
<td>1.0</td>
<td>0.33</td>
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<td>1.790</td>
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<tr>
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<td>1.0</td>
<td>0.33</td>
<td>0.450</td>
<td>1.659</td>
</tr>
<tr>
<td>120</td>
<td>1.0</td>
<td>0.33</td>
<td>0.413</td>
<td>1.653</td>
</tr>
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</table>

Particle size range = 45-53 \( \mu \)m
Channel length = 0.66
Inclination angle to horizontal = 35°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1
TABLE A24

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

Particle size range = 45-53 μm
Channel length = 0.66 m
Inclination angle to horizontal = 45°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate (Q₀)expt l/min</th>
<th>Underflow Rate (Qₚ)expt l/min</th>
<th>Concentration of Solids in Feed, c₀ % V/V</th>
<th>Concentration of Solids in Underflow, cᵤ % V/V</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>0.9</td>
<td>0.3</td>
<td>0.458</td>
<td>0.0</td>
</tr>
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<td>0.3</td>
<td>0.410</td>
<td>1.593</td>
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<tr>
<td>45</td>
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<td>0.3</td>
<td>0.370</td>
<td>1.599</td>
</tr>
<tr>
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<td>0.3</td>
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<td>1.721</td>
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<td>0.362</td>
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<tr>
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<td>0.3</td>
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<tr>
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<td>0.3</td>
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<tr>
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</table>
TABLE A25

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

Particle size range = 45-53 μm
Channel length = 0.66 m
Inclination angle to horizontal = 55°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate $(Q_o)_{expt}$ l/min</th>
<th>Underflow Rate $(Q_u)_{expt}$ l/min</th>
<th>Concentration of Solids in Feed, $c_o$ % v/v</th>
<th>Concentration of Solids in Underflow, $c_u$ % v/v</th>
</tr>
</thead>
<tbody>
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<td>0.26</td>
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<td>0.26</td>
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<td>0.26</td>
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<td>2.134</td>
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<td>0.26</td>
<td>0.481</td>
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<td>0.26</td>
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TABLE A26

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

Particle size range = 45-53 μm
Channel length = 0.95 m
Inclination angle to horizontal = 35°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{expt}) l/min</th>
<th>Underflow Rate ((Q_u)_{expt}) l/min</th>
<th>Concentration of Solids in Feed, (c_0) % v/v</th>
<th>Concentration of Solids in Underflow, (c_u) % v/v</th>
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</thead>
<tbody>
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<td>0</td>
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<td>0.51</td>
<td>0.230</td>
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<tr>
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<td>0.51</td>
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<tr>
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TABLE A27

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

<table>
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<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{\text{expt}} ) l/min</th>
<th>Underflow Rate ((Q_u)_{\text{expt}} ) l/min</th>
<th>Concentration of Solids in Feed, (c_0) % V/V</th>
<th>Concentration of Solids in Underflow, (c_u) % V/V</th>
</tr>
</thead>
<tbody>
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<td>0.0</td>
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<td>1.3</td>
<td>0.43</td>
<td>0.328</td>
<td>1.264</td>
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<td>1.193</td>
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Particle size range = 45-53 \(\mu\)m
Channel length = 0.95 m
Inclination angle to horizontal = 45°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1
TABLE A28

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate $Q_0$expt l/min</th>
<th>Underflow Rate $Q_u$expt l/min</th>
<th>Concentration of Solids in Feed, $c_0$ % v/v</th>
<th>Concentration of Solids in Underflow, $c_u$ % v/v</th>
</tr>
</thead>
<tbody>
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<td>0.500</td>
<td>0.0</td>
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<td>0.35</td>
<td>0.393</td>
<td>1.692</td>
</tr>
<tr>
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<td>0.35</td>
<td>0.423</td>
<td>1.604</td>
</tr>
<tr>
<td>90</td>
<td>1.05</td>
<td>0.35</td>
<td>0.396</td>
<td>-</td>
</tr>
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<td>105</td>
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<td>0.35</td>
<td>0.336</td>
<td>1.607</td>
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<td>1.05</td>
<td>0.35</td>
<td>0.383</td>
<td>1.493</td>
</tr>
</tbody>
</table>

- Particle size range = 45-53 μm
- Channel length = 0.95m
- Inclination angle to horizontal = 55°
- Channel spacing = 3.4 cm
- Channel width = 4 cm
- Number of settling channels = 1
TABLE A29
SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

Particle size range = 45-53 μm
Settler length = 1.12 m
Inclination angle to horizontal = 35°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{expt}) l/min</th>
<th>Underflow Rate ((Q_u)_{expt}) l/min</th>
<th>Concentration of Solids in Feed, (c_0) % V/V</th>
<th>Concentration of Solids in Underflow, (c_u) % V/V</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
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<td>0.58</td>
<td>0.524</td>
<td>0.0</td>
</tr>
<tr>
<td>15</td>
<td>1.75</td>
<td>0.58</td>
<td>0.394</td>
<td>1.998</td>
</tr>
<tr>
<td>30</td>
<td>1.75</td>
<td>0.58</td>
<td>0.414</td>
<td>1.825</td>
</tr>
<tr>
<td>45</td>
<td>1.75</td>
<td>0.58</td>
<td>0.419</td>
<td>1.840</td>
</tr>
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<td>60</td>
<td>1.75</td>
<td>0.58</td>
<td>0.408</td>
<td>1.639</td>
</tr>
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<td>1.75</td>
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<td>0.400</td>
<td>1.720</td>
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<td>90</td>
<td>1.75</td>
<td>0.58</td>
<td>0.394</td>
<td>1.798</td>
</tr>
<tr>
<td>105</td>
<td>1.75</td>
<td>0.58</td>
<td>0.388</td>
<td>1.764</td>
</tr>
<tr>
<td>120</td>
<td>1.75</td>
<td>0.58</td>
<td>0.370</td>
<td>1.697</td>
</tr>
</tbody>
</table>
TABLE A30
SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

