The Development of an Exxpress Unit for the Dewatering of Waterworks Sludges and the Production of Potable Water

MP Pryor • DJ Mullan

Report to the Water Research Commission
by
Umgeni Water

WRC Report No 568/1/98

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ISBN 1 86845 385 5

Executive Summary

The Tubular Filter Press Unit (TFP) at the Umgeni Water H.D. Hill Waterworks has been in operation since 1987. During an evaluation of the unit for the application of dewatering waterworks sludges, it was decided that there were significant weaknesses in the design, causing tube blockages and other operating problems, which need to be rectified in order to produce a reliable process. It was concluded that an important factor in the successful operation of the Tubular Filter Press, was the availability of a full-scale or pilot plant for experimental and development purposes.

In addition to its application for the dewatering of waterworks sludge, the tubular filtration process can provide potable water quality water (without the addition of chemicals) at high fluxes and low energy consumption. It was on this basis that a submission was made to the Water Research Commission (WRC) for a project to pursue further experimental work at the Umgeni Water Process Evaluation Facility. Following the resolution of licensing agreements between the WRC and Hi-Tech Water the project was initiated in 1994.

OBJECTIVES

The Aims of the Project were to :

- Develop and demonstrate the Exxpress Process for the dewatering of water works sludges.
- Apply the Exxpress Process to the production of potable water from river water.

In order to achieve the objectives and improve the design of the Tubular Filter Press, a number of specific areas were identified that required attention. These included :

1. Improving the design of the Tubular Filter Press to reduce the occurrence of cloth splits and tube blockages, and addressing aspects of the manifold layout to improve the flow distribution and cleaning.
2. Developing a commercially operational unit for waterworks sludges and providing a detailed design of the new plant.
3. Monitoring the capital and operating costs and comparing these to similar processes in industry.
4. Assessing the performance of the new design by extended operation during the project.
5. Assessing the use of the new design for potable water production.

6. Developing techniques for determining design parameters for the Tubular Filter Press by extending the filtration model proposed by Dr. Rencken and incorporating these into a design procedure.

DESIGN OF A VERTICAL TUBULAR FILTER PRESS

A design sub-committee was convened to address specific aspects of the design, and it was proposed that a vertically mounted tube, shorter in length and of a larger tube diameter be used. A single tube pilot plant was constructed at the Umgeni Water Process Evaluation Facility to demonstrate the proposed configuration. Following successful small-scale vertical tube trials, detailed mechanical drawings were approved for the manufacture and assembly of a full-scale demonstration unit at the Umgeni Water Wiggins Waterworks.

Construction of the unit was completed during September 1995 and was successfully demonstrated to delegates attending the International Water Supply Association (IWSA) Conference in Durban. As the supply of curtain fabric material had to be negotiated with Gelvenor Textiles, the plant was initially operated using single tubes of another material. This material eventually split and when replaced with curtains manufactured by Gelvanor Textiles, no further tube splitting was experienced. The operation of the Vertical Tubular Filter Press is detailed by the following stages.

Filtration Cycle - Feed sludge is pumped into the tubes under pressure. The formation of a filter cake occurs rapidly on the inside of the tubes, and the permeate is directed to a collection tank. The highly resistive nature of the filter cake results in a reduction in filtration rate (flux) until an operating limit is reached.

Tube Discharge - Once a final permeate flux has been achieved the tube discharge is initiated. The discharge valve at the bottom of the tubes is opened and the tubes are emptied onto a conveyor belt. In this way, dilute sludge is returned to the feed tank and filter cake is separated for waste disposal.

Flush Cycle - Sludge from the feed tank is then pumped at a high flowrate through the tubes and any remaining cake is washed onto the conveyor belt. The efficiency of the removal is dependent on the amount of solids deposited in the tubes and the nature of the dewatered solids. Under certain operating conditions the flush cycle may not totally clean the inside of the tubes which will result in a decrease in the performance of the unit with time.

Roller Action - A double roller is provided which squeezes the double row of tubes together. creating a restriction through which the flushing fluid is pumped. The increase in velocity through the restriction ensures a reliable and consistent cleaning of the tubes. It has been found that under certain operating conditions the rollers may not be necessary. Once complete, the operating cycle is repeated.

EXPERIMENTAL OPERATION TO DATE

Laboratory-scale compression-permeability cell (C-P cell) tests were performed to evaluate these tests as an effective means of obtaining sludge characteristics. The experimental method was investigated and found to be reliable, but some difficulties were experienced. A new C-P cell was manufactured to measure the transmitted pressure, to determine whether the effect of wall friction is significant when compressing highly resistive waterworks sludge.

The operation of a single tube pilot plant demonstrated the use of the vertical tubular technology for the dewatering of waterworks sludges. The effects of varying the operating parameters of feed solids concentration, operating pressure and final permeate flux were examined.

Once the new large-scale plant had been installed a number of practical problems were experienced, including losses of sludge from the conveyor during flushing, unstable roller movement, power tripping during flushing, inefficient mixing in the feed tank and the collection of solids from the conveyor. Other safety related issues have also been addressed by the inclusion of a high pressure switch on the feed pump and the installation of a splash cover around the material tubes.

The Vertical Tubular Filter Press was operated intermittently in a batch mode of operation where the recovery of the solids and dewatering performance were monitored. During the pilot-plant operation, 150 m^3 of sludge was filtered and the feed solids concentration varied between 7 and 30 g/l (0,7 and 3,0 % m/v). By varying the operating parameters of operating pressure and final flux before cleaning, and monitoring the feed solids concentration some interesting trends were observed.

The filtration model proposed by Rencken was extended to include a new theory of *Area Contact* between particles in the filter cake. A method was also developed to use pilot-scale or full-scale operating data to obtain parameters to describe the sludge characteristics. These parameters would normally be obtained from laboratory tests. The solution methodologies that have been developed are being incorporated into user-friendly Windows based software. This work is still to be completed.

SUMMARY OF RESULTS

The deficiencies in the operation of the Tubular Filter Press at the Umgeni Water H.D. Hill Waterworks were identified and modifications to the design were proposed. A single-tube pilot plant was constructed and successfully operated in a new vertical configuration, demonstrating that the proposed design was feasible. The single tube pilot plant was found to accurately represent the filtration and cake removal, and can be used effectively to obtain design information for the sizing of dewatering applications.

A Vertical Tubular Filter Press was manufactured and constructed at the Umgeni Water Wiggins Waterworks in Durban, and was demonstrated to delegates of the IWSA conference during September 1995.

The performance of the filter has been reasonable producing cake concentrations between 20 and 32 percent solids (m/m), at cake recoveries up to 75 %. Although a trend of increasing cake solids with increasing pressure and increasing feed concentration was observed, the nature of the sludge from Wiggins was variable during the period of operation. The suspended solids in the raw water, the amount of bentonite, lime, hypochlorite and coagulant addition all have an affect on the nature of the sludge which will impact on the performance of the unit for sludge dewatering.

The cake recovery during cleaning was found to be dependent on the filtration time (or the final filtration rate before cleaning) as a sludge which has been allowed to reach compression equilibrium, will be more resistant to the vigorous effects of flushing. A cake which has not been compressed sufficiently will reslurry more easily and will pass through the conveyor belt thereby lowering the cake recovery, or more significantly the solids will remain on the inside of the filter tubes resulting in a shorter filtration time during subsequent plant operation, and hence poorer plant performance. The solids dewatering rate was found to depend on the filtration pressure, feed solids concentration and final permeate flux allowed before flushing.

The plant operation was found to be highly dependent on the sludge characteristics, not only with regard to the filtration cycle (cake formation) but also the flush cycle (cake removal). Cake recovery is a complex function of operating pressure, final flux and feed solids concentration. In order to optimally operate a tubular filter not only should the filtration characteristics be determined, but also the recovery *characteristic function*. It was found that under certain operating conditions, flushing without the use of a roller may be sufficient to effectively remove the cake from the tubes. This may not always be the case as the sludge characteristics were found to vary considerably during the operation of the plant.

Tube blockages (previously experienced at the H.D. Hill Waterworks) were completely eliminated by increasing the tube diameter to 60 mm and decreasing the tube length. The increased tube diameter did not result in any occurrences of tube splitting or failure using the fabric produced by Gelvanor. Cake release and conveyance out of the tubes improved as the vertical orientation of the tubes assisted this by collapsing during the flush cycle.

The addition of lime to a waterworks sludge was found to improve the filterability of the sludge by altering the sludge characteristics. This was determined by C-P Cell tests, and evident during the continuous operation of the Vertical Tubular Filter Press. The addition of lime can have a negative impact on the plant operation as the pH of sludge and permeate increases significantly, and fouling of the woven tube fabric may occur due to the precipitation of calcium carbonate. Wall friction in the C-P Cell tests was investigated to determine if this was significant for waterworks sludges, and whether the standard C-P cell test was accurate in determining sludge characteristics. Although wall friction was observed the difference in the sludge characteristics obtained did not appear to be significant.

A new generalised *Area Contact Model* has been proposed for the constant pressure compressible cake filtration. Solution methodologies have been developed to regress for cake characteristics from operating plant data and to account for the period of pressurisation at the start of the filtration cycle. Once the software has been totally developed the accuracy of the *Area Contact Model* can be determined.

A pilot plant (comprising a single vertical tube) was operated in dead-end filtration mode to assess the use of the new design in raw water filtration for the production of potable water. It was shown that a precoat of limestone is required to reduce the turbidity of the raw water to below 1 NTU, but the efficiency of pre-coating in dead end mode (in a vertical tube) was poor. This process was clearly not adequate when compared to the Crossflow Microfiltration process for potable water production.

In meeting the objectives of the project a Vertical Tubular Filter Press was designed and developed for the dewatering of waterworks sludges. The limitations of the previous design, especially tube blockages were totally eliminated and provided the process can operate without the use of rollers for sludge removal and tube cleaning a reliable and inexpensive process has been developed.

The use of the new *vertical configuration* operating in dead-end filtration mode is not recommended for potable water production as the effectiveness of pre-coating using limestone is poor, and the turbidity of the final water occasionally exceeds the SABS guidelines for potable water (<1 NTU).

RECOMMENDATIONS

It is recommended that this technology be actively marketed for dewatering of sludges where few problems are likely to occur. Areas to be avoided are organic effluents and highly variable industrial effluents where problems may arise as a result of inadequate information. To achieve this purchasing agreements need to be set up with the curtain supplier, and a small business could be initiated for the manufacture of curtains.

The Vertical Tubular Filter Press at the Wiggins Waterworks should be moved to a site where it can be optimised and then operated on a continuous basis for an extended period of time in order to fully demonstrate and market the technology. The application should be specifically chosen where the sludge supply and characteristics are more consistent (more particulate in nature). Sufficient instrumentation should be installed to adequately monitor the continuous operation.

A limitation of the single tube pilot plant is that it can only be operated in a batch mode. It is also recommended that this be automated such that it can be utilised for process investigations of new applications and together with the computer model form a total design package.

Further investigations into the aspects of cake removal are recommended with the objective to develop a mathematical recovery function based on the parameters that have been identified.

It is recommended that the software be completed and fully evaluated before being developed into a marketable product that will compliment the technology.

The single-tube pilot plant should be maintained by Umgeni Water, and hired to the licensed companies for pilot-plant studies to promote the use of the vertical tubular technology. The hiring charges should be sufficient to adequately maintain and repair, as well as improve or modify the operation of the pilot plant.

Acknowledgments

The research in this report emanates from a project funded jointly by the Water Research Commission and Umgeni Water and entitled :

The development of an Exxpress Unit for the dewatering of waterworks sludges and the production of potable water.

The Steering Committee responsible for this project consisted of the following persons :

Dr T Erasmus	Water Research Commission (Chairman)
Mr P Smit / W van Greuning	Water Research Commission (Secretary)
Dr G Offringa	Water Research Commission
Prof C Buckley	University of Natal, Durban
Mr K Treffry Goatley	Explochem Water Treatment (Pty) Ltd.
Mr D Kerdachi / D. Nozaic	Umgeni Water
Mr R B Townsend	Explochem Water Treatment (Pty) Ltd.
Dr M Starzak	University of Natal, Durban
Dr G Rencken	U.S Filter

The financing of the project by the Water Research Commission, the provision of laboratory, pilot-plant and office facilities by Umgeni Water and the contribution of the members of the Steering Committee is gratefully acknowledged.

In addition the authors would like to express their gratitude to the following people for their co-operation and assistance with the project.

K Treffry Goatley and P Schwatrz for their continued assistance throughout the project, particularly with regard to the commissioning and trouble shooting during the initial operation of the pilot plants.

The University of Natal (Durban), Chemical Engineering Department, and the staff in the department for the use of the workshop facilities and manufacturing the Compression-Permeability Cells.

The Umgeni Water Wiggins Waterworks for allowing the construction and operation of the Vertical Tubular Filter Press at the sludge plant, and for ensuring availability of sludge when operating the unit continuously.

Mr R Rajagopaul (Chief Technician), Mr T Nyawose (Contract Technician) for their dedication and assistance during the course of the project.

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Nomenclature

A	=	area of plane perpendicular to the direction of filtrate flow, (m^2)
A_c	=	interparticle contact area in the plane perpendicular to the direction of filtrate flow, (m^2)
A_{cell}	=	area of cake in C-P cell, (m^2)
A_0	=	coefficient of area contact, (-)
a	=	empirical constant, (-)
B	=	empirical constant, (-)
b	=	empirical constant, (-)
c	=	cohesive force between the side wall and the compressed cake, (Pa)
D	=	inside diameter of the cell, (m)
d_f	=	mean diameter of particle aggregates, (m)
F	=	empirical constant, (-)
F_s	=	accumulated frictional drag force on particles, (N)
f	=	coefficient of friction, (-)
$f_A(...)$	=	area contact function, (-)
G	=	the number of experimental observations, (-)
g	=	constant of gravitational acceleration, (m/s^2)
H_∞	=	final height of sediment, (m)
j	=	integer number, (-)
K	=	permeability, (m^2)
K_i	=	permeability of cake when $p_s \le p_{si}$, (m^2)
K_0	=	permeability at $p_s = 0$, (m^2)
k_0	=	the coefficient of earth pressure at rest, (-)
L	=	compressed equilibrium thickness of the cake, (m)
l	=	tube length or axial length of cake, (m)
$M \ge 1$	=	the number of permeability correlation parameter sets, (-)
m	=	mass fraction of moisture in cake, (-)
N	=	number of particles in the plane perpendicular to the direction of filtrate flow, (-)
n	=	integer number, (-)
P_0	=	operating pressure, (Pa)
ΔP_c	=	cake pressure drop, (Pa)
ΔP_m	=	medium pressure drop, (Pa)
p_A	=	the applied pressure at the top of the cake, (Pa)
p_L	=	liquid pressure, (Pa)
p_s	=	solids compressive pressure, (Pa)
p_{sa}	=	packing orientation yield compressive pressure, (Pa)
p_{si}	=	solids compressive pressure below which the permeability and porosity are assumed constant, (Pa)
p_T	=	transmitted pressure through the cake, (Pa)
p_V	=	vertical solids pressure in the cake, (Pa)
p_a	=	empirical constants, (-)
Δp_c	=	hydrostatic pressure drop across cake in C-P cell, (Pa)
Q_f	=	volumetric flow rate of filtrate, (m^3/s)
R_m	=	medium resistance, (m-1)

r	=	radius, (m)
r_1	=	internal tube radius, (m)
r_2	=	internal radius of the cake, (m)
$T \geq 1$	=	the number of porosity correlation parameter sets, (-)
t	=	filtration time, (s)
Δt_c	=	thickness of cake in C-P cell, (m)
u_l	=	apparent liquid velocity relative to solids, (m/s)
V_f	=	volume of filtrate, (m^3)
V_0	=	initial settling velocity of surface of sediment, (m/s)
v	=	volume filtrate per unit medium area, (m^3/m^2)
W_1	=	cake pressure drop weighting factor, (-)
W_2	=	average porosity weighting factor, (-)
x	=	distance from medium in planar filtration, (m)
y	=	distance measured from bottom of cylinder, (m)
z	=	distance from the top of the cake, (m)

Greek Symbols

α	=	objective function result, (-)
β	=	empirical constant, (-)
δ	=	empirical constant, (-)
γ	=	empirical constant, (-)
ε	=	porosity, (-)
ε_{av}	=	average porosity of the cake
ε_i	=	porosity of cake when $p_s \leq p_{si}$, (-)
ε_{if}	=	internal porosity of particle aggregates, (-)
ε_{in}	=	initial porosity of suspension
ε_0	=	porosity at $p_s = 0$, (-)
θ	=	cylindrical co-ordinate, (radians)
μ_f	=	liquid viscosity, (Pa.s)
λ	=	empirical constant, (-)
ρ_l	=	liquid density, (kg/m^3)
ρ_s	=	solids density, (kg/m^3)
σ	=	root-mean-squared deviation, (-)
ϕ	=	angle of shearing resistance, (radians)
ϕ_s	=	volume fraction solids in feed sludge, (-)
ω	=	volume of dry solids per unit cross-sectional area, measured from bottom of cylinder, (m^3/m^2)
ω_c	=	volume of cake dry solids per unit medium area, (m^3/m^2)
ω_0	=	total volume of dry solids per unit cross sectional area, (m^2)
$\Pi_{density}$	=	local particulate density, (m^{-3})
$\Pi_{orientation}$	=	local particle orientation, (-)
Ψ_{shape}	=	local particle shape distribution, (-)
Ψ_{size}	=	local particle size distribution, (-)

1 Introduction

The Tubular Filter Press Unit at the Umgeni Water H.D. Hill Waterworks in Pietermaritzburg has been in operation since 1987. Although the unit operates regularly, the process has not been developed commercially because of tube blockages and other operational problems experienced on the plant. One of the problems encountered was that the research team was unable to effectively optimise the design and process operation, as the plant was under the control of an operating waterworks and was required on a routine basis. It was concluded that an important factor in the successful operation of the Tubular Filter Press Plant, was the availability of a full-scale or pilot plant for experimental and developmental purposes.

The Water Research Commission holds the patent for the Tubular Filter Press which has been given the trade name *Exxpress* by the licensees, Hi-Tech Water (HTW). The joint venture company Exxpress Technologies, which was formed between HTW and Yorkshire Water Plc. has failed. One of the problems was that there was insufficient technology transfer between Umgeni Water, Explochem, the University of Natal and HTW on the one hand and Exxpress Technologies on the other. The basic equipment was not at fault, rather the drive or will to succeed was not present. In order to get the technology into the market place it is necessary for the champions of the technology (Umgeni Water, Explochem and the University of Natal) to become directly involved in the development and demonstration of the technology.