Particle size range = 45-53 μm
Settler length = 1.12 m
Inclination angle to horizontal = 45°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_0)_{\text{expt}}) l/min</th>
<th>Underflow Rate ((Q_U)_{\text{expt}}) l/min</th>
<th>Concentration of Solids in Feed, (c_0) % v/v</th>
<th>Concentration of Solids in Underflow, (c_u) % v/v</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>1.45</td>
<td>0.48</td>
<td>0.5</td>
<td>0.0</td>
</tr>
<tr>
<td>15</td>
<td>1.45</td>
<td>0.48</td>
<td>0.302</td>
<td>1.920</td>
</tr>
<tr>
<td>30</td>
<td>1.45</td>
<td>0.48</td>
<td>0.301</td>
<td>-</td>
</tr>
<tr>
<td>45</td>
<td>1.45</td>
<td>0.48</td>
<td>0.326</td>
<td>1.506</td>
</tr>
<tr>
<td>60</td>
<td>1.45</td>
<td>0.48</td>
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</tr>
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<td>0.48</td>
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</table>
### TABLE A31

**SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE**

- Particle size range = 45-53 μm
- Settler length = 1.12 m
- Inclination angle to horizontal = 55°
- Channel spacing = 3.4 cm
- Channel width = 4 cm
- Number of settling channels = 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_0)_{\text{expt}}) l/min</th>
<th>Underflow Rate ((Q_u)_{\text{expt}}) l/min</th>
<th>Concentration of Solids in Feed, (c_0) % v/v</th>
<th>Concentration of Solids in Underflow, (c_u) % v/v</th>
</tr>
</thead>
<tbody>
<tr>
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<td>0.5</td>
<td>0.0</td>
</tr>
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<td>0.4</td>
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<td>0.314</td>
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</table>
**TABLE A32**

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COCURRENT MODE

- Particle size range = 45-53 μm
- Settler length = 0.49 m
- Inclination angle to horizontal = 35°
- Channel spacing = 3.4 cm
- Channel width = 4 cm
- Number of settling channels = 1

<table>
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<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_0)_{expt}) l/min</th>
<th>Underflow Rate ((Q_u)_{expt}) l/min</th>
<th>Concentration of Solids in Feed, (c_0) % v/v</th>
<th>Concentration of Solids in Underflow, (c_u) % v/v</th>
</tr>
</thead>
<tbody>
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TABLE A33

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COCURRENT MODE

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate $(Q_o)_{expt}$ l/min</th>
<th>Underflow Rate $(Q_u)_{expt}$ l/min</th>
<th>Concentration of Solids in Feed, $c_o$ % v/v</th>
<th>Concentration of Solids in Underflow, $c_u$ % v/v</th>
</tr>
</thead>
<tbody>
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<td>0.500</td>
<td>0.0</td>
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<tr>
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<td>0.23</td>
<td>0.393</td>
<td>-</td>
</tr>
<tr>
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<td>0.7</td>
<td>0.23</td>
<td>0.311</td>
<td>1.959</td>
</tr>
<tr>
<td>75</td>
<td>0.7</td>
<td>0.23</td>
<td>0.351</td>
<td>1.835</td>
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<td>0.7</td>
<td>0.23</td>
<td>0.349</td>
<td>1.663</td>
</tr>
<tr>
<td>105</td>
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<td>0.23</td>
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<td>1.751</td>
</tr>
<tr>
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<td>0.7</td>
<td>0.23</td>
<td>0.304</td>
<td>1.643</td>
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</table>

Particle size range = 45-53 μm
Settler length = 0.49 m
Inclination angle to horizontal = 45°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1
TABLE A34
SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COCURRENT MODE

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ( (Q_0)_{\text{expt}} ) l/min</th>
<th>Underflow Rate ( (Q_u)_{\text{expt}} ) l/min</th>
<th>Concentration of Solids in Feed, ( c_0 ) % v/v</th>
<th>Concentration of Solids in Underflow, ( c_u ) % v/v</th>
</tr>
</thead>
<tbody>
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<tr>
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<tr>
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<td>0.16</td>
<td>0.329</td>
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</tr>
<tr>
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<td>0.16</td>
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<td>0.16</td>
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<td>1.875</td>
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</table>

Particle size range = 45-53 \( \mu m \)
Settler length = 0.49 m
Inclination angle to horizontal = 55°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1
TABLE A35

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COCURRENT MODE

Particle size range = 45-53 μm
Settler length = 0.66 m
Inclination angle to horizontal = 35°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{expt}) l/min</th>
<th>Underflow Rate ((Q_u)_{expt}) l/min</th>
<th>Concentration of Solids in Feed, (c_o) % v/v</th>
<th>Concentration of Solids in Underflow, (c_u) % v/v</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
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<td>0.500</td>
<td>0.0</td>
</tr>
<tr>
<td>15</td>
<td>1.1</td>
<td>0.36</td>
<td>0.335</td>
<td>1.763</td>
</tr>
<tr>
<td>30</td>
<td>1.1</td>
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</tr>
<tr>
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<td>0.36</td>
<td>0.314</td>
<td>1.472</td>
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<td>0.315</td>
<td>1.405</td>
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<tr>
<td>90</td>
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<td>0.36</td>
<td>0.293</td>
<td>1.348</td>
</tr>
<tr>
<td>105</td>
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<td>0.274</td>
<td>1.348</td>
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<td>0.262</td>
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</table>
### TABLE A36

**SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COCURRENT MODE**

- Particle size range = 45-53 μm
- Settler length = 0.66 m
- Inclination angle to horizontal = 45°
- Channel spacing = 3.4 cm
- Channel width = 4 cm
- Number of settling channels = 1

<table>
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<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{expt}) l/min</th>
<th>Underflow Rate ((Q_u)_{expt}) l/min</th>
<th>Concentration of Solids in Feed, (c_o) % V/V</th>
<th>Concentration of Solids in Underflow, (c_u) % V/V</th>
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</thead>
<tbody>
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<td>0</td>
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</tr>
<tr>
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</tr>
<tr>
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## Table A37

**Solids Concentration in Feed and Underflow as a Function of Operating Time in the Cocurrent Mode**

Particle size range = 45-53 μm  
Settler length = 0.66 m  
Inclination angle to horizontal = 55°  
Channel spacing = 3.4 cm  
Channel width = 4 cm  
Number of settling channels = 1

<table>
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<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{expt}) l/min</th>
<th>Underflow Rate ((Q_u)_{expt}) l/min</th>
<th>Concentration of Solids in Feed, (c_o) % v/v</th>
<th>Concentration of Solids in Underflow, (c_u) % v/v</th>
</tr>
</thead>
<tbody>
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TABLE A38

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

Particle size range = 5-10 µm
Channel length = 0.49m
Inclination angle to horizontal = 40°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_0)_{\text{expt}}) cc/min</th>
<th>Underflow Rate ((Q_u)_{\text{expt}}) cc/min</th>
<th>Concentration of Solids in Feed, (c_0) % v/v</th>
<th>Concentration of Solids in Underflow, (c_u) % v/v</th>
</tr>
</thead>
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<td>0.780</td>
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TABLE A39
SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{\text{expt}}) cc/min</th>
<th>Underflow Rate ((Q_u)_{\text{expt}}) cc/min</th>
<th>Concentration of Solids in Feed, (c_0) % v/v</th>
<th>Concentration of Solids in Underflow, (c_u) % v/v</th>
</tr>
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</table>

Particle size range = 5-10 μm
Settler length = 0.49 m
Inclination angle to horizontal = 45°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1
TABLE A40
SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

Particle size range = 5-10 µm
Channel length = 0.49 m
Inclination angle to horizontal = 50°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{expt}) cc/min</th>
<th>Underflow Rate ((Q_u)_{expt}) cc/min</th>
<th>Concentration of Solids in Feed, (c_o) % V/V</th>
<th>Concentration of Solids in Underflow, (c_u) % V/V</th>
</tr>
</thead>
<tbody>
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</tr>
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<td>11.65</td>
<td>0.462</td>
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</tr>
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</tr>
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<td>0.463</td>
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<td>0.463</td>
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</table>
### TABLE A41

**SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE**

- Particle size range: 5-10 µm
- Channel length: 0.49 m
- Inclination angle to horizontal: 55°
- Channel spacing: 3.4 cm
- Channel width: 4 cm
- Number of settling channels: 1

<table>
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<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate $(Q_o)_{expt}$ cc/min</th>
<th>Underflow Rate $(Q_u)_{expt}$ cc/min</th>
<th>Concentration of Solids in Feed, $c_0$ % v/v</th>
<th>Concentration of Solids in Underflow, $c_u$ % v/v</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
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<td>9.96</td>
<td>0.5</td>
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</tr>
<tr>
<td>30</td>
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<td>9.96</td>
<td>0.476</td>
<td>0.542</td>
</tr>
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<td>9.96</td>
<td>-</td>
<td>0.781</td>
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<td>9.96</td>
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<td>0.456</td>
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<td>9.96</td>
<td>0.459</td>
<td>0.890</td>
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<tr>
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<td>9.96</td>
<td>0.458</td>
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<tr>
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<td>9.96</td>
<td>0.457</td>
<td>0.901</td>
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<td>9.96</td>
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<td>0.881</td>
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<tr>
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<td>9.96</td>
<td>0.451</td>
<td>0.827</td>
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<td>0.454</td>
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<td>0.792</td>
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<td>9.96</td>
<td>0.45</td>
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</tr>
</tbody>
</table>
### TABLE A42

**SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE**

- Particle size range: 5-10 μm
- Channel length: 0.66 m
- Inclination angle to horizontal: 40°
- Channel spacing: 3.4 cm
- Channel width: 4 cm
- Number of settling channels: 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ($Q_o$) expt cc/min</th>
<th>Underflow Rate ($Q_u$) expt cc/min</th>
<th>Concentration of Solids in Feed, $c_o$ % v/v</th>
<th>Concentration of Solids in Underflow, $c_u$ % v/v</th>
</tr>
</thead>
<tbody>
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<td>0</td>
<td>18.66</td>
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<td>0.0</td>
</tr>
<tr>
<td>50</td>
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<td>17.72</td>
<td>0.432</td>
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</tr>
<tr>
<td>70</td>
<td>18.05</td>
<td>18.28</td>
<td>-</td>
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</tr>
<tr>
<td>90</td>
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<td>18.45</td>
<td>0.416</td>
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<tr>
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<td>17.96</td>
<td>18.53</td>
<td>-</td>
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</tr>
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<td>-</td>
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<td>0.805</td>
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<td>-</td>
<td>0.724</td>
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<td>0.394</td>
<td>0.804</td>
</tr>
<tr>
<td>230</td>
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<td>18.58</td>
<td>-</td>
<td>0.742</td>
</tr>
</tbody>
</table>
TABLE A43

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

Particle size range = 5-10 μm
Channel length = 0.66m
Inclination angle to horizontal = 45°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate $(Q_o)_{expt}$ cc/min</th>
<th>Underflow Rate $(Q_u)_{expt}$ cc/min</th>
<th>Concentration of Solids in Feed, $c_0$ % V/V</th>
<th>Concentration of Solids in Underflow, $c_u$ % V/V</th>
</tr>
</thead>
<tbody>
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<td>0</td>
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<td>16.3</td>
<td>0.5</td>
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</tr>
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</tr>
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<td>16.4</td>
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<td>0.799</td>
</tr>
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<td>0.469</td>
<td>0.817</td>
</tr>
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TABLE A44

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

Particle size range = 5-10 μm
Channel length = 0.66m
Inclination angle to horizontal = 50°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

<table>
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<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate $(Q_o)_{expt}$ cc/min</th>
<th>Underflow Rate $(Q_u)_{expt}$ cc/min</th>
<th>Concentration of Solids in Feed, $c_o$ % v/v</th>
<th>Concentration of Solids in Underflow, $c_u$ % v/v</th>
</tr>
</thead>
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<td>0.769</td>
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TABLE A45

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

Particle size range = 5-10 μm
Channel length = 0.66 m
Inclination angle to horizontal = 55°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

<table>
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<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate (Q₀)expt cc/min</th>
<th>Underflow Rate (Qᵤ)expt cc/min</th>
<th>Concentration of Solids in Feed, cₓ % v/v₀</th>
<th>Concentration of Solids in Underflow, cᵤ % v/v</th>
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</thead>
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<td>11.69</td>
<td>-</td>
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<td>13.20</td>
<td>0.418</td>
<td>0.800</td>
</tr>
<tr>
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<td>-</td>
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<tr>
<td>Operating Time (mins)</td>
<td>Maximum Overflow Rate ((Q_o)_{expt}) cc/min</td>
<td>Underflow Rate ((Q_u)_{expt}) cc/min</td>
<td>Concentration of Solids in Feed, (c_o) % v/v</td>
<td>Concentration of Solids in Underflow, (c_u) % v/v</td>
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<td>25.3</td>
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<td>0.402</td>
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<td>25.3</td>
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<td>0.728</td>
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</table>
TABLE A47