One of the difficulties in providing assistance to the licensees is that the Tubular Filter Press at the Umgeni Water HD Hill Waterworks is operated as a production unit and is not available for tests. It is important for the successful operation of plants that relevant plant-scale advice can be given. The publication of Rencken (1992) has yielded a large body of knowledge on the mechanisms that determine cake formation, cake recovery and tube blockages. Procedures have been developed to aid the design and operation of the process. This work needs to be applied to a full-scale plant.

In addition to its application for the dewatering of waterworks sludge the Exxpress Process can provide potable quality water (without the addition of chemicals) at high fluxes and at low energy consumption. Results obtained from the Tubular Filter Press at the H.D. Hill Waterworks indicated that the water fluxes and the permeate quality achieved were comparable to the results obtained from some of the poorly performing Crossflow Microfiltration Plants. These results have been confirmed during the course of WRC project No 238, *Research on the Design Criteria for*

Cross-flow Microfiltration. On a synthetic river water feed with a turbidity of 73 NTU fluxes of 40 l/m^2h were obtained over periods of 36 h at applied pressures of 200 kPa. Dead-end filtration was used. The turbidity of the permeate was less than 0,3 NTU which is below the World Health Organisation guidelines for potable water. These results indicate that the Exxpress equipment can be used to produce drinking water at very low energy consumption. A natural water head of 10 to 40 m would be suitable to produce permeate. The roller desludging technique could be used to clean the tubes, and the dead-end operation with cake production indicated that very high water recoveries are possible.

During an evaluation of the previous projects, it was decided that there were significant weaknesses in the design of the Tubular Filter Press at H.D. Hill Waterworks which needed to be rectified in order to produce a marketable product and obtain effective technology transfer. It was on this basis that a submission was made to the Water Research Commission for a project to pursue further experimental work under the supervision of the researchers at the Umgeni Water Process Evaluation Facility, the University of Natal Pollution Research Group and Explochem Water Treatment (Pty) Ltd. It was proposed that the process be installed at the Umgeni Water Process Evaluation Facility and operated so that delegates to the International Water Supply Association (IWSA) conference in Durban 1995 would be able to assess the technology.

1.1 OBJECTIVES

The aims of the project were to :

- Develop and demonstrate the Exxpress Process for the dewatering of Waterworks sludges.
- Apply the Exxpress Process to the production of potable water from river water.

In order to structure the research on improving the design of the Tubular Filter Press, and assessing the performance of the unit for the dewatering of waterworks sludges and the production of potable water, specific key areas which were investigated are outlined below.

1.1.1 Improve the Design of the Tubular Filter Press

During the development and operation of the Tubular Filter Press at the H.D. Hill Waterworks, weaknesses in the design were identified. These included cloth splits, tube blockages, cleaning heads and manifold layout. The design of the Tubular Filter Press to be installed at the Umgeni Water Process Evaluation Facility at Wiggins Waterworks should address these wherever possible.

1.1.2 Develop a Commercially Operational Unit for Waterworks Sludges

Close liaison with Explochem, Dr. Rencken and the University of Natal, should ensure that the design of the plant is such that it will adequately demonstrate the process as an operational unit. To achieve this, a detailed design of the new plant should be produced.

This will include :

- Detailed mechanical drawings
- Detailed layout of the plant
- Detailed electrical drawings
- Final costing for capital installation

1.1.3 Monitor the Capital and Operating Costs

Capital and operating costs should be lower than similar processes in the industry, and the ease of operability should be addressed. Every effort will be made to produce a more cost effective process. (The low capital budget will assist in limiting the expenditure). An estimation of the operating costs should be performed once the demonstration plant is operational.

1.1.4 *Assess the Performance of the New Design*

Operating data from the new plant and other single tube configurations should be used to establish the operating parameters and performance of the new design.

1.1.5 Assess the Use of the New Design for Potable Water Production

The filtrate from the dewatering of waterworks sludges will be sampled and analysed to determine whether the filtrate quality complies with World Health Organisation guidelines for potable water quality. The use of similar technology, Crossflow Microfiltration, has resulted in excellent potable water results. It is proposed that the new design should be operated in a similar manner to produce an acceptable quality potable water.

1.1.6 Develop Techniques for Determining Design Parameters for the Tubular Filter Press

Previous work by Rencken (1992) showed that there are distinct differences in the filterability of water works and other sludges. His thesis details laboratory filtration tests which need to be performed in determining filtration characteristics of sludges. Dr Rencken also proposed a model for the filtration of waterworks sludges in the Tubular Filter Press. It is proposed that the theoretical filtration model and experimental techniques for determining model parameters, documented by Rencken be extended, and a mathematical model be developed for the new Tubular Filter Press and that a design procedure be recommended.

1.2 LITERATURE REVIEW

A comprehensive literature review on filtration theory has been provided by Rencken (1992). A further literature review was conducted in order to :

- determine if there had been any subsequent advances or significant improvements to the existing models and techniques documented by Rencken (1992),
- identify any weaknesses in the theoretical model or experimental techniques proposed by Rencken (1992), and
- identify and assess any alternative filtration theories that could be used to model the Tubular Filter Press.

No improvements in the compressible constant pressure filtration modelling techniques or the experimental techniques to determine model parameters (permeability and porosity correlations) subsequent to those utilised by Rencken (1992) where identified.

A number of assumptions are made in the conventional filtration theory which could possibly lead to problems in the application of the filtration model developed by Rencken (1992) to solid/liquid systems other than those studied by Rencken.

Equilibrium porosities and hence equilibrium cake structures are assumed to be attained instantaneously with changing solids compressive pressure during filtration. No models were identified in the literature, and it seems impossible to develop a *Dynamic Filtration Model* that can account for consolidation of filter cake structures during the filtration. The effects of this assumption are discussed in **Section 3.7**.

The solid particles of the filter cake are assumed to be in point contact with one another. All models identified in the literature use this assumption. A novel generalised area contact model has been developed that can account for any area contact that may exist in filter cakes. It is assumed that cake structures obtained in compression-permeability cell testing are representative of those obtained during filtration. Certain inadequacies in compression-permeability cell testing are identified and discussed in Section 3.5.

A relatively new filtration theory based upon the multiphase equations of change has been identified (Willis and Tosun 1980; Willis, Collins et al. 1983; Willis 1983; Tosun and Willis 1985; Willis, Tosun et al. 1985; Tosun and Sahioglu 1987; Tosun and Willis 1989; Willis, Tosun et al. 1989; Tosun, Yetis et al. 1993). The new *Single Resistance Model* differs considerably from the conventional *Two Resistance Model* utilised by Rencken (1992).

Generalised multiphase equations of change are developed from the fundamental concepts of the conversion of mass and momentum, using volume averaging techniques to shift from the unmeasurable microscopic local property level (i.e. within the pores of the filter cake) to the mathematically smooth, volume averaged level. These generalised expressions are then simplified by a series of assumptions that are most likely to be encountered in filtration practice e.g. that the process is isothermal, that a single solid particulate phase is non-deformable and insoluble in a single Newtonian liquid phase etc. to obtain the set of four multiphase equations of change (a continuity and motion equation for each phase) that describe the filtration process.

The development of this theory differs entirely from conventional filtration theory which was essentially developed on an analogy with Ohm's law for two resistances in series (the medium and the filter cake) in an electrical circuit (the filtration pressure and filtrate flow rate being analogous to the electrical potential and current respectively).

It was decided that this new filtration theory would be unsuitable as an alternative theory to model the Tubular Filter Press, as the model is mathematically rigorous and extending the model to the internal cylindrical geometry would be extremely difficult. The model also requires that the pressure and porosity profiles through the cake need to be empirically determined. These profiles are normally measured by pressure taps and electroconductive porosity probes within the filter cake, the pressure taps are connected to pressure transducers for local fluid pressure measurement and the porosity probes are designed to measure local electrical conductivity which is correlated to local porosity. These experimental techniques and equipment are far more sophisticated than those required by the conventional *Two Resistance Model*.

1.3 SUMMARY OF PROGRESS IN MEETING THE OBJECTIVES

A design committee was formed to discuss the weaknesses in the design of the H.D. Hill Tubular Filter Press and propose solutions to the operating problems experienced. It was decided that a completely new design using vertical tube orientation would adequately solve most of the operating problems. A single-tube pilot plant was constructed at the Umgeni Water Process Evaluation Facility and following the successful operation of the pilot plant, a Vertical Tubular Filter Press was designed and constructed prior to the IWSA conference in Durban (1995).

A new filtration model based on the operating experiences of the Vertical Tubular Filter Press was developed, and this model can be used to assist with the design and optimisation of new installations. The model is being incorporated into a user-friendly computer software package.

This report details the methods and procedures used for the operation of the laboratory test equipment and the pilot plants. In an assessment of the design the results of the operation of the plant are reported. The vertical tubular configuration was assessed for the production of potable water from river/surface water, and the problems that were experienced are documented.

2 Design of Full-Scale Express Unit

In order to structure the research on improving the design of the Tubular Filter Press, the operation of the plant at the H.D. Hill Waterworks was assessed. As shown in **Figure 2.1**, waterworks sludge is pumped through a set of horizontal tubes in crossflow, and the pressure inside the tubes is controlled, thereby allowing filtration of the sludge through the permeable woven fabric. During the filtration the solids form a cake inside the tubes and once the flowrate reaches a preset value, the tubes are cleaned by an automated roller which moves along the length of the tubes creating a high velocity through the inside of the tubes effectively removing the solids onto a conveyor belt.

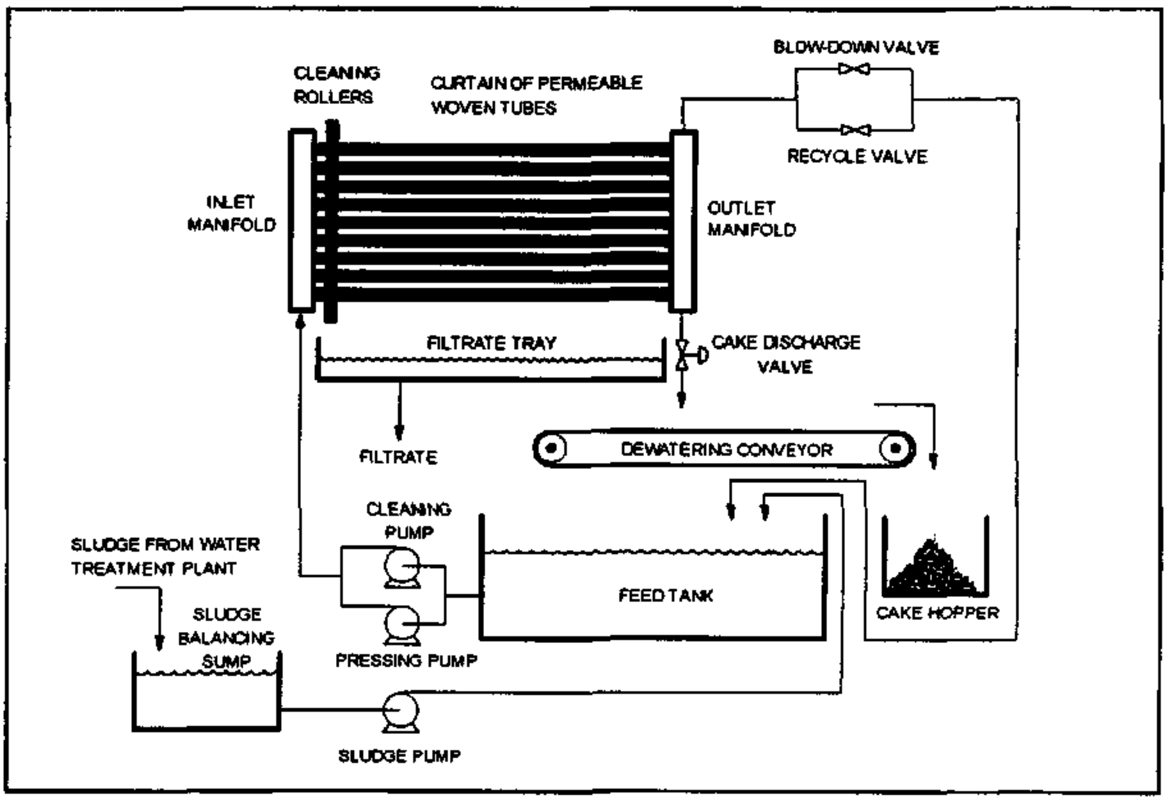

Figure 2.1 **Process Flow Diagram of the H.D. Hill Tubular Filter Press**

2.1 PROBLEMS EXPERIENCED WITH THE PREVIOUS DESIGNS

During an assessment of the performance of the plant a number of issues were raised that required attention. The weaknesses of the previous design at the H.D. Hill Waterworks were identified as follows :

- Cloth splits resulted in significant down time and loss of production as well as excessive operating costs in replacement of curtain material.
- Tube blockages as a result of uneven distribution of waterworks sludge during the filtration stage and insufficient flow during cleaning of the tubes. This was exaggerated by the narrow tube diameter used on the plant, as well as the length of the tubes and the associated hydraulic pressure gradients.
- Cake recovery was found to be low and needs to be improved to increase the capacity of the Tubular Filter Press.
- Cleaning heads were often problematic and required a large degree of mechanical maintenance.
- The separation of the permeate and the sludge during filtration and washing of the tubes is not adequate and can be improved.
- Conveyor belt blockages were also experienced causing inadequate draining of the cake, and some carryover of thin sludge together with the dewatered cake reduces the efficiency of the process.
- The cleaning carriage was driven along the length of the curtain by a motor on one side of the curtain. Frequent slippage was noticeable and the carriage would move out of alignment.
- The control strategy of the plant was not optimised and some attention is needed to ensure a smooth process operation.
- In order to overcome some of the tube blockage problems and improve the reliability of the process, the properties (especially the viscosity) of the cleaning fluid need to be considered.
- The spray hose which feeds water to the cleaning mechanism rolls onto a hose reel on one side of the cleaning carriage. This is seen to be an added mechanical system which adds to the complexity of the process.
- The manifold layout is considered to be problematic, and a primary reason for the poor flow distribution and tube blockages.
- Recording and instrumentation is required to ensure that the plant is optimised and that sufficient data are recorded to fully understand the process.

2.2 DESIGN SUB-COMMITTEE

A design sub-committee was proposed to address the problems experienced at the H.D. Hill Tubular Filter Press. It was stressed that the faults should be correctly engineered, and included in the design of the Exxpress Unit.

A technical subcommittee included the following members or alternates:

K. Treffry-Goatley, E. Coopmans	(Explochem)
G. Rencken	(Private Capacity)
M. Pryor	(Umgeni Water)
C. Buckley, V.L. Pillay	(University of Natal)
V. van Eck	(Thompson & van Eck / Explochem)

Two meetings of the technical sub-committee were held in Johannesburg, at which specific design and related issues were discussed.

2.3 DESIGN SPECIFICATIONS

As an alternative to the H.D. Hill design, it was decided that vertically mounted tubes, shorter in length with a larger diameter would have the following advantages over the previous design :

- Tube blockage potential is likely to be lower because of the shorter length and larger diameter of tubes.
- Cake release and conveyance out of the tubes is likely to be easier because of gravity. This may result in improved cake recovery.
- Cloth hanging mechanisms are simpler and the footprint of the unit will be smaller.
- Tubes can be employed easier in a vertical configuration than a horizontal configuration.

It was considered essential that a single-tube pilot plant be built, before a full-scale unit be developed. The project programme was structured to provide an operating unit in time for the International Water Supply Association (IWSA) Conference in Durban during September 1995. Based on the initial results of the single-tube experiments, the decision was made to continue with the design and construction of a vertical-tube pilot plant at Wiggins Waterworks. The design proposed by the technical subcommittee is shown in **Figure 2.2**.

Further discussions centred around specific aspects of the design. The most intricate and complicated aspect being the design of the cleaning mechanism and the operation strategy for the plant. The following aspects were discussed and resolved in the design meetings.

2.3.1 Modular Design

It was decided that modules should be constructed, each consisting of a double curtain. It was suggested that the Vertical Tubular Filter Press at Wiggins Waterworks should consist of three such modules, each with its own roller mechanism. In this way the plant could easily be sized and adapted in future for larger applications.

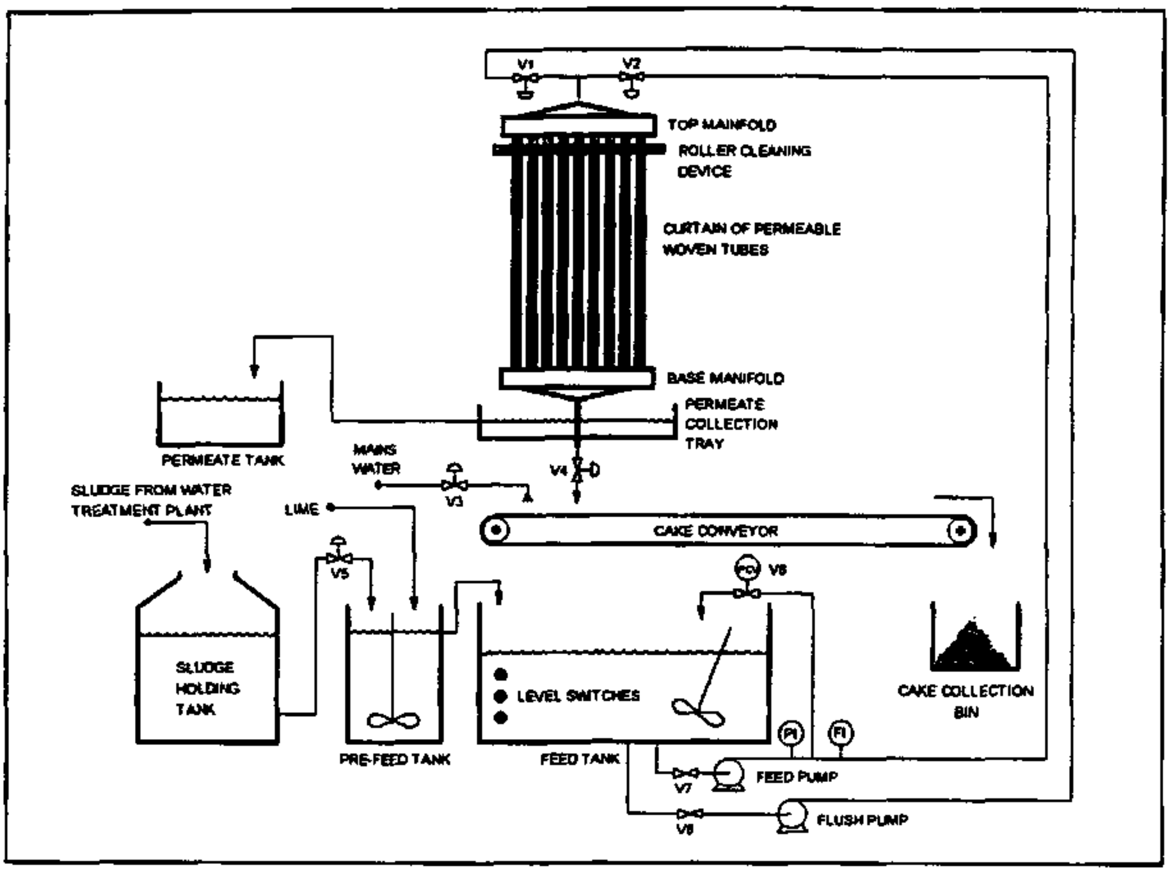

Figure 2.2 **Process Flow Diagram for the Vertical Tubular Filter Press**

2.3.2 Double Curtain per Module

A double curtain, **Figure 2.3** was considered to achieve a higher filtration area per square metre of floor area. The double curtain would be moulded into a top and bottom manifold. The cleaning rollers would then squeeze the two curtains together, forming the restriction for effective cleaning, whereas previously at the H.D. Hill Waterworks the single horizontal curtain was cleaned by a system of two rollers per curtain.