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

Particle size range = 5-10 µm
Channel length = 0.95 m
Inclination angle to horizontal = 45°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_0)_{\text{expt}}) cc/min</th>
<th>Underflow Rate ((Q_u)_{\text{expt}}) cc/min</th>
<th>Concentration of Solids in Feed, (c_0) % v/v</th>
<th>Concentration of Solids in Underflow, (c_u) % v/v</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>20.0</td>
<td>20.0</td>
<td>0.5</td>
<td>0.0</td>
</tr>
<tr>
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<td>20.0</td>
<td>20.0</td>
<td>0.457</td>
<td>0.617</td>
</tr>
<tr>
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</tr>
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</tr>
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</tr>
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<td>20.0</td>
<td>0.396</td>
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</table>
### TABLE A48

**SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE**

- Particle size range = 5-10 μm
- Channel length = 0.95 m
- Inclination angle to horizontal = 50°
- Channel spacing = 3.4 cm
- Channel width = 4 cm
- Number of settling channels = 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{expt} ) cc/min</th>
<th>Underflow Rate ((Q_u)_{expt} ) cc/min</th>
<th>Concentration of Solids in Feed, (c_o ) % V/V</th>
<th>Concentration of Solids in Underflow, (c_u ) % V/V</th>
</tr>
</thead>
<tbody>
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<td>0.50</td>
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<td>0.846</td>
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<td>17.6</td>
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<td>0.837</td>
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</tr>
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<td>0.826</td>
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<td>0.804</td>
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</table>
TABLE A49

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

Particle size range = 5-10 μm
Channel length = 0.95 m
Inclination angle to horizontal = 55°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{expt}) cc/min</th>
<th>Underflow Rate ((Q_u)_{expt}) cc/min</th>
<th>Concentration of Solids in Feed, (c_0) % v/v</th>
<th>Concentration of Solids in Underflow, (c_u) % v/v</th>
</tr>
</thead>
<tbody>
<tr>
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<td>0.839</td>
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<td>0.443</td>
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TABLE A50
SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

<table>
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<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate (Q_o)expt cc/min</th>
<th>Underflow Rate (Q_u)expt cc/min</th>
<th>Concentration of Solids in Feed, C_o % v/v</th>
<th>Concentration of Solids in Underflow, C_u % v/v</th>
</tr>
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<tbody>
<tr>
<td>0</td>
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Particle size range = 5-10 μm
Channel length = 1.12m
Inclination angle to horizontal = 35°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1
TABLE A51

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

- Particle size range = 5-10 μm
- Channel length = 1.12 m
- Inclination angle to horizontal = 40°
- Channel spacing = 3.4 cm
- Channel width = 4 cm
- Number of settling channels = 1

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<th>Maximum Overflow Rate (Q₀) expt cc/min</th>
<th>Underflow Rate (Qᵢ) expt cc/min</th>
<th>Concentration of Solids in Feed, c₀ % V/V</th>
<th>Concentration of Solids in Underflow, cᵢ % V/V</th>
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<td>31.0</td>
<td>0.406</td>
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TABLE A52
SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

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<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ($Q_o^{expt}$) cc/min</th>
<th>Underflow Rate ($Q_u^{expt}$) cc/min</th>
<th>Concentration of Solids in Feed, $c_0$ % V/V</th>
<th>Concentration of Solids in Underflow, $c_u$ % V/V</th>
</tr>
</thead>
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Particle size range = 5-10 μm
Channel length = 1.12m
Inclination angle to horizontal = 45°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1
TABLE A53

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

- Particle size range = 5-10 µm
- Channel length = 1.12 m
- Inclination angle to horizontal = 50°
- Channel spacing = 3.4 cm
- Channel width = 4 cm
- Number of settling channels = 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{expt}) cc/min</th>
<th>Underflow Rate ((Q_u)_{expt}) cc/min</th>
<th>Concentration of Solids in Feed, (c_0) % v/v</th>
<th>Concentration of Solids in Underflow, (c_u) % v/v</th>
</tr>
</thead>
<tbody>
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<td>0</td>
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<td>0.452</td>
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</table>


TABLE A54

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

Particle size range = 5-10 \( \mu \)m
Channel length = 1.12 m
Inclination angle to horizontal = 55°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{expt} ) cc/min</th>
<th>Underflow Rate ((Q_u)_{expt} ) cc/min</th>
<th>Concentration of Solids in Feed, ( c_o ) % V/V</th>
<th>Concentration of Solids in Underflow, ( c_u ) % V/V</th>
</tr>
</thead>
<tbody>
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<tr>
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<td>86</td>
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<td>19.2</td>
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<td>19.2</td>
<td>0.416</td>
<td>0.872</td>
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TABLE A55
SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate $(Q_0)_{\text{expt}}$ cc/min</th>
<th>Underflow Rate $(Q_u)_{\text{expt}}$ cc/min</th>
<th>Concentration of Solids in Feed, $c_0$ % v/v</th>
<th>Concentration of Solids in Underflow, $c_u$ % v/v</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
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<td>28.85</td>
<td>-</td>
<td>0.831</td>
</tr>
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</table>

Particle size range = 5-10 μm
Channel length = 0.66 m
Inclination angle to horizontal = 40°
Channel spacing = 1.5 cm
Channel width = 4 cm
Number of settling channels = 2
TABLE A56

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

Particle size range = 5-10 μm
Channel length = 0.66 m
Inclination angle to horizontal = 45°
Channel spacing = 1.5 cm
Channel width = 4 cm
Number of settling channels = 2

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate $(Q_o)_{expt}$ cc/min</th>
<th>Underflow Rate $(Q_u)_{expt}$ cc/min</th>
<th>Concentration of Solids in Feed, $c_0$ % V/V</th>
<th>Concentration of Solids in Underflow, $c_u$ % V/V</th>
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</thead>
<tbody>
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<td>0.0</td>
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<td>23.70</td>
<td>23.78</td>
<td>-</td>
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</tr>
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<td>23.93</td>
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<td>-</td>
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<td>-</td>
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<td>24.04</td>
<td>0.420</td>
<td>0.812</td>
</tr>
<tr>
<td>230</td>
<td>23.75</td>
<td>24.20</td>
<td>-</td>
<td>0.945</td>
</tr>
</tbody>
</table>
TABLE A57
SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE

Particle size range = 5-10 \mu m
Channel length = 0.66 m
Inclination angle to horizontal = 50^\circ
Channel spacing = 1.5 cm
Channel width = 4 cm
Number of settling channels = 2