2.3.3 Inlet Feed Manifold Design

The manifold design was identified as a potential problem area. The even distribution of sludge to each tube of the curtain is necessary to ensure effective cleaning and reduce the likelihood of tube blockages. The vertical tube design, with considerably shorter and larger diameter tubes, results in reduced pressure drop along the length of the tube, and subsequently fewer flow related problems.

The proposed design will also be operated in *Dead-End Filtration* which will favourably affect the pressure distribution.

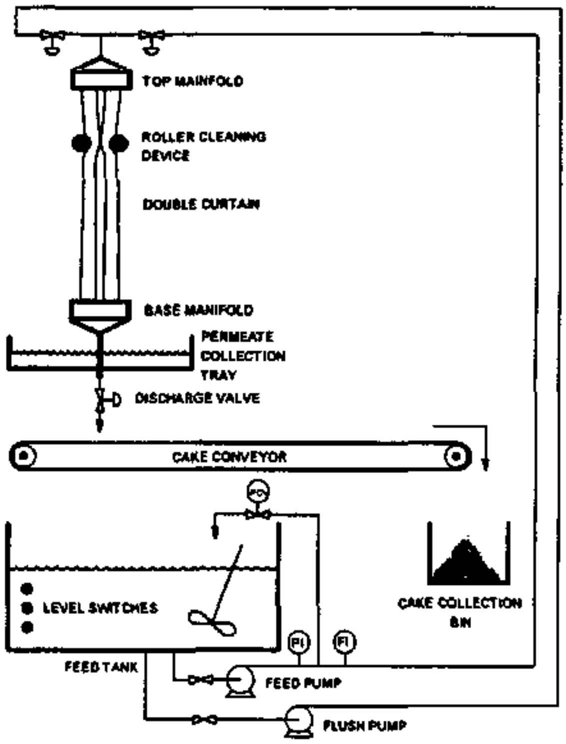

Figure 2.3 Roller Cleaning of Double Curtain

The inlet manifold was designed with an orifice restriction at the inlet to each tube. This will create an even flow distribution while the tubes are filling up, and hopefully result in an initial even distribution of solids. The cleaning of the tubes is critical for even flow distribution. It has already be shown in the WRC project 386, *The development of a Crossflow Microfilter* that blockages of tubes can result from uneven distribution of wash water. The orifice plate in the inlet manifold should assist in creating even wash flow through the tubes.

2.3.4 Size of Tubes

It was agreed that the diameter of the tubes at the Tubular Filter Press at the H.D. Hill Waterworks was a major contributing factor to the occurrence of tube blockages. It was proposed that a larger size tube be considered. Previous work during the development of the Tubular Filter Press had shown that a 40 mm tube diameter was unable to withstand as high a pressure as the smaller diameter tubes.

A new weave of fabric has since been used in the manufacture of the curtains. The fabric has demonstrated a higher resistance to pressure and wear during cleaning. For these reasons a larger diameter tube was considered. The Vertical Tubular Filter Press at Wiggins Waterworks was built using 60 mm tubes.

Although the larger tube will prevent the likelihood of tube blockages, a more expensive and stronger cleaning mechanism is required to withstand the forces during cleaning. The Vertical Tubular Filter Press built at Wiggins Waterworks was designed to withstand the forces for 100 mm tubes. This will give the flexibility to test different tube sizes in the future.

2.3.5 Cleaning Mechanism Design

The cleaning mechanism has the most intricate design in the whole plant, and its design took a lot longer than was originally anticipated. The design includes a pair of rollers which are engaged at the top of the curtain module, move down the curtains and are released once they reach the bottom of the curtain.

The problem with the design was mainly the structural strength of the roller mechanism. In order to produce an even gap between the rollers over the full width of the curtain, the forces in the centre of the roller had to be calculated. It was agreed that the design should be flexible for tube sizes between 40 mm and 100 mm. The rollers had to be sized for the worst case scenario i.e. when the forces are maximum - 100 mm tubes. This resulted in a delay in the programme.

A limitation in the Tubular Filter Press at the H.D. Hill Waterworks was the mechanism which moves the cleaning mechanism along the curtain length. In the new design, the length of travel of the cleaning carriage is reduced. After various suggestions of using pneumatics and hydraulic drivers, the final design incorporates the use an electric motor with a gear and chain mechanism to move the cleaning carriage vertically.

2.3.6 Lime Addition

Previous work at the H.D. Hill Waterworks established that the addition of lime to a bentonite sludge creates a more filterable slurry with a lower specific cake resistance and improved cake recovery. It is necessary to include the facility to dose lime into the sludge to assess the performance with and without lime addition.

2.3.7 Future Expansion

Although a single module plant (consisting of a double curtain) was designed, it is necessary to make provision in the design to increase the capacity of the plant to a three-module plant. The pumps and peripheral equipment has to be sized accordingly, but only one module assembled. The option of increasing the size of the plant will be to demonstrate the operation of a full-scale unit.

2.4 CONSTRUCTION OF THE DEMONSTRATION UNIT

Following the completion and approval of detailed mechanical drawings, an order was placed with Explochem for the manufacture and construction of the Vertical Tubular Filter Press as a demonstration unit. The time schedule for the completion of the unit before the IWSA conference was extremely tight. Umgeni Water constructed a pilot-scale testing bay at the sludge plant at Wiggins Waterworks to house the demonstration unit.

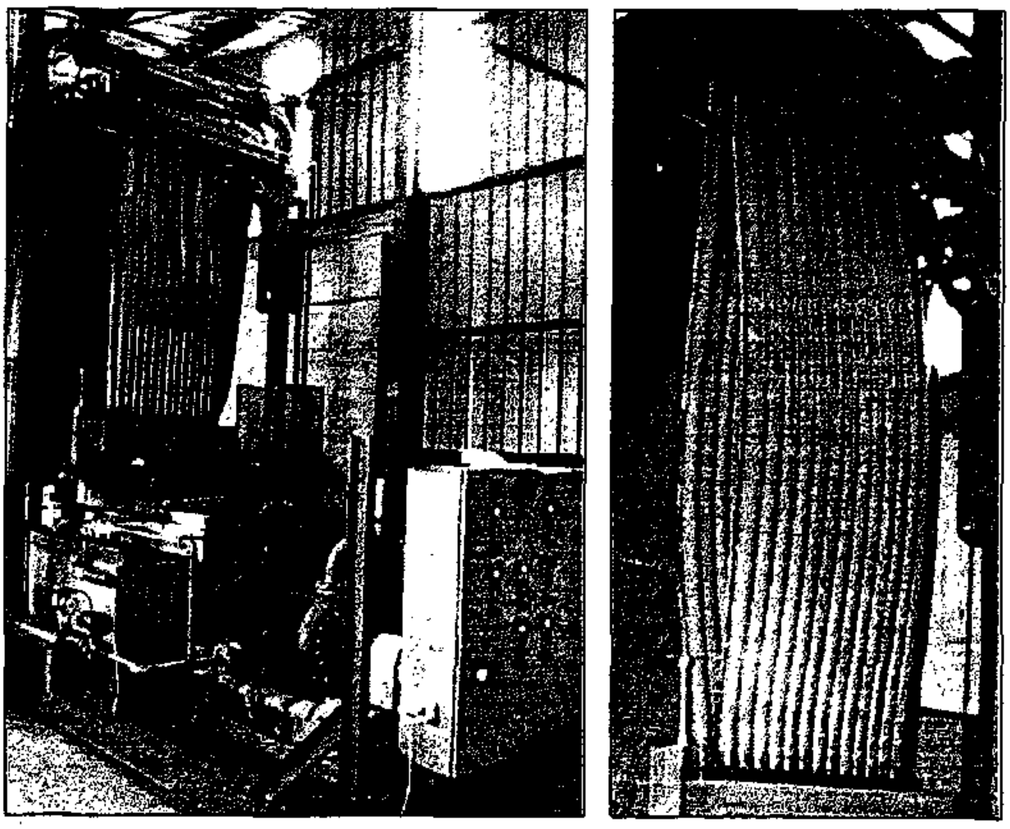

Figure 2.4a **The Vertical Tubular Filter Press Constructed at Wiggins Waterworks Showing the Skid and the Double Curtain During Filtration**

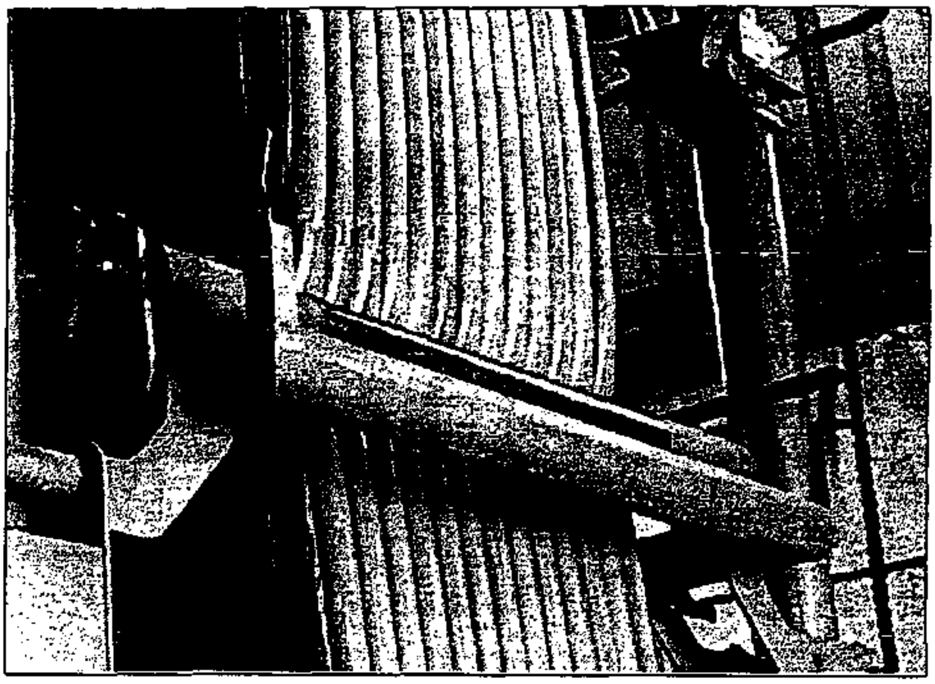

Figure 2.4b **Roller Cleaning of the Vertical Tubular Filter Press**

During the manufacture of the unit, supply problems were encountered in the purchase of curtain material from Gelvenor Textiles as only a short length of fabric was required. Encouraged by the successful testing of single tubes on the small-scale rig, the unit was modified to accommodate 22 single tubes (from Swiss Silk Textiles), instead of a double curtain. The moulding (connection) of the single tubes to the manifold and end blocks was also much simpler.

During the IWSA conference a technical tour to Wiggins Waterworks was scheduled, and during the tour sludge was pumped into the tubes at a low pressure to demonstrate the technology.

2.5 MODIFICATIONS AND IMPROVEMENTS TO THE DESIGN

During the commissioning of the unit a number of areas were highlighted as requiring modification to comply with operational and safety requirements at Umgeni Water, OHSA Act (1993) :

- **Splash Box** - During flushing, fluid and sludge inside the tubes is discharged onto a conveyor belt. The force and flowrate of the sludge, caused significant splashing resulting in extreme wastage. A splash guard was manufactured in order to contain the sludge on the conveyor belt. This was unsuccessful, but after a number of attempts a splash box was manufactured which prevents the splashing.

A further problem was encountered whereby some of the sludge removed from the tubes was washed off the conveyor belt into the feed tank, resulting in lower recoveries. It is presumed that the splash box adequately prevents this from occurring.

- **Splash Curtain** - During the action of rollers on the tubes, a large amount of splashing occurred from the top of the tubes. A concern of the project team was the safety of operating material tubes under pressure, and the compliance with pressure vessel regulations of the Occupational Health and Safety Act. The Tubular Filter Press is however not classed as a pressure vessel as no gaseous phase is under pressure. It was recognised that should a tube burst, not only could an operator become drenched in sludge but that the feed pump would continue pumping sludge out of the unit into the surrounding building. A plastic curtain was erected around the tubes to contain any splashing and sludge within the bounds of the feed tank.

- **Feed Magnetic Flowmeter** - During the initial operation of the unit, permeate flux was measured as the filtrate flowrate leaving the unit. This was found to be difficult as at low flowrates. One of the limitations of the Tubular Filter Press at the H.D. Hill Waterworks was identified to be inadequate instrumentation to perform optimisation. A magnetic flowmeter was installed in the feed line, which not only gives an instantaneous flowrate of feed sludge to the filter, but also displays a total volume of sludge filtered. This made monitoring of the unit much easier. Unfortunately due to a limited budget additional instrumentation could not be purchased.

- **Automation of the Flushing Sequence** - It is required that the unit clean the tubes once a specific minimum permeate flux is reached. The Programmable Logic Controller (PLC) was initially set up to measure the period of time that it took for the level in the feed tank to drop between two level probes. At very low fluxes, it was found that this time was as long as 30 min., and when the feed concentration was low (< 8 g/l) this was often too long for effective monitoring and the operation of the filter had to be stopped prematurely. This was considered to be inaccurate and once the magnetic flowmeter had been installed, a method was devised by measuring the time interval between subsequent pulses generated by the magnetic flowmeter (1 pulse per litre) to signal the start to the flushing sequence.

- **Mixing of the Feed Tank** - During the operation of the Vertical Tubular Filter Press, difficulties were experienced in the settling of solids in the feed tank. It was assumed that with continual operation, there would be sufficient movement in the tank to avoid the necessity for the installation of a mixer. During the filtration cycle at low fluxes, it was found that the feed solids concentration entering the tubes was not uniform, creating pressure fluctuations and inconsistencies in the flowrate. It was decided that a mixer in the feed tank would significantly improve the operation of the plant.

- **Roller Movement** - During initial operation, the roller mechanism was operated by an electric motor and chain hoist from one end only. Whilst the rollers were engaged and moving in a downward direction, the movement was smooth. A problem was observed when the rollers were being raised to the top position, the opposite end of the roller mechanism was found to occasionally snag, which caused a vibration in the whole structure. A modification was made to drive the roller mechanism from both ends by introducing an axle across the top of the unit connected to a similar lifting chain on the opposite end of the roller, thereby improving the operation.

- **Power Tripping of the Flush Pump** - Extensive problems were experienced with continual tripping of the flush pump. The flush pump has a large motor requiring a star / delta starting sequence. On the change over from star to delta, the power would often trip causing many runs to be aborted. This was eventually solved by installing a larger size circuit breaker to the motor power supply in order to prevent motor tripping during times of high starting currents. This emphasised the need to allow for contingencies in the budget for electrical specification and design.

- **Availability of Sludge from Wiggins Waterworks** - During 1996/7 the Wiggins Waterworks underwent an upgrade in capacity from 175 to 350 Ml/day. Unfortunately during the course of this project the sludge plant was upgraded and the availability of sludge was erratic. As a result of a large sludge holding tank being taken out of commission and the Dissolved Air Flotation thickening unit not performing adequately, the feed concentration was not consistent and varying between 6 and 30 g/l.

- **Sludge Solids Collection** - Initially the conveyor discharged the dewatered cake into a tank where it was re-slurried with the permeate from the filter, and returned to the sludge plant at Wiggins Waterworks. The problems with this were that not all the dry solids could be collected for calculation of recoveries. A sludge tray was manufactured and placed over the reslurry tank to collect all the solids from the conveyor.

Unfortunately this was not large enough and an accurate mass balance could not be performed. In order to effectively collect all the solids, the reslurry tank was moved and a separate sludge hopper was installed under the conveyor to collect the dry product. The hopper with sludge is then weighed and after sampling of the dry product, the remaining sludge is manually tipped into the reslurry tank. Further modification was required prior to the operation of the the unit on a continuous basis.

- **Tube Failure and Splitting** - After approximately six months of operation, the single tubes started to progressively develop splits and rupture. This began after trying to operate the filter at pressures above 350 kPa. It had been a concern of the design team that tubes larger than 40 mm would not be able to withstand high operating pressures.

- **Installation of a Tube Curtain** - Although it has been proven that the latest curtain fabric can withstand higher operating pressures, and considering that the curtain was not available at the time of plant construction, the single tubes were installed. The occurrence of tube failure emphasises the need for a durable fabric which must be proven adequate to withstand the recommended operating pressures (up to 450 kPa). A length of fabric wass acquired from Gelvanor Textiles, moulded into two curtains and installed on the Vertical Tubular Filter Press.

- **Spray on the Conveyor** - Over a period of time it was noticed that the solids that were collected contained a large amount of liquid sludge, implying that the liquid was not draining fast enough or that the conveyor aperture was too small. The speed of the conveyor could not be reduced to increase the rate of draining and a larger aperture conveyor could not be obtained. A clean water spray was installed to wash the conveyor.

- **Blockage of the Conveyor Belt** - The installation of a spray over the conveyor belt improved the cleaning of the conveyor to an extent. The aperture of the openings in the conveyor appeared too small. A larger aperture belt was eventually sourced and especially manufactured. Once installed the draining of the sludge improved with a possible small loss in recovery form the filter.

- **Pressure Trip Switch on the Feed Pump -** During a safety inspection of the plant, further items were identified as requiring modification to comply ot the safety requirements, OHSA Act (1993). All pressure gauges should have an indication of maximum safe working pressure. Mechanisms should also be provided to prevent the occurrence of excessive pressure where a limit is indicated.

  A pressure switch was installed in the feed line after the feed pump. The signal from the switch was fed as an input into the PLC as an automatic cut off which stops the plant should the tubes block to such an extent that they are subjected to excessive pressures. The pressure regulating valve could fail or the wash cycle may not initiate resulting in situations which may also create over-pressurisation. The pressure switch therefore provides protection against all these situations.

3 Constant Pressure Compressible Cake Filtration Model

In this chapter a mathematical model for constant pressure compressible cake filtration is presented and the experimental techniques required to obtain empirical parameters for the model are documented. A new generalised *Area Contact Model* is developed and some aspects unique to the modelling of the Tubular Filter Press are discussed.

3.1 PERMEABILITY AND POROSITY CORRELATIONS

The permeability and porosity of compressible cakes are functions of the solids compressive pressure. The solids compressive pressure arises from the frictional drag force on the particles as the filtrate flows through the cake, and is defined as the cumulative frictional drag force divided by the cake cross-sectional area perpendicular to the direction of filtrate flow. It is observed (Tiller and Cooper, 1962) (Tiller and Leu, 1980), that the functional relationship is exponential with respect to the solids compressive pressure and the two commonly accepted correlations are presented below.

Tiller and Cooper (1962) proposed the following set of equations:

$$K = Fp_s^{-\delta} \qquad p_s > p_{si} \qquad (3.1.a)$$

$$K = K_i = Fp_{si}^{-\delta} \qquad p_s \le p_{si} \qquad (3.1.b)$$

$$(1-\varepsilon) = Bp_s^{\beta} \qquad p_s > p_{si} \qquad (3.1.c)$$

$$(1-\varepsilon_i) = Bp_{si}^{\beta} \qquad p_s \le p_{si} \qquad (3.1.d)$$

where F, δ, B, β = empirical constants, (-)

K = permeability, (m^2)

ε = porosity, (-)

p_s = solids compressive pressure, (Pa)

K_i = permeability of cake when $p_s \le p_{si}$, (m^2)

ε_i = porosity of cake when $p_s \le p_{si}$, (-)

p_{si} = solids compressive pressure below which the permeability and porosity are assumed constant, (Pa) and Tiller and Leu (1980) presented an alternative set of equations:

$$K = K_0\left(1 - \frac{p_s}{p_a}\right)^{-\gamma} \tag{3.2.a}$$

$$(1 - \varepsilon) = (1 - \varepsilon_0)\left(1 - \frac{p_s}{p_a}\right)^{\lambda} \tag{3.2.b}$$

where γ, λ, p_a = empirical constants, (-)

K_0 = permeability at $p_s = 0$, (m^2)

ε_0 = porosity at $p_s = 0$, (-)

Various tests can be performed to obtain permeability and porosity data over a range of solids compressive pressures and the parameters in the above equations are obtained by numerical regression on the data. Both the correlations above have the same number of empirical parameters.

Compression-Permeabilty cell (C-P cell) tests are used to obtain permeability and porosity data in the high solids compressive range, approximately 50 to 500 kPa. Although these tests have been used at lower pressures, reservations have been expressed on the accuracy of the data at low pressures mainly due to wall friction effects. Murase et al. (1989) proposed a centrifuge method for the determination of porosity data in the intermediate solids compressive pressure range, approximately 1 to 100 kPa. Rencken (1992) found that when the results from centrifuge tests were incorporated into the filtration model, the accuracy of the model output decreased. As such, centrifuge tests will not be considered as an important method for obtaining data for the correlations. Shirato et al. (1983) proposed a batch settling test for the determination of permeability and porosity data in the low solids compressive pressure range, typically below 1 kPa. These settling tests require simple and inexpensive testing equipment.

Equations 3.1 assume that the permeability and porosity are constant below some low solids compressive pressure, p_{si}. Caution should be exercised in determining the value of p_{si} since relatively small variations in the value can have marked effects on the filtrate volume versus time predictions of the filtration model. According to Tiller and Leu (1980) the location of p_{si} is entirely empirical and relatively arbitrary, as C-P cell tests are not sufficiently accurate at low solids compressive pressures to accurately determine its value. Tiller and Leu (1980) introduced **Equations 3.2** to overcome this problem, by defining a permeability and porosity at a solids compressive pressure of zero. However, Shirato et al. (1983) subsequently introduced the settling method for determining permeability and porosity data at very low solids compressive pressures, which enables the location of p_{si} to be determined more accurately.

Equations 3.2 are restricted in that they are intended to be fitted over the entire range of permeability and porosity data, **Equations 3.1** however can be extended to include multiple empirical parameter sets which are valid over sections of the solids compressive pressure range. This is important because various authors (Tiller et al. 1987) have expressed reservations about the validity of these correlations for highly compressible cakes as the range over which these equations apply decreases with increasing compressibility. Rencken (1992) found that the fit between **Equations 3.2** and experimental data were not good and when the regressed parameters were incorporated into the internal cylindrical filtration model the agreement between experimental data and the model predictions were poor.

The extended form of **Equations 3.1** for multiple correlation parameter sets is as follows:

$$K = K_i = F_1 p_{si}^{-\delta_1} = F_1 p_{s1}^{-\delta_1} \qquad p_s \leq p_{s1} \qquad (3.3.a)$$

$$K = F_n p_s^{-\delta_n} \qquad p_{sn} < p_s \leq p_{s(n+1)} \qquad (3.3.b)$$

$$K = F_{(n+1)} p_s^{-\delta_{(n+1)}} \qquad p_s > p_{s(n+1)} \qquad (3.3.c)$$

$$(1-\varepsilon_i) = B_1 p_{si}^{\beta_1} = B_1 p_{s1}^{\beta_1} \qquad p_s \leq p_{s1} \qquad (3.3.d)$$

$$(1-\varepsilon) = B_j p_s^{\beta_j} p_{sj} < p_s \leq p_{s(j+1)} \qquad (3.3.e)$$

$$(1-\varepsilon) = B_{(j+1)} p_s^{\beta_{(j+1)}} \qquad p_s > p_{s(j+1)} \qquad (3.3.f)$$

$$n \in [1, M]$$

$$j \in [1, T]$$

where $M \geq 1$ = the number of permeability correlation parameter sets, (-)

$T \geq 1$ = the number of porosity correlation parameter sets, (-)

n, j = integer numbers, (-)

Depending on the compressibility of the cake, there will be a correlation parameter set for each test type utilised, or if a test type covers a large solids compressive pressure range, the data within a test type may be sub-divided and parameter sets obtained in each region.

3.2 A NEW GENERALISED AREA CONTACT MODEL

Previously, the development of filtration models assumed that the particles of the filter cake are in *point contact* with one another and the liquid pressure is effective over the entire cross-sectional area of the cake (Tiller, 1966). The *point contact* model is however idealistic and in actual filter cakes there will be area contact between the particles, see **Figure 3.1**.

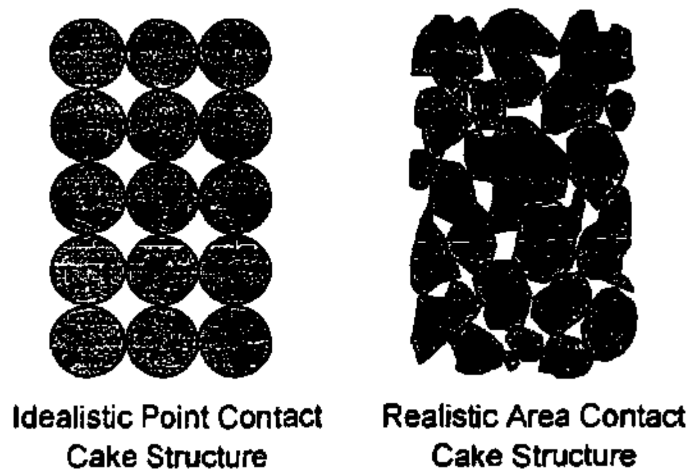

Figure 3.1 Representation of Contact Between Particles in a Filter Cake

A new generalised filtration model that can account for any area contact that may exist between particles in a filter cake is presented below.

3.2.1 Area Contact Function

The local interparticle contact area is a function of the local cake structure or packing of the particles within the cake. An area contact function describing the relationship between the interparticle contact area in the plane perpendicular to the direction of filtrate flow and the cake structure is developed heuristically.

The cake structure is a function of the particle shape distribution, particle size distribution, particulate density (the number of particles present in a given volume), and the orientation of the particles with respect to one another. The interparticle contact area in the plane perpendicular to the filtrate flow direction through the cake will therefore be equal to the product of the area of this plane, and an area contact function:

$$A_c = A \times f_A(\Psi_{shape}, \Psi_{size}, \Pi_{density}, \Pi_{orientation}) \tag{3.4}$$

where

A	=	area of plane perpendicular to the direction of filtrate flow, (m^2)
A_c	=	interparticle contact area in the plane perpendicular to the direction of filtrate flow, (m^2)
$f_A(\ldots)$	=	area contact function, (-)
$\Pi_{density}$	=	local particulate density, (m^{-3})
$\Pi_{orientation}$	=	local particle orientation, (-)
Ψ_{shape}	=	local particle shape distribution, (-)
Ψ_{size}	=	local particle size distribution, (-)

For a given particle shape and size distribution, the local particulate density is directly related to the local solids volume fraction within the cake and hence the local cake porosity. Its effect can therefore be separated from the area contact function. It is assumed that the interparticle contact area is directly proportional to the local solids volume fraction within the cake. For compressible cakes the porosity is a function of the solids compressive pressure within the cake.

$$f_A = f_A(\Psi_{shape}, \Psi_{size}, \Pi_{orientation}) \times [1 - \varepsilon(p_s)] \tag{3.5}$$

The direct proportionality between interparticle contact area and the local solids volume fraction is only expected to hold for relatively high solids compressive pressures where the changes in porosity with respect to solids compressive pressure are not as pronounced, as the cake structure is more consolidated. At lower solids compressive pressures, small changes in solids compressive pressure can result in relatively large changes in cake porosity (and hence cake structure) and the relationship between cake porosity and area contact will be difficult to determine, but it is unlikely that it will behave proportionally as shown in **Equation 3.5**.

Particle shape and size distributions have a direct and strong influence on the area contact function. Particles may be in the shape of spheres, fibres, flakes etc. or more complex shapes due to particle association (flocculation and coagulation) to form particle structures. For example, flake-like particles will exhibit greater area contact behaviour than spherical particles and systems with a broad particle size distribution may exhibit greater area contact behaviour because the smaller particles occupy the interstices between the larger particles. For a given slurry type (ignoring the effects of small-scale solids migration), the particle shape and size distributions are assumed to be constant with respect to filtration time and distance through the cake, the area contact function can now be written as:

$$f_A = f_A'(\Pi_{orientation}) \times [1 - \varepsilon(p_s)] \tag{3.6}$$

For particles which are not symmetrical about a point (a sphere is an example of a particle that displays point symmetry), the orientation of the particles with respect to one another and the plane perpendicular to the filtrate flow direction within the cake will affect the cake structure and hence the interparticle contact area. It is assumed that the particle orientation is a function of the solids compressive pressure.

$$\Pi_{orientation} = g(p_s) \tag{3.7}$$

At the cake surface, the particles are deposited with a random orientation and the resulting cake structure is therefore random and unstable. As subsequent layers of cake are deposited, the solids compressive pressure increases. The increasing solids compressive pressure will begin to stress the underlying random cake structure and the particles will begin to re-align and re-orientate themselves in such a way as to form a more uniform and stable cake structure (this includes the deformation of associated particle structures such as flocs). Substituting **Equation 3.7** into **Equation 3.6** yields:

$$f_A = f_A'''(p_s) \times [1 - \varepsilon(p_s)] \tag{3.8}$$

The region of the cake near the cake surface where the solids compressive pressure is low is characterised by a cake structure which is very porous and composed of randomly orientated particles. The particles in this region will be assumed to be in point contact with one another and only as the solids compressive pressure increases towards the septum, and the cake structure becomes more compact and the particles begin to orientate themselves will they begin to be in area contact with one another, see **Figure 3.2**.

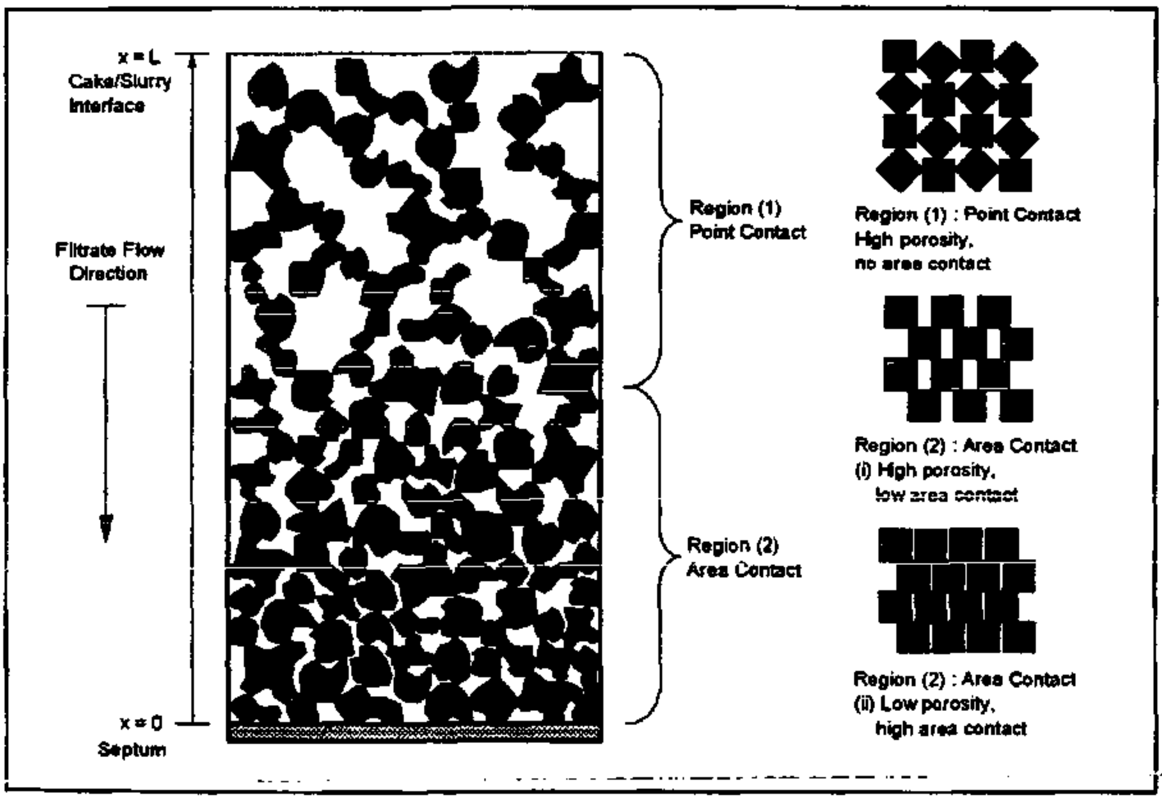

Figure 3.2 **Relationship Between the Porosity and Particle Orientation and the Inter-Particle Contact Area in the Cake**

For a given system, the area contact function is composed of two principle parts, the particulate density component and the particle orientation component, both of which are functions of solids compressive pressure. The cross over from *point contact* to *area contact* may be gradual or sudden depending on the behaviour of these two components with respect to solids compressive pressure. The behaviour of the particulate density component is characterised by the functional dependence of the cake porosity, however it is not expected to hold for low solids compressive pressures. The behaviour of the particle orientation component is difficult to ascertain and is assumed that it behaves as a step function i.e. as if the packing orientation exhibits a yield compressive stress.

$$f_A'''(p_s) = 0 \qquad 0 \le p_s < p_{sa} \qquad (3.9.a)$$

$$f_A'''(p_s) = A_0 \qquad p_s \ge p_{sa} \qquad (3.9.b)$$

where p_{sa} = packing orientation yield compressive pressure, (Pa)

A_0 = coefficient of area contact, (-)

It is also assumed that the particle density component of the area contact function holds for solids compressive pressures greater than the packing orientation yield compressive pressure, as the behaviour of the cake structure with respect to solids compressive pressure will be more predictable. As a result the packing orientation yield compressive pressure can be redefined as the *point-area* compressive pressure, and represents the solids compressive pressure at which the interparticle contact changes from point contact to area contact.

The coefficient of area contact, A_0, is assumed to be constant for a given system. The area contact in the plane perpendicular to the direction of filtrate flow can be determined from the following set of equations:

$$A_c = 0 \qquad 0 \le p_s < p_{sa} \qquad (3.10.a)$$

$$A_c = A[A_0[1 - \varepsilon(p_s)]] \qquad p_s \ge p_{sa} \qquad (3.10.b)$$

The area contact function states that below the *point-area compressive pressure* the particles in the filter cake are in point contact with one another. Above the point-area compressive pressure the particles begin to experience area contact, the extent of this area contact is given by the product of the local solids volume fraction and a constant area contact coefficient.

The ratio of the interparticle contact area to the area of the plane perpendicular to the filtrate flow direction is zero if the particles are in point contact with one another and has maximum possible value of 1. Since the porosity ranges from 0 to 1, the coefficient of area contact has a minimum value of 0 and a maximum value of 1.

$$A_0 = A_c/A \times 1/(1 - \varepsilon) \qquad A_0 \in [0, 1] \qquad (3.11)$$

3.2.2 Relationship between Liquid and Solids Compressive Pressure

In filtration, drag stresses arise from the interfacial transfer of momentum from the liquid to the solid particles as the liquid flows around the solid particles towards the medium in the direction of decreasing liquid pressure. The viscous drag occurs only if there is relative motion between the liquid and solids particles and is zero within the slurry and at the cake-slurry interface. Since the particles are in contact with one another, the frictional drag force on the solid particles is cumulative and reaches a maximum at the medium.