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate (Q_o)_{expt} cc/min</th>
<th>Underflow Rate (Q_u)_{expt} cc/min</th>
<th>Concentration of Solids in Feed, c_o % V/V</th>
<th>Concentration of Solids in Underflow, c_u % V/V</th>
</tr>
</thead>
<tbody>
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<td>20.36</td>
<td>20.39</td>
<td>-</td>
<td>0.848</td>
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<tr>
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<td>20.50</td>
<td>20.67</td>
<td>0.437</td>
<td>0.847</td>
</tr>
<tr>
<td>110</td>
<td>20.83</td>
<td>20.60</td>
<td>-</td>
<td>0.825</td>
</tr>
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<td>20.84</td>
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<td>-</td>
<td>0.832</td>
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<td>20.83</td>
<td>-</td>
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<td>20.75</td>
<td>-</td>
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<td>20.50</td>
<td>20.82</td>
<td>-</td>
<td>0.910</td>
</tr>
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</table>
**TABLE A58**

**SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COUNTERCURRENT MODE**

Particle size range  
Channel length  
Inclination angle to horizontal  
Channel spacing  
Channel width  
Number of settling channels

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ($Q_o$)expt cc/min</th>
<th>Underflow Rate ($Q_u$)expt cc/min</th>
<th>Concentration of Solids in Feed, $c_o$ % v/v</th>
<th>Concentration of Solids in Underflow, $c_u$ % v/v</th>
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</thead>
<tbody>
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<td>0.0</td>
</tr>
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<td>-</td>
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<td>-</td>
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</tr>
<tr>
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<td>15.70</td>
<td>0.424</td>
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<td>15.81</td>
<td>-</td>
<td>0.826</td>
</tr>
<tr>
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<td>15.77</td>
<td>15.82</td>
<td>0.419</td>
<td>0.784</td>
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<tr>
<td>190</td>
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<td>15.93</td>
<td>-</td>
<td>0.856</td>
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<td>15.57</td>
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<td>16.16</td>
<td>-</td>
<td>0.803</td>
</tr>
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- Particle size range = 5-10 μm
- Channel length = 0.66m
- Inclination angle to horizontal = 55°
- Channel spacing = 1.5 cm
- Channel width = 4 cm
- Number of settling channels = 2
TABLE A59

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COCURRENT SUBCRITICAL MODE

Particle size range = 5-10 μm
Channel length = 0.66 m
Inclination angle to horizontal = 35°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

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<th>Operating Time (mins)</th>
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<th>Underflow Rate $Q_u$ (cc/min)</th>
<th>Concentration of Solids in Feed, $c_0$ (% v/v)</th>
<th>Concentration of Solids in Underflow, $c_u$ (% v/v)</th>
</tr>
</thead>
<tbody>
<tr>
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<td>0</td>
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<td>22.33</td>
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<td>0.793</td>
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<td>-</td>
<td>0.798</td>
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TABLE A60

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COCURRENT SUBCRITICAL MODE

<table>
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<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{expt}) cc/min</th>
<th>Underflow Rate ((Q_u)_{expt}) cc/min</th>
<th>Concentration of Solids in Feed, (c_o) % v/v</th>
<th>Concentration of Solids in Underflow, (c_u) % v/v</th>
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</thead>
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<td>19.05</td>
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<td>19.33</td>
<td>-</td>
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<td>0.775</td>
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Particle size range = 5-10 μm
Channel length = 0.66m
Inclination angle to horizontal = 40°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1
**TABLE A61**

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COCURRENT SUBCRITICAL MODE

Particle size range = 5-10 μm  
Channel length = 0.66 m  
Inclination angle to horizontal = 45°  
Channel spacing = 3.4 cm  
Channel width = 4 cm  
Number of settling channels = 1

<table>
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<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{expt}) cc/min</th>
<th>Underflow Concentration of Solids in Feed, (C_o) % v/v</th>
<th>Concentration of Solids in Underflow, (C_u) % v/v</th>
</tr>
</thead>
<tbody>
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<td>17.73</td>
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<td>18.17</td>
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<td>-</td>
</tr>
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<td>0.417</td>
</tr>
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<td>18.20</td>
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</table>
TABLE A62

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COCURRENT SUBCRITICAL MODE

Particle size range = 5-10 μm
Channel length = 0.66 m
Inclination angle to horizontal = 50°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

<table>
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<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ( (Q_o)_{expt} ) cc/min</th>
<th>Underflow Rate ( (Q_u)_{expt} ) cc/min</th>
<th>Concentration of Solids in Feed, ( c_0 ) % v/v</th>
<th>Concentration of Solids in Underflow, ( c_u ) % v/v</th>
</tr>
</thead>
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</table>
TABLE A63

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COCURRENT SUBCRITICAL MODE

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{\text{expt}}) cc/min</th>
<th>Underflow Rate ((Q_u)_{\text{expt}}) cc/min</th>
<th>Concentration of Solids in Feed, (c_0) % v/v</th>
<th>Concentration of Solids in Underflow, (c_u) % v/v</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>12.3</td>
<td>12.66</td>
<td>0.5</td>
<td>0.0</td>
</tr>
<tr>
<td>50</td>
<td>12.54</td>
<td>13.41</td>
<td>0.425</td>
<td>0.447</td>
</tr>
<tr>
<td>70</td>
<td>12.72</td>
<td>12.56</td>
<td>-</td>
<td>0.578</td>
</tr>
<tr>
<td>90</td>
<td>12.95</td>
<td>12.49</td>
<td>0.374</td>
<td>0.843</td>
</tr>
<tr>
<td>110</td>
<td>12.99</td>
<td>12.53</td>
<td>-</td>
<td>0.798</td>
</tr>
<tr>
<td>130</td>
<td>13.20</td>
<td>12.75</td>
<td>0.368</td>
<td>0.740</td>
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<tr>
<td>150</td>
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<td>-</td>
<td>0.779</td>
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<tr>
<td>170</td>
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<td>12.62</td>
<td>0.357</td>
<td>0.705</td>
</tr>
<tr>
<td>190</td>
<td>13.16</td>
<td>12.55</td>
<td>-</td>
<td>0.761</td>
</tr>
<tr>
<td>210</td>
<td>13.17</td>
<td>12.43</td>
<td>0.366</td>
<td>0.770</td>
</tr>
<tr>
<td>230</td>
<td>13.29</td>
<td>12.46</td>
<td>-</td>
<td>0.734</td>
</tr>
</tbody>
</table>

Particle size range = 5-10 µm
Channel length = 0.66m
Inclination angle to horizontal = 55°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1
TABLE A64

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COCURRENT SUPERCritical MODE

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_\text{expt} \text{ cc/min})</th>
<th>Underflow Rate ((Q_u)_\text{expt} \text{ cc/min})</th>
<th>Concentration of Solids in Feed, (c_o % \text{ v/v})</th>
<th>Concentration of Solids in Underflow, (c_u % \text{ v/v})</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>24.52</td>
<td>24.23</td>
<td>0.5</td>
<td>0.0</td>
</tr>
<tr>
<td>30</td>
<td>24.64</td>
<td>24.17</td>
<td>0.459</td>
<td>1.050</td>
</tr>
<tr>
<td>50</td>
<td>24.86</td>
<td>24.46</td>
<td>-</td>
<td>0.914</td>
</tr>
<tr>
<td>70</td>
<td>24.98</td>
<td>23.39</td>
<td>0.432</td>
<td>0.879</td>
</tr>
<tr>
<td>100</td>
<td>24.83</td>
<td>24.20</td>
<td>-</td>
<td>0.857</td>
</tr>
<tr>
<td>120</td>
<td>24.97</td>
<td>24.67</td>
<td>0.438</td>
<td>0.630</td>
</tr>
<tr>
<td>140</td>
<td>25.50</td>
<td>25.40</td>
<td>-</td>
<td>0.841</td>
</tr>
<tr>
<td>160</td>
<td>24.84</td>
<td>24.91</td>
<td>0.449</td>
<td>0.917</td>
</tr>
<tr>
<td>180</td>
<td>25.03</td>
<td>25.00</td>
<td>-</td>
<td>0.878</td>
</tr>
<tr>
<td>200</td>
<td>25.08</td>
<td>24.92</td>
<td>0.422</td>
<td>0.850</td>
</tr>
<tr>
<td>220</td>
<td>24.84</td>
<td>23.72</td>
<td>-</td>
<td>0.810</td>
</tr>
</tbody>
</table>

Particle size range = 5-10 \(\mu\)m
Channel length = 0.66m
Inclination angle to horizontal = 35°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1
TABLE A65

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COCURRENT SUPERCritical MODE

Particle size range = 5-10 μm
Channel length = 0.66 m
Inclination angle to horizontal = 40°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{\text{expt}}) cc/min</th>
<th>Underflow Rate ((Q_u)_{\text{expt}}) cc/min</th>
<th>Concentration of Solids in Feed, (c_0) % v/v</th>
<th>Concentration of Solids in Underflow, (c_u) % v/v</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>21.7</td>
<td>21.73</td>
<td>0.5</td>
<td>0.0</td>
</tr>
<tr>
<td>70</td>
<td>22.48</td>
<td>22.10</td>
<td>0.408</td>
<td>0.847</td>
</tr>
<tr>
<td>92</td>
<td>22.49</td>
<td>22.33</td>
<td>0.427</td>
<td>0.804</td>
</tr>
<tr>
<td>113</td>
<td>21.48</td>
<td>22.19</td>
<td>0.417</td>
<td>0.834</td>
</tr>
<tr>
<td>171</td>
<td>21.67</td>
<td>21.98</td>
<td>0.392</td>
<td>0.755</td>
</tr>
<tr>
<td>191</td>
<td>21.84</td>
<td>21.68</td>
<td>0.400</td>
<td>0.797</td>
</tr>
<tr>
<td>211</td>
<td>21.64</td>
<td>22.13</td>
<td>0.394</td>
<td>0.726</td>
</tr>
<tr>
<td>235</td>
<td>21.86</td>
<td>21.62</td>
<td>0.390</td>
<td>0.776</td>
</tr>
<tr>
<td>258</td>
<td>21.99</td>
<td>22.32</td>
<td>0.387</td>
<td>0.748</td>
</tr>
<tr>
<td>278</td>
<td>22.00</td>
<td>22.34</td>
<td>0.388</td>
<td>0.734</td>
</tr>
<tr>
<td>302</td>
<td>21.71</td>
<td>22.14</td>
<td>0.390</td>
<td>0.773</td>
</tr>
<tr>
<td>327</td>
<td>21.74</td>
<td>22.60</td>
<td>0.371</td>
<td>0.709</td>
</tr>
<tr>
<td>365</td>
<td>21.39</td>
<td>23.06</td>
<td>0.384</td>
<td>0.725</td>
</tr>
</tbody>
</table>
TABLE A66

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COCURRENT SUPERCRITICAL MODE

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{expt}) cc/min</th>
<th>Underflow Rate ((Q_u)_{expt}) cc/min</th>
<th>Concentration of Solids in Feed, (c_0) % v/v</th>
<th>Concentration of Solids in Underflow, (c_u) % v/v</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>18.09</td>
<td>18.18</td>
<td>0.5</td>
<td>0.0</td>
</tr>
<tr>
<td>60</td>
<td>18.29</td>
<td>18.54</td>
<td>0.411</td>
<td>0.806</td>
</tr>
<tr>
<td>80</td>
<td>18.03</td>
<td>18.54</td>
<td>0.425</td>
<td>0.917</td>
</tr>
<tr>
<td>100</td>
<td>18.33</td>
<td>19.17</td>
<td>0.402</td>
<td>0.844</td>
</tr>
<tr>
<td>120</td>
<td>18.31</td>
<td>18.13</td>
<td>0.401</td>
<td>0.862</td>
</tr>
<tr>
<td>160</td>
<td>18.58</td>
<td>17.94</td>
<td>0.413</td>
<td>0.824</td>
</tr>
<tr>
<td>210</td>
<td>18.46</td>
<td>18.10</td>
<td>0.393</td>
<td>0.875</td>
</tr>
<tr>
<td>235</td>
<td>18.30</td>
<td>17.89</td>
<td>0.421</td>
<td>0.862</td>
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<tr>
<td>255</td>
<td>18.58</td>
<td>18.49</td>
<td>0.414</td>
<td>0.854</td>
</tr>
<tr>
<td>275</td>
<td>18.35</td>
<td>18.34</td>
<td>0.413</td>
<td>0.843</td>
</tr>
<tr>
<td>295</td>
<td>18.45</td>
<td>18.61</td>
<td>0.402</td>
<td>0.804</td>
</tr>
<tr>
<td>340</td>
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<td>18.65</td>
<td>0.473</td>
<td>1.054</td>
</tr>
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</table>