The theoretical relationships between the liquid and solids compressive pressures are derived from a force balance over a differential element of cake.

3.2.2.1 Planar Filtration

For one-dimensional filtration, the plane perpendicular to the direction of filtrate flow is any plane parallel to the septum. Summing the individual particle properties over the number of particles in the plane perpendicular to the direction of filtrate flow yields:

$$\sum_{i=1}^{N} A_i = A \tag{3.12}$$

$$\sum_{i=1}^{N} (A_c)_i = A_c \tag{3.13}$$

$$\sum_{i=1}^{N} (A_c + dA_c)_i = A_c + dA_c \tag{3.14}$$

$$\sum_{i=1}^{N} (F_s)_i = F_s \tag{3.15}$$

$$\sum_{i=1}^{N} (F_s + dF_s)_i = F_s + dF_s \tag{3.16}$$

where F_s = accumulated frictional drag force on particles, (N)

N = number of particles in the plane perpendicular to the direction of filtrate flow, (-)

A force balance on a planar differential element of cake neglecting inertial and accelerative effects, **Figure 3.3**, yields:

$$p_L(A - A_c) + F_s = (p_L + dp_L)(A - (A_c + dA_c)) + (F_s + dF_s) \tag{3.17}$$

where p_L = liquid pressure, (Pa)

The solids compressive pressure, which for planar filtration is given by the accumulated frictional drag force divided by the cake cross-sectional area perpendicular to the direction of the filtrate flow, is defined as:

$$p_s = F_s/A \tag{3.18}$$

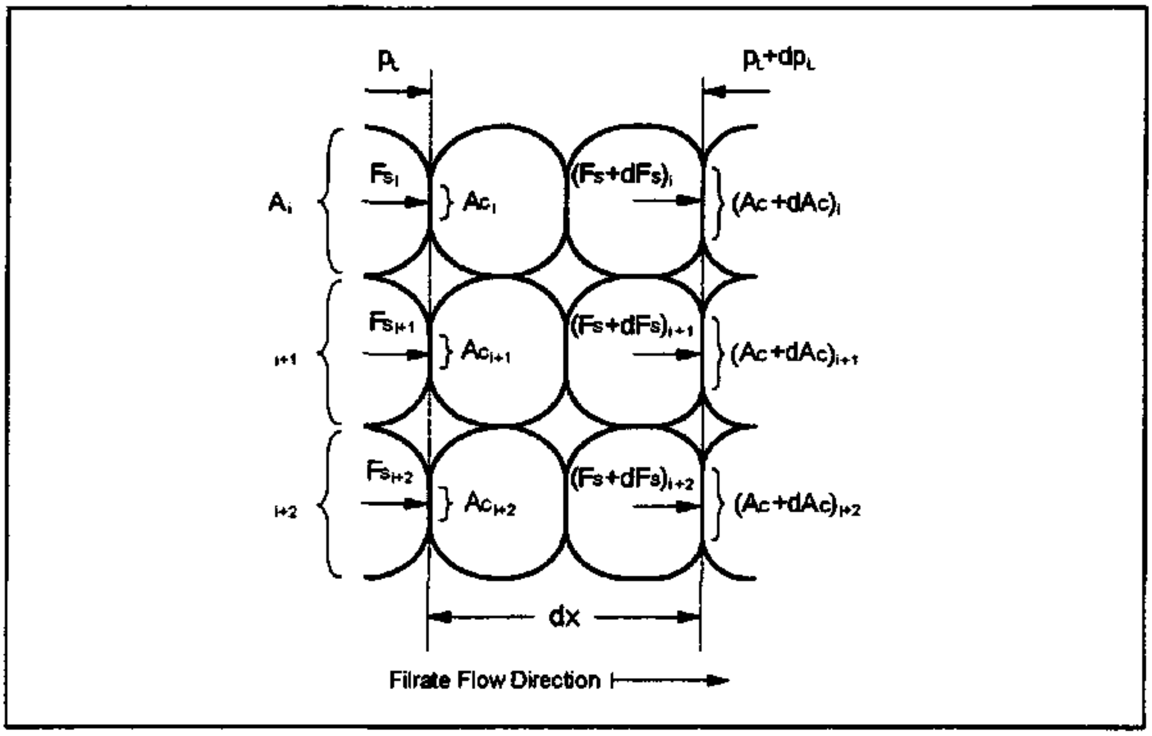

Figure 3.3 **Force Balance on a Planar Differential Element of Cake**

Combining **Equation 3.18** with **Equation 3.17** and simplifying, leads to:

$$dp_L(1 - A_c/A) - (p_L + dp_L)dA_c/A + dp_s = 0 \ (3.19)$$

From **Equation 3.10.b**, for $p_s \geq p_{sa}$ we have:

$$dA_c/A = -A_0 d\varepsilon \tag{3.20}$$

Substituting **Equations 3.10.b** and **Equation 3.20** into **Equation 3.19** leads to:

$$(1 - A_0 + A_0\varepsilon)dp_L + A_0 p_L d\varepsilon + dp_s = 0 \tag{3.21}$$

Equation 3.21 is novel. For $0 \leq p_s < p_{sa}$ and $A_0 = 0$ (i.e. point contact), **Equation 3.21** reduces to the familiar expression relating solids compressive pressure to liquid pressure for planar filtration.

$$dp_L + dp_s = 0 \tag{3.22}$$

3.2.2.2 Internal Cylindrical Filtration

For internal cylindrical filtration, the plane perpendicular to the filtrate flow direction is a cylindrical surface of constant radius, the set of **Equations 3.12** to **3.16** are valid over this surface. **Figure 3.4** shows the forces acting on a differential element of cake.

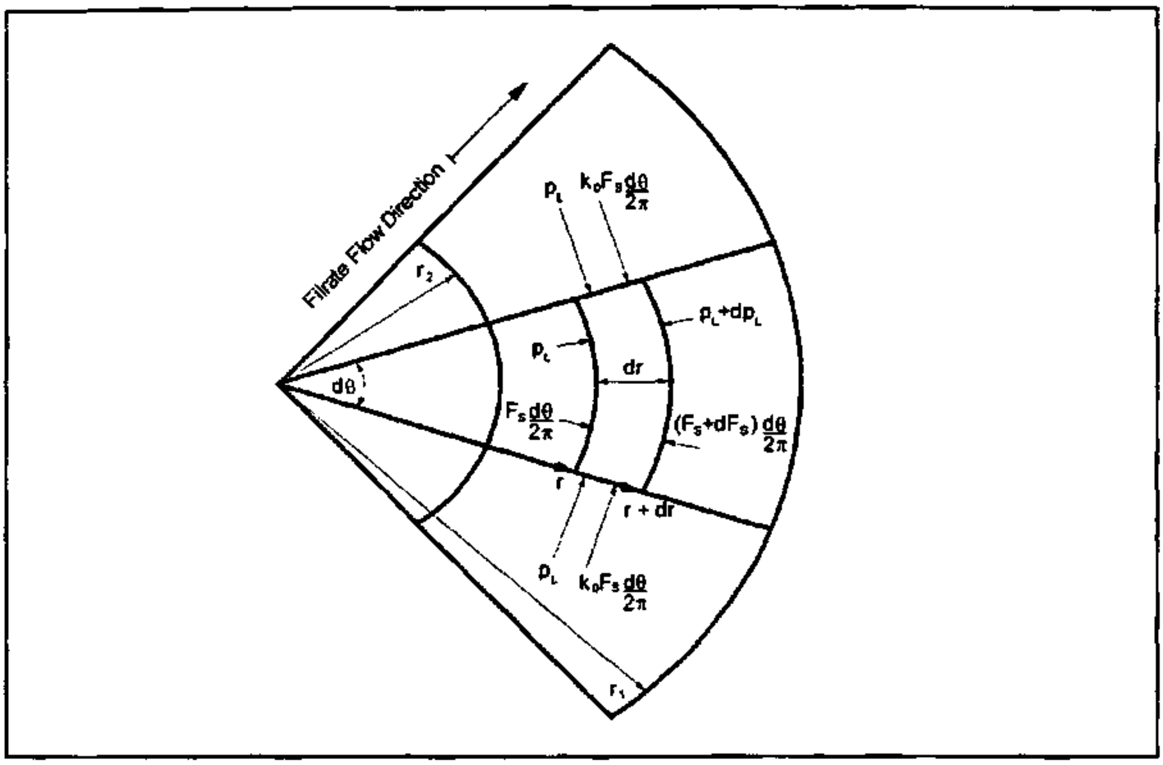

Figure 3.4 **Force Balance on an Internal Cylindrical Differential Element of Cake**

In addition to the forces acting in the radial direction, there are forces acting laterally on the sides of the element of cake. These forces arise from the liquid pressure acting on the sides of the element of cake and an effective solids force resulting from the accumulated frictional force acting on the element of cake in the radial direction **Section 3.4**. In planar filtration these forces are antagonistic and have no component acting in the direction of filtrate flow and are ignored in the force balance. In cylindrical filtration however, the geometry is such that these forces have a component acting in the radial direction that must be taken into account. **Figure 3.5** shows how these forces may be resolved into components.

Despite the fact that the area of the sides of the differential element of cake are parallel and not perpendicular to the direction of filtrate flow, it is assumed that the interparticle contact area in these surfaces can be approximated by the area contact function.

A force balance on the cylindrical differential element of cake, neglecting inertial and accelerative effects, yields:

$$p_L(rd\theta l)\left[1-\frac{A_c}{A}\right]+F_s\frac{d\theta}{2\pi}+2\sin\left(\frac{d\theta}{2}\right)\left[k_0F_s\frac{d\theta}{2\pi}+p_L(ldr)\left[1-\frac{A_c}{A}\right]\right]$$
$$=(p_L+dp_L)(r+dr)d\theta l\left[1-\left(\frac{A_c}{A}+\frac{dA_c}{A}\right)\right]+(F_s+dF_s)\frac{d\theta}{2\pi} \qquad (3.23)$$

where k_0 = coefficient of earth pressure at rest, (-)

l = axial length of cake or tube length, (m)

r = radius, (m)

θ = cylindrical co-ordinate, (radians)

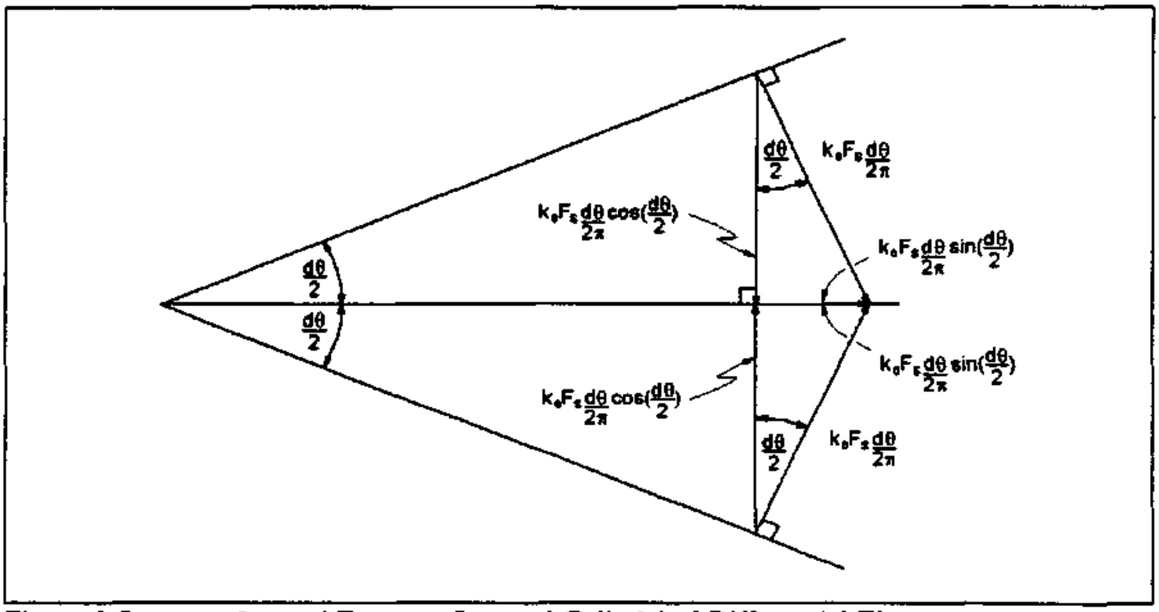

Figure 3.5 Lateral Force on Internal Cylindrical Differential Element of Cake Resolved into Components

Defining the solids compressive pressure as the solids force acting perpendicularly on a superficial area of cake divided by that superficial area of cake, **Equation 3.23** becomes:

$$p_L(rd\theta l)\left[1-\frac{A_c}{A}\right]+p_s(rd\theta l)+2\sin\left(\frac{d\theta}{2}\right)\left[k_0p_s(ldr)+p_L(ldr)\left[1-\frac{A_c}{A}\right]\right]$$
$$=(p_L+dp_L)(r+dr)d\theta l\left[1-\left(\frac{A_c}{A}+\frac{dA_c}{A}\right)\right]+(p_s+dp_s)(r+dr)d\theta l \qquad (3.24)$$

For $p_s \geq p_{sa}$, substituting **Equations 3.10.b** and **Equation 3.20** into **Equation 3.24** and simplifying , leads to:

$$2\sin\left(\frac{d\theta}{2}\right)(k_0+[1-A_0(1-\varepsilon)]p_L)dr$$
$$=d(rp_s)d\theta+[1-A_0(1-\varepsilon)]d(rp_L)d\theta+[A_0d\varepsilon](rp_L)d\theta \qquad (3.25)$$

Now as $d\theta \to 0$, $\sin(d\theta/2) \to d\theta/2$, so **Equation 3.25** reduces to:

$$[1 - A_0(1-\varepsilon)]\frac{dp_L}{dr} + A_0 p_L \frac{d\varepsilon}{dr} + \frac{dp_s}{dr} = (k_0 - 1)\frac{p_s}{r} \qquad (3.26)$$

Equation 3.26 is novel. For $0 \le p_s < p_{sa}$ and $A_0 = 0$ (i.e. point contact), **Equation 3.26** reduces to **Equation 3.27** relating solids compressive to liquid pressure for internal cylindrical filtration, (Rencken, 1992)

$$\frac{dp_L}{dr} + \frac{dp_s}{dr} = (k_0 - 1)\frac{p_s}{r} \qquad (3.27)$$

Equation 3.26 is equally valid for external cylindrical filtration and can be derived in an analogous way as presented above.

3.2.3 Solids Compressive Pressure and liquid Pressure Gradients

In the determination of the solids compressive pressure and liquid pressure gradients through the cake, and hence the overall solids compressive pressure and liquid pressure profiles, the cake can be thought to exist of a number of distinct hypothetical regions. The cake structure and hence the behaviour of the cake properties, is different in each region.

Previously, with the use of point contact models and the assumed functional relationship of permeability and porosity with solids compressive pressure, as described by the correlations given in Section 3.1, the cake could be thought to exist of two distinct hypothetical regions.

The first region extends from the cake surface where $p_s = 0$, to the point within the cake where $p_s = p_{si}$. In this region, the particles of the cake are in point contact with one another in a stable cake structure that consequently exhibits a constant porosity and permeability independent of the solids compressive pressure.

The second region extends from the point where $p_s = p_{si}$, to the filter medium. In this region the cake structure is no longer stable but compressible, and the porosity and permeability are functions of the solids compressive pressure. The particles in the cake are still assumed to be in point contact with one another.

With the introduction of the area contact model, a third hypothetical region of cake can be thought to exist. The second region described above is now restricted from $p_s = p_{si}$ to a point within the cake where the point-area compressive pressure is reached, $p_s = p_{sa}$. The third region of cake now extends from the point-area compressive pressure to the filter medium. In this region the porosity and permeability are functions of the solids compressive pressure as before, except now the particles in the cake are assumed to be in area contact with one another and this area contact is also a function of the solids compressive pressure.

Since the first hypothetical region given by $0 < p_s \leq p_{si}$ represents a *mandatory* region of constant porosity and permeability and hence a constant, fixed cake structure, the crossover from point contact to area contact cannot occur in this region. So generally, $p_{sa} \geq p_{si}$. Equality represents an extreme case where the second hypothetical region is excluded altogether.

A fourth hypothetical region can exist for internal cylindrical compressible filtration. This region is in addition to the three regions which have been discussed. This region will be discussed in more detail in **Section 3.2.3.2** below.

3.2.3.1 Planar Filtration

For planar filtration the expression describing the liquid pressure gradient for a fluid flowing through a porous medium is given as follows (D' Arcy, 1856):

$$\frac{dp_L}{dx} = \frac{\mu_f Q_f}{AK} \tag{3.28}$$

where μ_f = liquid viscosity, (Pa.s)

Q_f = volumetric flow rate of filtrate, (m^3/s)

x = distance from medium, (m)

The corresponding equations relating the solids compressive pressure and liquid pressure, are given by **Equations 3.21** and **Equation 3.22** respectively:

$$\frac{dp_s}{dx} = [A_0(1-\varepsilon)-1]\frac{dp_L}{dx} - A_0 p_L \frac{d\varepsilon}{dx} \qquad p_s > p_{sa} \tag{3.29.a}$$

$$\frac{dp_s}{dx} = -\frac{dp_L}{dx} \qquad 0 \leq p_s \leq p_{sa} \tag{3.29.b}$$

Both the porosity and permeability are functions of the solids compressive pressure and their assumed functional relationships as described by **Equations 3.3** in **Section 3.1**. The solids compressive pressure and liquid pressure gradients can now be determined in each of the hypothetical cake regions.