Particle size range = 5-10 µm
Channel length = 0.66m
Inclination angle to horizontal = 45°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1
<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_o)_{expt}) cc/min</th>
<th>Underflow Rate ((Q_u)_{expt}) cc/min</th>
<th>Concentration of Solids in Feed, (c_o) % V/V</th>
<th>Concentration of Solids in Underflow, (c_u) % V/V</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>16.01</td>
<td>15.84</td>
<td>0.5</td>
<td>0.0</td>
</tr>
<tr>
<td>60</td>
<td>16.06</td>
<td>15.79</td>
<td>0.411</td>
<td>0.633</td>
</tr>
<tr>
<td>80</td>
<td>16.25</td>
<td>15.87</td>
<td>0.405</td>
<td>0.813</td>
</tr>
<tr>
<td>100</td>
<td>16.08</td>
<td>15.65</td>
<td>0.390</td>
<td>0.886</td>
</tr>
<tr>
<td>120</td>
<td>16.21</td>
<td>15.67</td>
<td>0.397</td>
<td>0.806</td>
</tr>
<tr>
<td>140</td>
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<td>15.83</td>
<td>0.396</td>
<td>0.782</td>
</tr>
<tr>
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<td>16.03</td>
<td>16.09</td>
<td>0.403</td>
<td>0.802</td>
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<td>15.97</td>
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<td>0.805</td>
</tr>
<tr>
<td>200</td>
<td>16.29</td>
<td>16.15</td>
<td>0.403</td>
<td>0.794</td>
</tr>
<tr>
<td>220</td>
<td>15.99</td>
<td>15.94</td>
<td>0.398</td>
<td>0.775</td>
</tr>
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<td>16.06</td>
<td>15.64</td>
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<td>0.758</td>
</tr>
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</table>
TABLE A68

SOLIDS CONCENTRATION IN FEED AND UNDERFLOW AS A FUNCTION OF OPERATING TIME IN THE COCURRENT SUPERCritical MODE

Particle size range = 5-10 μm
Channel length = 0.66 m
Inclination angle to horizontal = 55°
Channel spacing = 3.4 cm
Channel width = 4 cm
Number of settling channels = 1

<table>
<thead>
<tr>
<th>Operating Time (mins)</th>
<th>Maximum Overflow Rate ((Q_0)_{expt}) cc/min</th>
<th>Underflow Rate ((Q_U)_{expt}) cc/min</th>
<th>Concentration of Solids in Feed, (c_0) % v/v</th>
<th>Concentration of Solids in Underflow, (c_{u1}) % v/v</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>12.05</td>
<td>12.01</td>
<td>0.50</td>
<td>0.0</td>
</tr>
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<td>12.19</td>
<td>12.44</td>
<td>0.450</td>
<td>0.809</td>
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<tr>
<td>70</td>
<td>12.16</td>
<td>12.31</td>
<td>-</td>
<td>0.922</td>
</tr>
<tr>
<td>90</td>
<td>12.32</td>
<td>12.60</td>
<td>0.417</td>
<td>0.840</td>
</tr>
<tr>
<td>110</td>
<td>12.39</td>
<td>12.24</td>
<td>-</td>
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<td>12.23</td>
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<td>12.34</td>
<td>12.54</td>
<td>-</td>
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<td>12.29</td>
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<td>190</td>
<td>12.57</td>
<td>12.66</td>
<td>-</td>
<td>0.891</td>
</tr>
<tr>
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<td>12.52</td>
<td>12.39</td>
<td>0.420</td>
<td>0.889</td>
</tr>
<tr>
<td>230</td>
<td>12.71</td>
<td>12.51</td>
<td>-</td>
<td>0.821</td>
</tr>
</tbody>
</table>
## NOMENCLATURE