REGION 1: $0 \le p_s \le p_{si}$

$$\frac{dp_s}{dx} = -\frac{dp_L}{dx} \tag{3.30.a}$$

$$\frac{dp_L}{dx} = \frac{\mu_f Q_f}{AK_i} \tag{3.30.b}$$

REGION 2: $p_{si} < p_s \le p_{sa}$

$$\frac{dp_s}{dx} = -\frac{dp_L}{dx} \tag{3.31.a}$$

$$\frac{dp_L}{dx} = \frac{\mu_f Q_f p_s^{\delta}}{AF} \tag{3.31.b}$$

REGION 3: $p_s > p_{sa}$

From equation (3.3.e) the following is obtained:

$$\frac{d\varepsilon}{dx} = \frac{d\varepsilon}{dp_s}\frac{dp_s}{dx} = -B\beta p_s^{(\beta-1)}\frac{dp_s}{dx} \tag{3.32}$$

Substituting equation (3.32) into equation (3.29.a), yields:

$$\frac{dp_s}{dx} = \frac{\left[A_0 B p_s^{\beta} - 1\right]}{\left[1 - A_0 p_L B\beta p_s^{(\beta-1)}\right]}\frac{dp_L}{dx} \tag{3.33.a}$$

with: $$\frac{dp_L}{dx} = \frac{\mu_f Q_f p_s^{\delta}}{AF} \tag{3.33.b}$$

3.2.3.2 Internal Cylindrical Filtration

For internal cylindrical filtration the expression describing the liquid pressure gradient for a fluid flowing through a porous medium is given as follows (Rencken 1992):

$$\frac{dp_L}{dr} = -\frac{\mu_f Q_f}{2\pi r l K} \tag{3.34}$$

The corresponding equations relating the solids compressive pressure and liquid pressure, are **Equations (3.26)** and **(3.27)** respectively:

$$\frac{dp_s}{dr} = (k_0 - 1)\frac{p_s}{r} - A_0 p_L \frac{d\varepsilon}{dr} - [1 - A_0(1-\varepsilon)]\frac{dp_L}{dr} \qquad p_s \ge p_{sa} \tag{3.35.a}$$

$$\frac{dp_s}{dr} = (k_0 - 1)\frac{p_s}{r} - \frac{dp_L}{dr} \qquad 0 \le p_s < p_{sa} \tag{3.35.b}$$

Again, both the porosity and permeability are functions of the solids compressive pressure and their assumed functional relationships as described by **Equations (3.3)** in **Section 3.1**. The solids compressive pressure and liquid pressure gradients can now be determined in each of the hypothetical cake regions.

REGION 1: $0 \le p_s \le p_{si}$

$$\frac{dp_s}{dr} = (k_0 - 1)\frac{p_s}{r} - \frac{dp_L}{dr} \tag{3.36.a}$$

$$\frac{dp_L}{dr} = -\frac{\mu_f Q_f}{2\pi r l K_i} \tag{3.36.b}$$

REGION 2: $p_{si} < p_s \le p_{sa}$

$$\frac{dp_s}{dr} = (k_0 - 1)\frac{p_s}{r} - \frac{dp_L}{dr} \tag{3.37.a}$$

$$\frac{dp_L}{dr} = -\frac{\mu_f Q_f p_s^{\delta}}{2\pi r l F} \tag{3.37.b}$$

REGION 3: $p_s > p_{sa}$

From equation (3.3.e) the following is obtained:

$$\frac{d\varepsilon}{dr} = \frac{d\varepsilon}{dp_s}\frac{dp_s}{dr} = -B\beta p_s^{(\beta-1)}\frac{dp_s}{dr} \tag{3.38}$$

Substituting equation (3.38) into equation (3.35.a), yields:

$$\frac{dp_s}{dr} = \frac{\left[(k_0 - 1)\frac{p_s}{r} - \left[1 - A_0 B p_s^{\beta}\right]\frac{dp_L}{dr}\right]}{\left[1 - A_0 p_L B\beta p_s^{(\beta-1)}\right]} \tag{3.39.a}$$

with $$\frac{dp_L}{dr} = -\frac{\mu_f Q_f p_s^{\delta}}{2\pi r l F} \tag{3.39.b}$$

REGION 4:

Rencken (1992) observed that for internal cylindrical filtration, as the cake thickness increases, it is possible for the solids compressive pressure at the medium and in cake layers close to the medium to decrease. This effect was found to be a strong function of the coefficient of earth pressure at rest. For $k_0 = 0$, the effect was at its maximum, whilst for $k_0 = 1$, the effect was not observed.

A similar effect can be observed in external cylindrical filtration (Tiller and Yeh, 1985), where the solids compressive pressure at the medium can increase above the applied filtration pressure as the cake thickness increases.

Both the above effects may be explained as follows: The frictional drag force over a layer of cake particles in the plane perpendicular to the direction of filtrate flow is transmitted in the direction of the filtrate flow to the subsequent layer of cake particles together with the accumulated frictional drag force from the previous layers. Since the frictional drag force over the cake layer is equal to the sum of frictional drag forces experienced by each individual particle in that layer, the amount by which the accumulated frictional drag force is incremented across a layer of particles is proportional to the number of particles in that layer. In planar filtration the average number of particles in each cake layer is constant, however for cylindrical filtration the average number of particles in the cake layer is a function of the radial distance through the cake. For thin cake , the ratio of the cake thickness to the radius of the medium is small and the filtration behaves similar to a planar filtration. As the cake thickness increases the filtration becomes more cylindrical in nature and the above effect becomes exaggerated. For internal cylindrical filtration, as the cake thickness increases, each subsequent deposited layer of cake has a reduced area and hence the solids compressive pressure transmitted through to the medium will decrease as the cake thickness increases. The opposite is true for external cylindrical filtration where each newly deposited layer of cake has an increased area, and so the solids compressive pressure at the medium can increase above the applied filtration pressure as the cake thickness increases. For planar filtration where each deposited layer of cake has a constant area and hence the solids compressive pressure at the medium remains essentially constant (note: this is not true for the early stages of the filtration where the effect of the medium resistance predominates).

To a degree, the above effect is compensated for by the radial component of the effective solids force acting laterally on the sides of the cylindrical element of cake. This effective solids force is directly proportional to the value of the coefficient of earth pressure at rest. For external cylindrical filtration the radial component of this force acts outwardly against the direction of filtrate flow and hence against the accumulated frictional drag force. For internal cylindrical filtration it acts inwardly with the accumulated frictional drag force. For planar filtration this force has no component that acts in the direction of filtrate flow.

For both internal and external cylindrical filtration, for $k_0 = 1$ the effect is no longer observed and like planar filtration the solids compressive pressure at the medium remains relatively constant (ignoring the effects of the medium resistance). For $k_0 = 1$, the equation relating solids compressive pressure to liquid pressure **Equation 3.26**, reduces to the corresponding equation for planar filtration **Equation 3.21** and cylindrical filtration begins to imitate planar filtration behaviour in terms of the solids compressive pressure and liquid pressure profiles through the cake. This is because the radial component of the effective solids force acting laterally on the sides of the cylindrical element of cake becomes sufficiently large so as to fully compensate for the radial dependence of the frictional drag force.

For external cylindrical filtration the effect of increasing solids compressive pressure at the medium does not present a problem with regard to modelling the pressure profiles through the cake, however for internal cylindrical filtration this is not the case. As the cake thickness increases, there will come into being a small region of cake extending from the medium that experiences a reduced solids compressive pressure. This region of cake will increase with increasing cake thickness.

It is unlikely that the cake structure in this region, and hence the cake properties such as permeability, porosity and area contact, once stressed at a higher solids compressive pressure profile, will *relax* to a new state as determined by the current solids compressive pressure profile. This would only occur if, for example, the cake particles themselves are deformable and exhibit an elastic behaviour as would be the case for latex beads or very strongly associated particle structures. It is more likely that the porosity and permeability profiles in this region will remain at their minimum values, and the area contact profile at its maximum as determined by the previous higher solids compressive pressure profile. As a result the porosity, permeability and area contact in this region become disassociated from their functional dependence on solids compressive pressure and rather become a *pseudo-function* of the cake radius in this region. The solids compressive pressure and liquid pressure gradients in this region can therefore be calculated as follows:

$$\frac{dp_s}{dr} = (k_0 - 1)\frac{p_s}{r} - A_0 p_L \frac{d\varepsilon(r)}{dr} - [1 - A_0(1 - \varepsilon(r))]\frac{dp_L}{dr} \qquad (3.40.a)$$

$$\frac{dp_L}{dr} = -\frac{\mu_f Q_f}{2\pi r l K(r)} \qquad (3.40.b)$$

If $\varepsilon(r)$ and $K(r)$ are known from a previous calculation history, the correct solids compressive and liquid pressure profiles can be calculated in this region.

3.2.4 Point-Area Compressive Pressure

As a result of various assumptions made in the development of the area contact function it became necessary to define a point-area compressive pressure, which is the solids compressive pressure above which the interparticle contact changes from point contact to area contact.

The point-area compressive pressure as well as the coefficient of area contact are empirically determined parameters. At this stage there is no documented experimental procedure to determine these parameters directly (except for regressing for model parameters from actual filtration data).

However, in context of the area contact model, a minimum feasible point-area compressive pressure can be identified.

3.2.4.1 Planar Filtration

During planar filtration the solids compressive pressure increases towards the medium.

$$\frac{dp_s}{dx} < 0 \tag{3.41}$$

For low solids compressive pressures (and hence relatively high liquid pressures) the term:

$$\left[1 - A_0 p_L B \beta p_s^{(\beta-1)}\right]$$

in **Equation 3.33.a** can become less than zero resulting in solids compressive pressure profiles which are not physically realisable (gradients which are greater than zero). To ensure physically realisable pressure profiles, a lower limit for solids compressive pressure needs to be determined for which **Equation 3.33.a** and the hence area contact model is valid.

By assuming continuity of the solids compressive pressure profiles at the cross-over from point contact to area contact (equating the solids compressive pressure gradients of **Equation 3.31.a** and **Equation 3.33.a** respectively) it can be shown that:

$$\frac{p_s}{p_L} = \beta \tag{3.42}$$

Therefore provided the ratio of the solids compressive pressure to the liquid pressure is greater than the exponential component of the porosity correlation (β), the area contact model will be valid and the pressure profiles physically realisable.

$$\frac{p_s}{p_L} > \beta \tag{3.43}$$

Equation 3.43 is a necessary condition for the area contact model. **Equation 3.43** also provides a means of estimating the lower limit for the point-area compressive pressure. The integrated form of **Equation 3.22** is:

$$p_s + p_L = \Delta P_c \tag{3.44}$$

where ΔP_c = cake pressure drop, (Pa)

Equation 3.44 will still be valid at the cross-over point from point to area contact, so substituting into **Equation 3.43** and solving for the solids compressive pressure gives the lower limit for the point-area compressive pressure. The cake pressure drop increases during the course of the filtration from zero at the start of the filtration, to close to the operating pressure. The lower limit of the point-area compressive pressure is determined using the maximum cake pressure drop which can be approximated by the applied operating pressure.

$$p_{sa} > P_0\left(\frac{\beta}{1+\beta}\right) \tag{3.45}$$

where P_0 = operating pressure, (Pa)

3.2.4.2 Internal Cylindrical Filtration

For internal cylindrical filtration at low solids compressive pressures (and hence relatively high liquid pressures), **Equation 3.39.a** can also describe solids compressive pressure gradients that are not physically realisable.

By assuming continuity of the solids compressive pressure profiles at the cross-over from point contact to area contact (equating the solids compressive pressure gradients of **Equation 3.37.a** and **Equation 3.39.a** respectively) it can be shown that:

$$\frac{p_s}{p_L} = \beta\left[\frac{(k_0 - 1)2\pi lF + \mu_f Q_f p_s^{(\delta-1)}}{\mu_f Q_f p_s^{(\delta-1)}}\right] \tag{3.46}$$

For $k_0 = 1$, **Equation 3.46** reduces to **Equation 3.42**, for $0 \le k_0 < 1$ the term in brackets in **Equation 3.46** will always be positive and therefore less than unity. Therefore provided the ratio of the solids compressive pressure to liquid pressure is greater than the exponential component of the porosity correlation, the area contact model will be valid and the pressure profiles will be physically realisable.

For $k_0 = 1$, **Equation 3.44** is also valid for internal cylindrical filtration and hence **Equation 3.44** can be used to determine the lower limit of the point-area compressive pressure. For $0 \le k_0 < 1$ **Equation 3.44** does not hold, but can be considered a sufficiently accurate approximation so that **Equation 3.45** can be used to estimate the lower limit of the point-area compressive pressure.

3.2.4.3 Multiple Porosity Correlation Data

A problem arises when estimating the lower limit for the point-area compressive pressure when there are multiple porosity correlation parameter sets. Once the condition given by **Equation 3.43** becomes true, and **Equation 3.45** used to estimate the point-area compressive pressure, it must hold for all subsequent porosity correlation data sets, if not, the lower limit of the point-area compressive pressure must be increased beyond the value of the solids compressive pressure where the condition is violated, so that all subsequent correlation data sets are consistent in this manner.

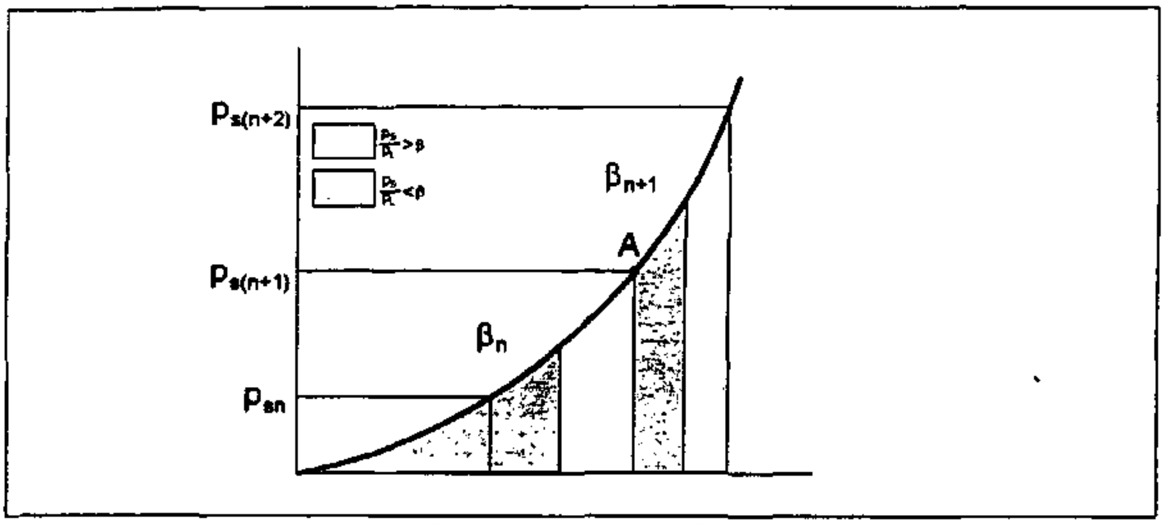

Figure 3.6 **Intersection of Porosity Correlation Data Showing an Inconsistency with Regard to the Applicability of the Area Contact Model**

Consider **Equations 3.3**, if point A in **Figure 3.6** represents the point where $p_s = p_{s(n+1)}$ and is a point where the above mentioned inconsistency holds, then the following expressions must be true:

$$\frac{p_{s(n+1)}}{P_0 - p_{s(n+1)}} > \beta_n \tag{3.47}$$

$$\frac{p_{s(n+1)}}{P_0 - p_{s(n+1)}} < \beta_{n+1} \tag{3.48}$$

Therefore if **Equations 3.47** and **Equations 3.48** hold, the minimum of the point-area compressive pressure will have to increased to $p_{s(n+1)}$. This procedure is repeated for subsequent correlation data sets until the consistency of condition given in **Equation 3.43** can be ensured.

3.3 SOLUTION PROCEDURE

The solution procedure is the application of the constant pressure compressible cake filtration model described above. The overlying procedure is the same for planar and internal cylindrical filtration models and hence will be described here only in the context of internal cylindrical filtration.

3.3.1 Predictive Model

The constant pressure compressible filtration model described above is used in conjunction with the empirical correlation parameters obtained from C-P cell experiments, settling experiments, or regressing on actual filtration data, to predict the filtration performance.

3.3.1.1 Constant Pressure

Equation 3.36, **Equation 3.37** and **Equation 3.39** which describe the solids compressive and liquid pressure gradients, represent a system of two simultaneous ordinary differential equations with initial conditions:

$$p_s(r_2) = 0 \tag{3.49}$$

$$p_L(r_2) = P_0 \tag{3.50}$$

where r_2 = internal radius of the cake, (m)

since the solids compressive pressure is zero at the cake slurry interface and the liquid pressure is equal to the applied operating pressure. For a given cake thickness the solids compressive and liquid pressure profiles can be determined by numerically integrating **Equation 3.36**, **Equation 3.37** and **Equation 3.39** with the initial conditions given by **Equation 3.49** and **Equation 3.50**. These expressions are too complex to obtain a full analytic solution. The cake pressure drop is given by the liquid pressure at the medium.

The pressure drop across the medium for internal cylindrical filtration (Rencken, 1992) is given by:

$$\Delta P_m = \frac{\mu_f Q_f R_m}{2\pi r_1 l} \tag{3.51}$$

where ΔP_m = medium pressure drop, (Pa)

R_m = medium resistance, (m^{-1})

r_1 = internal tube radius, (m)

l = tube length or axial length of cake, (m)

Both the cake and medium pressure drops are functions of the filtrate flow rate and the sum is equal to the applied operating pressure.

$$P_0(Q_f) = \Delta P_m(Q_f) + \Delta P_c(Q_f) \tag{3.52}$$

For a given cake thickness the correct filtrate flow rate can be determined using a numerical technique for solving non-linear equations such that **Equation 3.52** holds.

Once the correct filtrate flow rate has been found and the solids compressive pressure profile has been determined, the porosity profile can be determined using the porosity correlations, **Equations 3.3.d,e,f**. The average porosity of the cake is then given by:

$$\varepsilon_{av} = \frac{2\pi \int_{r_2}^{r_1} \varepsilon r dr}{\pi(r_1^2 - r_2^2)} \tag{3.53}$$

where ε_{av} = average porosity of the cake

Since the solids and the liquid are induvidually incompressible, a mass balance on a volumetric basis is:

$$\begin{bmatrix} \textit{Volume of} \\ \textit{slurry per unit} \\ \textit{medium area} \end{bmatrix} = \begin{bmatrix} \textit{Volume of} \\ \textit{cake per unit} \\ \textit{medium area} \end{bmatrix} + \begin{bmatrix} \textit{Volume of} \\ \textit{filtrate per unit} \\ \textit{medium area} \end{bmatrix}$$

or

$$\frac{\omega_c}{\phi_s} = \frac{\omega_c}{(1 - \varepsilon_{av})} + v \tag{3.54}$$

where ω_c = volume of cake dry solids per unit medium area, (m^3/m^2)

ϕ_s = volume fraction solids in feed sludge, (-)

v = volume filtrate per unit medium area, (m^3/m^2)

The volume of cake dry solids per unit medium area is given by:

$$\omega_c = \frac{(1 - \varepsilon_{av})(r_1^2 - r_2^2)}{2r_1} \tag{3.55}$$

Solving for v using **Equation 3.54** and **Equation 3.55** gives:

$$v = \frac{(1 - \varepsilon_{av} - \phi_s)(r_1^2 - r_2^2)}{\phi_s 2\pi r_1} \tag{3.56}$$

For constant pressure filtration, the filtration time can be calculated as follows:

$$t = \int_0^{V_f} \frac{dV_f}{dQ_f} \tag{3.57}$$

where V_f = volume of filtrate, (m^3)

t = filtration time, (s)

The volume of filtrate can be determined from **Equation 3.56**, and is given by:

$$V_f = \frac{(1 - \varepsilon_{av} - \phi_s)(r_1^2 - r_2^2)\pi l}{\phi_s} \tag{3.58}$$

This procedure is repeated for successive positive increments of the cake thickness and in so doing the cumulative and instantaneous filtration properties are obtained with respect to the filtration time which is determined by numerical integration of **Equation 3.57**.

3.3.1.2 Pseudo Variable Pressure

Often in the case of practical applications of constant pressure filtrations and in the case of the Tubular Filter Press there is a period of time before the actual final filtration pressure is attained. If this *pressurisation* time is significant the solution procedure will have to be modified to account for the pseudo *variable-pressure* portion of the filtration.

For the calculation of cumulative and instantaneous filtration properties during the pseudo variable pressure stage of the filtration, the applied pressure profile must be known with respect to the filtration time. The equations and techniques described above remain valid, except the overall approach of the calculation differs. Instead of calculating the change in filtration time for a given change of cake thickness at a known constant applied pressure, the change in cake thickness is calculated for a known change in filtration time at the average applied pressure over the time interval.

If the cake is incompressible, the thickness of the previously deposited cake layer would remain constant during the subsequent growth of cake at higher applied pressures. This is not the case for compressible cakes. It is assumed that any compression of the previously deposited cake takes place during, and is completed, over the growth of the cake at the higher average pressure over the next time interval. Although cake growth and compression occur simultaneously, the calculation is divided into a growth stage, followed by a compression stage before the next growth stage.

Due to complexities that may arise in calculating the solids compressive and liquid pressure profiles through the cake, the applied pressure profile will be restricted to increase or remain constant with respect to filtration time. If the applied pressure is reduced, as described in **Section 3.2.3.2**, **Equation 3.37** and **Equation 3.39** could not be used to calculate the pressure profiles through the cake.

For the growth stage of the calculation, the cake thickness is incremented and the filtration time calculated as described previously, until the filtration time equals that at the end of the time interval, using the average applied pressure over the interval. The cumulative and instantaneous filtration properties at this time are now known. For the first time interval there is no previously deposited cake so compression during this stage is ignored

Before the growth stage over the next time interval, the cake is compressed at the average applied pressure of the next time interval. Compression involves calculating the cake thickness that results in the same mass dry solids in the cake, but at the higher average applied pressure. This new cake structure represents the basis for the next growth stage. The difference between the calculated filtrate volume and the filtrate volume at the end of the last growth stage represents the fluid that is *squeezed* out of the cake during the compression, this compressed volume is added to the calculated filtrate volume at the end of the following growth stage to give the overall filtrate volume. The overall filtrate flow rate is also adjusted to include the compressed volume over the given time interval, but since compression is assumed to be independent of growth, care must be taken to ensure that filtration time calculations over the growth stage do not include these compressed volumes and rates.

Once the applied filtration pressure reaches the constant operational pressure, the calculation procedure can return to that previously described for constant pressure filtration.

3.3.2 Regressive Model

The constant pressure compressible filtration model is used in addition to data obtained from actual full-scale and/or laboratory-scale filtration tests to determine the empirical correlation parameters that would normally be obtained from C-P cell and/or settling tests.

The regressive filtration model is an optimisation problem with the following objective function which needs to be minimised:

$$a = W_1 \sqrt{\frac{1}{G}\Sigma\left[\frac{(\Delta P_c)_{calc} - (\Delta P_c)_{exp}}{(\Delta P_c)_{exp}} \times 100\right]^2} + W_2 \sqrt{\frac{1}{G}\Sigma\left[\frac{(\varepsilon_{av})_{calc} - (\varepsilon_{av})_{exp}}{(\varepsilon_{av})_{exp}} \times 100\right]^2} \qquad (3.59)$$

where a = objective function result, (-)

W_1 = cake pressure drop weighting factor, (-)

W_2 = average porosity weighting factor, (-)

G = number of experimental observations, (-)

subject to the following constrained parameter set:

$$\{F, \delta, B, \beta, p_{si}, A_0, P_{sa}\} > 0 \tag{3.60}$$

Based on the experimentally observed final cake thickness and final filtrate flow rate and an assumed set of parameters, the solid and liquid pressure profiles through the cake can be calculated and in so doing the cake pressure drop and average porosity of the cake can be calculated. These are compared to the experimentally observed cake pressure drop and average porosity and the value of the objective function is determined for this particular parameter set.

A *direct search* technique is employed to find the set of parameters that minimumises the objective function and results in the best agreement between experimental and calculated cake pressure drops and average cake porosities.

3.4 COEFFICIENT OF EARTH PRESSURE AT REST

From the expressions relating solids compressive pressure to liquid pressure (Section 3.2.2), it is evident that the only difference between planar and internal cylindrical models is the term:

$$(1-k_0)\frac{p_s}{r}$$

where k_0 = the coefficient of earth pressure at rest, (-)

For internal cylindrical filtration, as the internal tube radius increases, the cylindrical nature of the filtration becomes less significant for a given cake thickness, similarly, for a given tube radius, the cylindrical nature of the filtration decreases as the cake thickness decreases. Henry et al. (1976) modelled internal cylindrical compressible filtration inside a vertical porous tube assuming that the cake thickness is small compared to the tube radius and hence that planar theory could be used. He obtained good correlation between experimental results and predictions for a compressible lime neutralised acid mine sludge.

For more compressible cakes such as the cake studied by Rencken (1992), the above effect becomes even less significant since the filtration is essentially controlled by a highly compressed thin layer of filter cake on the filter medium, and the *apparent* cake thickness is considerably less than the actual cake thickness and hence the cylindrical nature of the filtration is even further reduced. This is evident in the results obtained by Rencken (1992), who assumed a value of k_0 = 0.34 and found that over the practical range of internal cake diameters, as applicable to the Tubular Filter Press, the coefficient of earth pressure at rest had no significant effect on the model output when varied from 0 to 1. Provided that the ratio of cake thickness to tube radius is small and/or the cake is compressible, internal cylindrical filtration could probably be more than adequately modelled using planar filtration theory.

Previously, a value for k_0 was assumed, however if one is to utilise the cylindrical filtration model, which incorporates this term, a method for determining the coefficient of earth pressure at rest should be identified and an attempt be made to determine its true value.

The coefficient of earth pressure at rest is a concept originating from the field of soil mechanics and represents the ratio of horizontal to vertical stress when a mass of soil (or filter cake) is subjected to a vertical stress. k_0 can vary between 0 and 1, reported values vary between 0.3 and 1. For a one dimensionally normally compressed soil (and hence a filter cake), k_0 can be approximated by the following expression (Muir-wood 1990):

$$k_0 \approx 1 - \sin\phi \tag{3.61}$$

where ϕ = angle of shearing resistance, (radians)

The angle of shearing resistance can be determined by a number of soil testing procedures such as the triaxial compression test, direct shear test and the unconfined compression test. It was decided that the most practical test, in term of the size of sample required and the sophistication of the test, would be a method known as direct shear testing.

When a sample of filter cake from the tubular filter press was tested, an angle of shearing resistance of 33.4° and hence a coefficient of earth pressure at rest of 0.45 was obtained.

3.5 COMPRESSION - PERMEABILITY CELL TESTS

Compression-Permeability cell (C-P cell) testing is the primary laboratory-scale test to evaluate the empirical parameters in the correlations relating permeability and porosity to solids compressive pressure, **Equations 3.3**. The primary assumption in C-P cell testing is that the local values of porosity and permeability in the filter cake are equal to those of a cake sample in a C-P cell, provided the mechanical pressure in the cell is equal to the local value of solids compressive pressure in the filter cake. There are however a number of problems associated with this primary assumption, e.g. is the overall cake structure in the C-P cell is representative of the local cake structure in the filter cake.

The results obtained from C-P cell testing may be affected by the testing procedure (Lu, Tiller et al. 1970). Firstly, the method by which the cake in the C-P cell is formed, whether by sedimentation or slow filtration, may affect the resulting cake structure and hence the experimental results. A cake formed by sedimentation and then consolidation in a C-P cell may also not be representative of a cake formed during filtration. Secondly, the method under which the test is conducted. Is a single cake used with successive increments of applied pressure (allowing for equilibrium to be attained between each step) or is a different, newly prepared cake used for each loading. If filtrate is allowed to pass through the cake for prolonged periods of time, the migration of small-scale solids could adversely affect permeability values due to *blinding* of the cake and/or filter medium (the permeability would be reduced, however the porosity would remain unaffected).

The cakes studied in C-P cells are static. After each loading, the cake in the cell is allowed to reach equilibrium, the point where cake consolidation ceases. Depending on the applied pressure and the thickness of the cake sample, this can take several hours. Cakes formed during filtration are on the other hand dynamic, as the stress on an element of cake increases continually as successive layers of cake are formed above. Overall filtration times are also normally always less than an hour, as such the equilibrium in filtration cakes may seldom, if ever, be reached for short filtration times.

However the main problem associated with C-P cell tests is that they are not accurate at low values of solids compressive pressure mainly due to the effects of wall friction. Wall friction in C-P cell testing causes non-uniformity in stress distributions within the cake which in turn affect local porosity and permeability values. Previously it was assumed that provided the cake thickness to cell diameter ratio (L/D) was less than 0.6, these effects would be negligible (Grace 1953). However it has been shown that even with (L/D) ratios as low as 0.2, the transmitted pressure in the C-P cell was less than 85 % of the applied pressure (Tiller and Lu 1972), (Lu, Tiller et al. 1970) even for highly compressible cakes (in the absence of wall friction the transmitted pressure would equal the applied pressure). Although the effects of wall friction are reduced for smaller (L/D) ratios it has been found that the repeatability of test results decreases for decreasing (L/D) ratios, particularly if (L/D) < 0.2 (Lu, Tiller et al. 1970).

Therefore unless an attempt is made to correct C-P cell data for the effects of side wall friction, the data could be in error and due care must be taken in all design since there can be no guarantee that the cake structure obtained will be representative of cakes formed during actual filtration. Correction methods have been developed by Tiller and Haynes et al. (1972) who suggested relations for correcting calculated values of porosity and specific cake resistance in terms of the ratio of applied to transmitted pressure. However the method proposed by Shirato et al. (1968) where the compressive pressure is corrected after statistically comparing calculated and experimentally determined transmitted pressures is probably more general, since adhesion between particles and the cell walls are taken into account.

3.5.1 Determining Correlation Data

The permeability of the cake in the C-P cell at a particular loading can be determined from D' Arcy's law (D' Arcy, 1856) and is given by:

$$K = \frac{\mu_f Q_f \Delta t_c}{A_{cell} \Delta p_c} \tag{3.62}$$

where Δt_c = thickness of cake in C-P cell, (m)

A_{cell} = area of cake in C-P cell, (m^2)

Δp_c = hydrostatic pressure drop across cake in C-P cell, (Pa)

It is assumed that the combined resistance of the porous plates and filter papers in the C-P cell is negligible.

The final porosity of the cake in the C-P cell at the end of the test can be determined from the final moisture content of the cake:

$$\varepsilon = \frac{\frac{m}{\rho_l}}{\frac{m}{\rho_l} + \frac{(1-m)}{\rho_s}} \tag{3.63}$$

where m = mass fraction of moisture in cake, (-)

ρ_l = liquid density, (kg/m^3)

3.5.2 Approximate Correction for Side Wall Friction

The following method for the correction of side wall friction in C-P cell testing was developed by Shirato et al. (1968).

Assuming that the vertical pressure is uniformly distributed across the cell and that there is a constant cohesive force at the wall, a force balance over a differential element of cake shown in **Figure 3.7** may be written as follows:

$$\frac{\pi D^2}{4}[p_V-(p_V+dp_V)]=(k_0 f p_V+c)\pi D\cdot dz \tag{3.64}$$

where f = coefficient of friction, (-)

D = inside diameter of the cell, (m)

z = distance from the top of the cake, (m)

c = cohesive force between the side wall and the compressed cake, (Pa)

p_V = vertical solids pressure in the cake, (Pa)

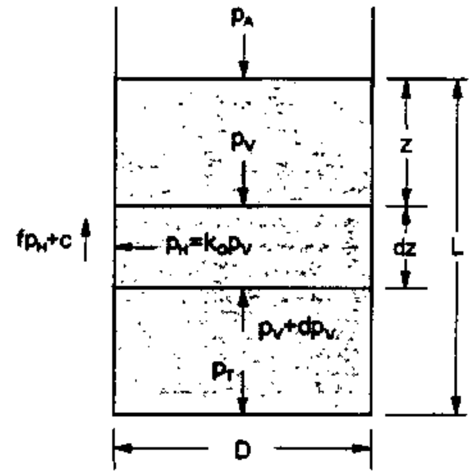

Figure 3.7 **Force Balance on Differential Element of Cake Inside the Compression-Permeability Cell**

Integrating **Equation 3.64**, assuming that $k_0 f$ and c are constant, from the top of the cake where the vertical pressure is equal to the applied pressure, to some point within the cake yields:

$$p_V=\frac{1}{k_0 f}\left[\frac{k_0 f p_A+c}{\exp(4k_0 fz/D)}-c\right] \tag{3.65}$$

where p_A = the applied pressure at the top of the cake, (Pa)

Or alternatively, integrating **Equation 3.64** from the top of the cake to the bottom of the cake where the vertical pressure will be equal to the transmitted pressure, yields:

$$p_T=\frac{1}{k_0 f}\left[\frac{k_0 f p_A+c}{\exp(4k_0 fL/D)}-c\right] \tag{3.66}$$

where p_T = transmitted pressure through the cake, (Pa)

L = compressed equilibrium thickness of the cake, (m)

The average compressive pressure through the cake is defined as:

$$p_S = \frac{1}{L}\int_0^L p_V \cdot dz \tag{3.67}$$

Substituting **Equation 3.65** into **Equation 3.67** yields:

$$p_S = \frac{p_A + (c/k_0 f)}{4k_0 fL/D}\{1 - \exp(-4k_0 fL/D)\} - \frac{c}{k_0 f} \tag{3.68}$$

Due to the variable stress within the cake, and hence variable porosity and permeability, the data obtained from the C-P cell test would be better correlated with the solids compressive pressure obtained from **Equation 3.68** as opposed to the applied pressure as in conventional C-P cell testing.

The terms $k_0 f$ and c may be obtained by statistically comparing experimentally measured transmitted pressures to those calculated from **Equation 3.66** and determining which values of $k_0 f$ and c result in a minimum root-mean-squared deviation as given by the expression:

$$\sigma = \sqrt{\frac{1}{G}\Sigma\left[\frac{(p_T)_{calc} - (p_T)_{exp}}{(p_T)_{exp}} \times 100\right]^2} \tag{3.69}$$

where σ = root-mean-squared deviation, (-)

G = the number of experimental observations, (-)

A new C-P cell has been constructed that will enable the measurement of thetransmitted pressure in the C-P cell (see **Section 4.2.1**) and hence the data can be corrected for the effects of wall friction.

3.6 SETTLING TESTS

The following techniques for the determination of permeability and porosity data at low solids compressive pressures was developed by Shirato, Murase et al. (1983) and are the same techniques as documented by Rencken (1992).

3.6.1 Porosity Correlation Data

When a suspension containing a dry solids volume per unit area, ω_0, settles in a cylinder, the final equilibrium height of the sediment is denoted by H_∞. When the initial volume of solids is increased by $d\omega_0$, the final height of the sediment will be given by $(H_\infty + dH_\infty)$. The porosity variation of the suspension AB is identical to that of the suspension CD in **Figure 3.8** As a result the total solids volume in sediment CE can be represented by:

$$\omega_0 + d\omega_0 = \omega_0 + dH_\infty(1-\varepsilon) \tag{3.70}$$

where H_∞ = final height of sediment, (m)

ω_0 = total volume of dry solids per unit cross sectional area, (m^2)

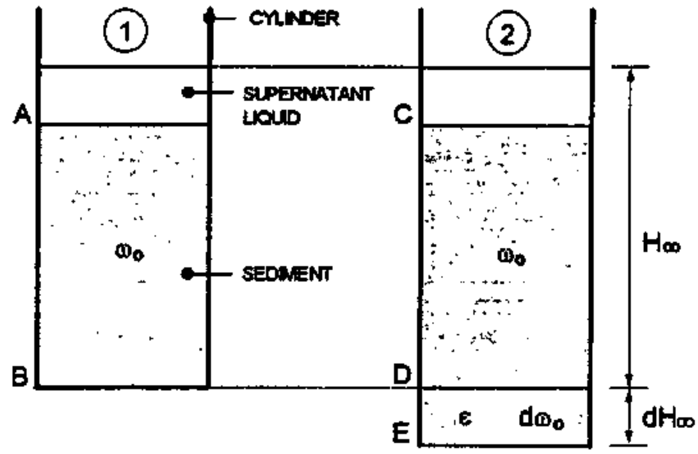

Figure 3.8 **Relationship Between Height of Sediment and Volume of Dry Solids per Unit Area**

solving for the porosity in **Equation 3.70**:

$$\varepsilon = 1 - \frac{d\omega_0}{dH_\infty} \tag{3.71}$$

ω_0 is related to the solids compressive pressure by the relation:

$$p_s = (\rho_s - \rho_l)g\omega_0 \tag{3.72}$$

where g = constant of gravitational acceleration, (m/s^2)

Shirato, Murase et al. (1983) found that on the basis of experimental data, H_∞ could be represented in terms of ω_0 by the following equation:

$$H_\infty = a\omega_0^b \tag{3.73}$$

where a, b = empirical constants, (-)

substituting **Equation 3.73** into **Equation 3.71** and eliminating ω_0 by means of **Equation 3.72**, one obtains **Equation 3.74**:

$$(1-\varepsilon) = Bp_s^{\beta} \tag{3.74}$$

where $$B = \frac{1}{ab[(\rho_s - \rho_l)g]^{(1-b)}} \tag{3.75}$$

and $$\beta = (1-b) \tag{3.76}$$

The empirical constants a and b may be obtained from linear regression using the experimental results of $\ln H_{\infty}$ and $\ln \omega_0$ and the linearised form of **Equation 3.73**:

$$\ln H_{\infty} = \ln a + b \ln \omega_0 \tag{3.77}$$

3.6.2 Permeability Correlation Data

Michaels and Bolger (1962) investigated the sedimentation behaviour of a flocculated suspensions of kaolin. For their sedimentation model they assumed that for flocculated suspensions, the basic flow units or settling entities are small clusters of particles or flocs. For gravity settling the flocs group into clusters called particle or floc aggregates. Michaels and Bolger (1962) found that the floc aggregates determine the sedimentation behaviour of flocculated suspensions.