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
<th>Dimensions</th>
</tr>
</thead>
<tbody>
<tr>
<td>a</td>
<td>Acceleration of solids down lower inclined slope</td>
<td>LT²</td>
</tr>
<tr>
<td>A</td>
<td>Floor area occupied by the lamella plates</td>
<td>L²</td>
</tr>
<tr>
<td>b</td>
<td>Channel spacing (plate spacing)</td>
<td>L</td>
</tr>
<tr>
<td>B</td>
<td>Dimensionless channel spacing</td>
<td>L</td>
</tr>
<tr>
<td>c₀</td>
<td>Concentration (volume fraction of particles) in feed to separator</td>
<td>-</td>
</tr>
<tr>
<td>cᵤ</td>
<td>Concentration of particles in underflow from separator</td>
<td>-</td>
</tr>
<tr>
<td>C₀, C₁, C₂, C₃, C₄</td>
<td>Constants of integration</td>
<td>-</td>
</tr>
<tr>
<td>D</td>
<td>Hydraulic diameter of settling channels</td>
<td>L</td>
</tr>
<tr>
<td>F</td>
<td>Empirical coefficient in Graham and Lama's and Ghosh's equations</td>
<td>-</td>
</tr>
<tr>
<td>Fₔ</td>
<td>Drag force per unit area</td>
<td>ML⁻¹T⁻²</td>
</tr>
<tr>
<td>F₉</td>
<td>Frictional force per unit area</td>
<td>ML⁻¹T⁻²</td>
</tr>
<tr>
<td>F₉</td>
<td>Gravitational force per unit area</td>
<td>ML⁻¹T⁻²</td>
</tr>
<tr>
<td>g</td>
<td>Gravitational constant</td>
<td>LT⁻²</td>
</tr>
<tr>
<td>h</td>
<td>Vertical height of suspension measured from the base of the upper inclined surface at time t (or characteristic length of macroscale motion)</td>
<td>L</td>
</tr>
<tr>
<td>H</td>
<td>Width of clear and suspension layers</td>
<td>L</td>
</tr>
<tr>
<td>j</td>
<td>Particle species</td>
<td>-</td>
</tr>
<tr>
<td>L</td>
<td>Length of lamella plate</td>
<td>L</td>
</tr>
<tr>
<td>m</td>
<td>Effective mass of sludge layer per unit area</td>
<td>ML⁻²</td>
</tr>
<tr>
<td>n</td>
<td>Number of settling channels</td>
<td>-</td>
</tr>
<tr>
<td>N</td>
<td>Number of species of particles/Number of settler plates</td>
<td>-</td>
</tr>
<tr>
<td>Symbol</td>
<td>Description</td>
<td>Dimensions</td>
</tr>
<tr>
<td>-------------</td>
<td>------------------------------------------------------------------------------</td>
<td>------------------</td>
</tr>
<tr>
<td>NGr</td>
<td>Sedimentation Grashof Number</td>
<td>-</td>
</tr>
<tr>
<td>NRe (or Re)</td>
<td>Sedimentation Reynolds Number</td>
<td>-</td>
</tr>
<tr>
<td>R</td>
<td>Liquid Reynolds Number</td>
<td>-</td>
</tr>
<tr>
<td>P</td>
<td>Pressure</td>
<td>ML(^{-1})T(^{-2})</td>
</tr>
<tr>
<td>Q</td>
<td>Volumetric flowrate through the separator</td>
<td>L(^3)T(^{-1})</td>
</tr>
<tr>
<td>Q(_f)</td>
<td>Volumetric feedrate to separator</td>
<td>L(^3)T(^{-1})</td>
</tr>
<tr>
<td>Q(_o)</td>
<td>Volumetric overflow rate from separator</td>
<td>L(^3)T(^{-1})</td>
</tr>
<tr>
<td>Q(_u)</td>
<td>Volumetric underflow rate from separator</td>
<td>L(^3)T(^{-1})</td>
</tr>
<tr>
<td>Q(_l)</td>
<td>Volumetric overflow per unit breadth</td>
<td>L(^2)T(^{-1})</td>
</tr>
<tr>
<td>S(j)(t)</td>
<td>Dimensionless instantaneous volumetric rate at which suspension is devoid of</td>
<td>-</td>
</tr>
<tr>
<td></td>
<td>species j</td>
<td></td>
</tr>
<tr>
<td>t</td>
<td>Time</td>
<td>T</td>
</tr>
<tr>
<td>t(_R)</td>
<td>Particle residence time in settling channel</td>
<td>T</td>
</tr>
<tr>
<td>T</td>
<td>Thickness of sludge layer</td>
<td>L</td>
</tr>
<tr>
<td>u</td>
<td>Longitudinal component of velocity in any layer</td>
<td>LT(^{-1})</td>
</tr>
<tr>
<td>v</td>
<td>Vertical settling velocity of particles in suspension</td>
<td>LT(^{-1})</td>
</tr>
<tr>
<td>V</td>
<td>Velocity of fluid flow</td>
<td>LT(^{-1})</td>
</tr>
<tr>
<td>W</td>
<td>Width of lamella plate</td>
<td>L</td>
</tr>
<tr>
<td>x</td>
<td>Coordinate in direction parallel to settler length</td>
<td>L</td>
</tr>
<tr>
<td>X</td>
<td>Dimensionless distance along the upper inclined surface measured from its base</td>
<td>-</td>
</tr>
<tr>
<td>y</td>
<td>Coordinate in direction perpendicular to settler length</td>
<td>L</td>
</tr>
<tr>
<td>(\alpha)</td>
<td>Angle of inclination (from horizontal)</td>
<td>degrees</td>
</tr>
<tr>
<td>(\delta)</td>
<td>Dimensionless clear liquid layer thickness</td>
<td>-</td>
</tr>
<tr>
<td>(\varepsilon)</td>
<td>Volume concentration of solids in the sludge layer</td>
<td>-</td>
</tr>
<tr>
<td>Symbol</td>
<td>Description</td>
<td>Dimensions</td>
</tr>
<tr>
<td>--------</td>
<td>-----------------------------------------------------------------------------</td>
<td>---------------------</td>
</tr>
<tr>
<td>$\eta$</td>
<td>Thickness of clear liquid layer</td>
<td>L</td>
</tr>
<tr>
<td>$\theta$</td>
<td>Angle of inclination (from the vertical)</td>
<td>degrees</td>
</tr>
<tr>
<td>$\Lambda$</td>
<td>Ratio of Sedimentation Grashof Number to the Sedimentation Reynolds Number</td>
<td>-</td>
</tr>
<tr>
<td>$\mu$</td>
<td>Viscosity of pure fluid</td>
<td>ML$^{-1}$T$^{-1}$</td>
</tr>
<tr>
<td>$\mu_F$</td>
<td>Coefficient of frictional resistance</td>
<td>-</td>
</tr>
<tr>
<td>$\pi$</td>
<td>Pi constant</td>
<td>-</td>
</tr>
<tr>
<td>$\rho$</td>
<td>Density of any layer in separator</td>
<td>ML$^{-3}$</td>
</tr>
<tr>
<td>$\rho_F$</td>
<td>Density of pure fluid</td>
<td>ML$^{-3}$</td>
</tr>
<tr>
<td>$\rho_p$</td>
<td>Density of particles</td>
<td>ML$^{-3}$</td>
</tr>
<tr>
<td>$\Delta$</td>
<td>Thickness of suspension layer</td>
<td>L</td>
</tr>
<tr>
<td>$\phi$</td>
<td>Angle to horizontal at which sludge flow down an inclined slope first occurs</td>
<td>degrees</td>
</tr>
</tbody>
</table>

**Superscript**

"$\sim$" denotes stretch variables

**Subscripts**

1. Clear liquid layer
2. Suspension layer
3. Sludge layer
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ERRATUM

1. Page 1, para. 2. For 'Clavenger' read 'Clevenger'
2. Page 10, para. 2. For 'greater' read 'lower'
3. Page 19, para. 1. For 'Clavenger' read 'Clevenger'
4. Page 24, para. 3. For 'Clavenger' read 'Clevenger'
5. Page 46, para. 2. For 'overflow' read 'underflow'
   For 'underflow' read 'overflow'
   para. 3. For '/ft^3' read 'ft^3/ft^2'
6. Page 107, para. 1. For '16" read '13''
7. Page 128, para. 1. For 'Herbolzheimer' read 'Herbolzheimer'
   For 'surpressed' read 'suppressed'
   For '(0(10^4) - 0(10^3))' read '(0(10^4) - 0(10^7))'
   For '(0(1) - 0(10))' read '(0(1) - 0(10))'
8. Page 129, para. 3. For 'Namakura' read 'Nakamura'
9. Page 165, para. 3. After '6.31' add 'and see page 67'
10. Page 183, para. 2. For '1.12' read '0.95'
11. Page 237, para. 2. For 'Herbolzheimer' read 'Herbolzheimer'
13. Page 98, For 'Figure 5.1: Layer Movement' read 'Figure 5.4: Fracture Flow'
   For 'Figure 5.2: Heap Movement' read 'Figure 5.3: Bulk Movement'
   For 'Figure 5.3: Bulk Movement' read 'Figure 5.2: Heap Movement'
   For 'Figure 5.4: Fracture Flow' read 'Figure 5.1: Layer Movement'