Shirato, Murase et al. (1983) confirmed the results of Michaels and Bolger (1962). By using zinc oxide, Mitsukuri Gairome clay and ferric oxide slurries, they showed that sedimentation behaviour may be classified into three general regions according to the initial porosity of the suspension, ε_{in}, as shown in **Figure 3.9**.

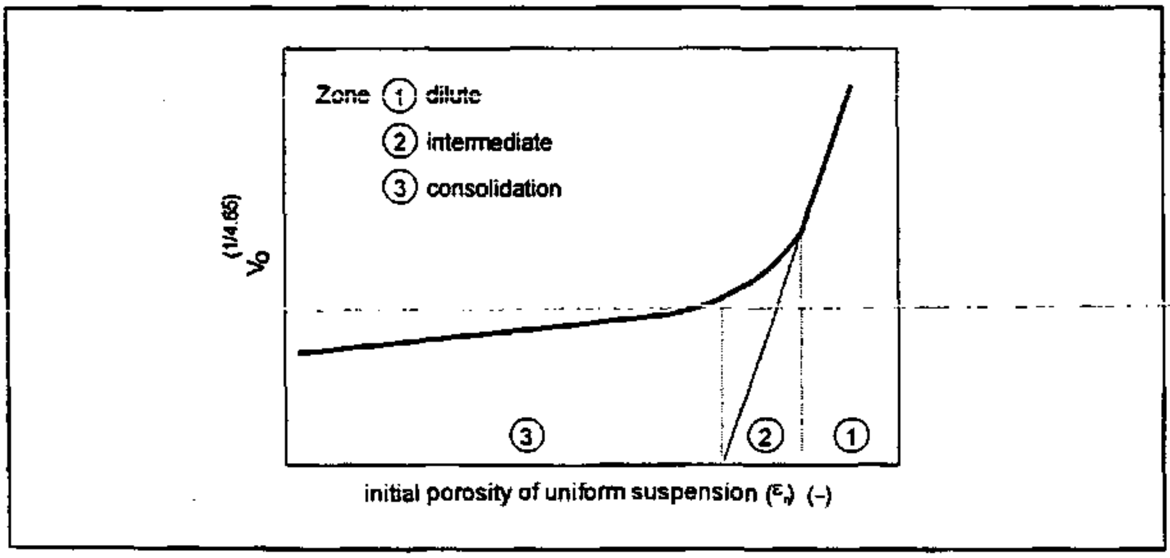

Figure 3.9 **Settling Regimes for a Slurry**

In the **dilute** concentration region, *free* or *hindered* settling of the individual particles or aggregates may occur. In the **intermediate** concentration region, sedimentation behaviour becomes unstable due to the partial collapse of particle aggregates. In the higher **consolidation** concentration region the supernatant liquid-sediment surface interface subsides slowly due to the *consolidation* or *compression* of the sediment. In order to describe the sedimentation behaviour in the dilute concentration region, Michaels and Bolger (1962) modified the equation developed by Richardson and Zaki (1954) for the hindered settling of uniform, spherical particles. According to the modified equation, for dilute suspensions, the initial settling velocity of the surface of the sediment can be related to the initial porosity by:

$$V_0^{\frac{1}{4.65}} = \left[\frac{g d_f^2(\rho_s - \rho_l)}{18\mu_f(1-\varepsilon_{if})^{3.65}}\right]^{\frac{1}{4.65}} (\varepsilon_{in} - \varepsilon_{if}) \tag{3.78}$$

where d_f = mean diameter of particle aggregates, (m)

V_0 = initial settling velocity of surface of sediment, (m/s)

ε_{if} = internal porosity of particle aggregates, (-)

ε_{in} = initial porosity of suspension

As shown in **Figure 3.9**, a straight line is obtained in the dilute region in accordance with **Equation 3.78**.

The technique for measuring porosity and permeability at low solids compressive pressure is only applicable in the high concentration region where sedimentation takes place due to consolidation of the sediment (Shirato et al. 1983). For batch sedimentation of a concentrated suspension inside a cylinder, the liquid pressure and solids compressive pressure increase towards the bottom of the cylinder. When a differential element of suspension settles due to consolidation, liquid has to flow through the element due to the liquid pressure gradient across the element as shown on **Figure 3.10**. The liquid pressure gradient across the element is caused by the weight of the particles lying above the element. The D' Arcy equation (D' Arcy, 1856) may be used to describe the liquid flow through the element:

$$u_l = -\frac{K}{\mu_f}\frac{\delta p_L}{\delta y} \tag{3.79}$$

where u_l = apparent liquid velocity relative to solids, (m/s)

y = distance measured from bottom of cylinder, (m)

But

$$d\omega = (1-\varepsilon)dy \tag{3.80}$$

where ω = volume of dry solids per unit cross-sectional area, measured from bottom of cylinder, (m^3/m^2)

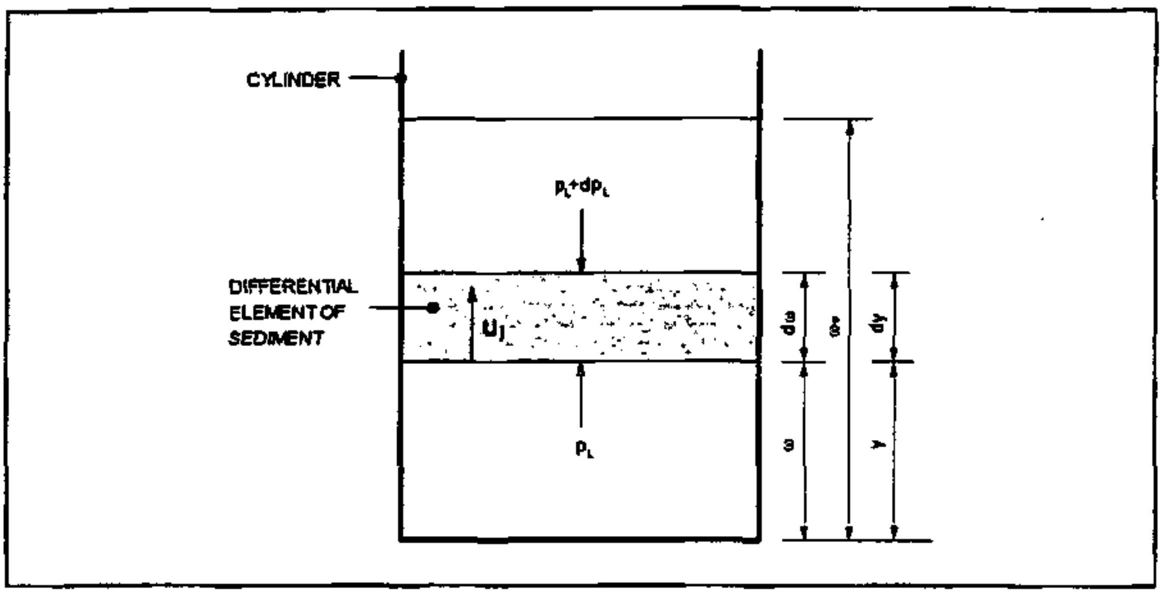

Figure 3.10 **Up-Flow of Liquid Through a Differential Element of Slurry Due to Liquid Pressure Gradient**

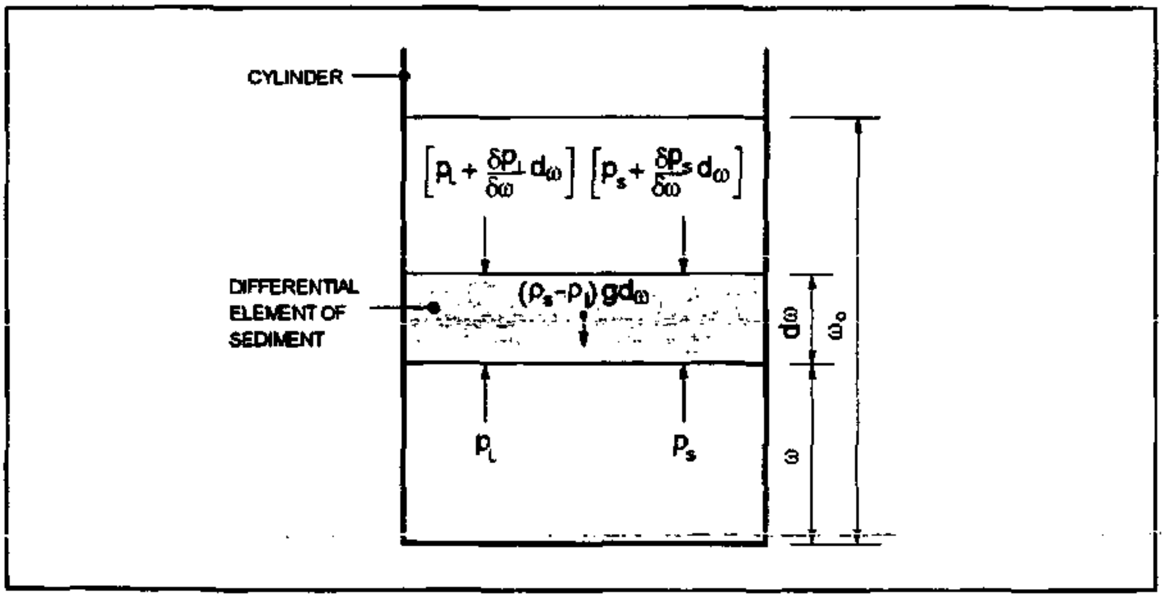

Figure 3.11 **Forces Exerted on a Differential Element of Slurry**

Therefore **Equation 3.79** can be written in terms of ω, as follows:

$$u_l = -\frac{K(1-\varepsilon)}{\mu_f}\frac{\delta p_L}{\delta\omega} \tag{3.81}$$

The forces acting on a differential element, $\delta\omega$, are shown in **Figure 3.11**. The following equation can be derived from Newton's second law of motion (assuming accelerative and inertial effects to be negligible):

$$\frac{\delta p_L}{\delta\omega}+\frac{\delta p_s}{\delta\omega}=-(\rho_s-\rho_l)g \qquad (3.82)$$

Wall friction is assumed to be negligible.

Since the solids compressive pressure can be assumed to be constant throughout the height of the cylinder at the beginning of a settling test, when the suspension is uniform (Shirato et al. 1983), the initial liquid pressure gradient can be obtained from **Equation 3.82**.

$$\left(\frac{\delta p_L}{\delta\omega}\right)_{t=0}=-(\rho_s-\rho_L)g \qquad (3.83)$$

At the start of the test, u_l, in **Equation 3.81** can be considered as a constant which is equal to the initial settling velocity of the sediment surface, V_0. Combing **Equation 3.81** and **Equation 3.83**:

$$K=\frac{V_0\mu_f}{(\rho_s-\rho_l)(1-e_{in})g} \qquad (3.84)$$

Both Shirato et al. (1983) and Rencken (1992) found the initial height of the uniform suspension in the cylinder, had no effect on the initial settling velocity of the suspension.

3.7 CAKE EQUILIBRIUM EFFECTS

A primary assumption in filtration theory is that equilibrium porosities are attained instantaneously with changing solids compressive pressure, i.e. the cake is always in equilibrium. However in actual filtration this is not the case, during the initial stages of filtration, cake growth is rapid and subsequent cake structures are not in equilibrium, only as the filtrate flow rate and hence the cake growth rate decreases, will the cake begin to obtain equilibrium as it has time to consolidate under the solids compressive pressure.

This can have a number of effects with regard to evaluating the performance of the filtration plants. For instance in terms of the filtration theory, the feed solids concentration should have no effect on the cake equilibrium, however in reality as feed solids concentration increases so does the initial cake growth rate and as such the resultant initial cake structure will be further from the equilibrium cake structure. Cakes formed from dilute feed solids concentrations are more likely to be in equilibrium because the cake structure is exposed to greater periods of solids compressive stress per unit mass solids deposited as the increased amount of filtrate (per unit solids deposited) flows through the cake structure.

Filtration can therefore be regarded as being divided into two conceptual phases. The *cake growth phase*, whilst the filtrate flow rates are still relatively high. The filtrate flow rate decreases rapidly and then levels off, cake growth becomes relatively insignificant and the filtration can now be regarded to be in a *cake consolidation phase* as the filtrate flows through the cake structure with little cake growth. This is evident in planar filtration tests where the theoretical dimensionless solids compressive pressure profile is constant and hence the porosity profile and average cake porosity should be constant, however filtration tests show that the cake solids concentrations are initially below the predicted value and increase slowly becoming relatively constant in the region of the predicted value.

A cake structure that is not in equilibrium will be more porous and therefore be more permeable than an equal mass of cake at equilibrium. Non-equilibrium cake structures will therefore deposit a greater mass of solids before a predetermined limiting flux is attained than equilibrium cake structures. As a result, if the operation of a plant is to terminate at a certain final flux, the higher the feed solids concentration, the larger the amount of solids that will be deposited before the filtration cycle terminates.

If the filtration is not allowed to operate in the consolidation phase for a sufficiently long period of time, the average cake solids concentration could be lower than expected. The unconsolidated cake structure may also be structurally weaker than an equilibrium cake structure, this could have adverse effects on the cake recovery for plants such as the Tubular Filter Press where the cake is hydraulically removed by the feed sludge. Obtaining an optimum dry solids production rate is then a trade off between operating in the cake consolidation phase for long enough to ensure a good recovery without adversely extending the filtration cycle time, and hence lowering production rates.

3.8 TUBULAR FILTER PRESS

In this section some aspects of the modelling process specific to the Tubular Filter Press are discussed.

3.8.1 Dynamic Dead-End Internal Cylindrical Filtration

During the modelling of the dead-end filtration cycle of the Tubular Filter Press the *dynamic* properties of the filter tube have been ignored, the entire tube has been treated as being uniform with respect to filtration pressure and cake thickness and dynamic effects such as the shearing of the filter cake have been ignored. In reality both the local filtration pressure and cake thickness will vary down the length of the tube. The overall filtration rate will be determined by the average of the local filtration rates along the tube, which depend on the local filtration pressure and cake thickness. Hence the significance of these factors should be investigated.

3.8.1.1 Dead-End Shear Model

Shearing of the filter cake has been modelled by Rencken (1992) during the cleaning cycle and during the cross-flow filtration mode of the Tubular Filter Press, but not during the dead-end filtration mode. Shearing of the cake within the tube will be a function of the local axial flow velocity and internal cake radius, both of which are dependent on the axial pressure profile within the tube. The local axial flow velocity will be a maximum at the top of the tube and zero at the bottom, hence if any shearing takes place, it will be at the top of the tube. However, unlike shearing during cross-flow mode, any sheared material will be redeposited further down the tube during dead-end mode, if this effect are significant it could lead to blockages further down the tube.

Intuitively one would expect that shearing effects will be negligible since the filtration rate and hence the axial flow rate will decrease sharply as soon as a thin layer of cake is deposited, during the initial stage the tube internal cross-sectional area available for flow is still large and hence the axial flow velocity will be low and the cake will be a thin highly consolidated layer that will be resistant to shearing. By the time the internal cross sectional area has decreased significantly, the axial flow rate will no longer be significant. If any shearing occurs it will probably be restricted to the tube inlet and its effects will be negligible. Shearing effects hence may only be of significance for very long tubes with a small internal diameters. For the purposes of this study, these effects have been ignored

3.8.1.2 Axial Pressure Profiles

The local filtration pressure in the tube will equal the overall applied pressure plus a hydrostatic contribution, minus a pressure drop associated with the flow of the slurry down the tube. The local hydrostatic contribution will depend on the orientation of the tubes, the density of the slurry and the position relative to the tube inlet. The pressure head loss due to fluid flow will depend on the internal cake radius, the axial flow profile within the tube and the physical properties of the slurry such as the viscosity. These factors will only be of significance for relatively low applied pressures, long filter tubes and dense highly viscous slurries. At this stage it appears justifiable to ignore the head loss due to the flow of the slurry down the tube and to treat the filtration pressure as constant down the tube and equal to the filtration pressure plus the average of the hydrostatic pressure along the tube.

3.8.1.3 Axial Feed Solids Concentration Profiles

For long filtration times, low applied pressures and low feed solids concentrations, the filtrate flow rate may be sufficiently low such that the sludge settling rate is greater than the axial flow rate in the tube. As a result, non-uniform feed solids concentration profiles may occur in the tubes. However, under normal operation of the tubular filter press axial feed solids concentration profiles will be assumed constant.

3.8.2 Cake Recovery

Rencken (1992) studied the cake recovery due to shearing of the cleaning fluid prior to the action of rollers and due to the combined action of the rollers and hydraulic conveyance of the flakes of cake during the cake removal cycle of the Tubular Filter Press. Rencken (1992) developed a shear model, and was able to predict the increase in average cake porosity and internal cake radius as the loosely consolidated layers of cake were sheared off by the cleaning fluid, but was unable to account for the combined action of the rollers and hydraulic conveyance.

For the vertical orientation of the Tubular Filter Press, the cake is not exposed to the shearing action of the cleaning fluid *in situ*, prior to being removed by the rollers, as is the case of the Horizontal Tubular Filter Press studied by Rencken (1992). The mechanism of cake removal for the Vertical Tubular Filter Press differs from that of the Horizontal Tubular Filter Press. For the Vertical Tubular Filter Press the cake is removed by the combined action of the cylindrical shape of tube curtain collapsing when not pressurised in dead-end mode as during the filtration cycle, and series of flush cycles. The rollers are only used during the final flush cycle to remove any remaining cake (which is usually negligible). As such, cake losses are mainly due to hydraulic conveyance only and the shear model developed by Rencken (1992) is not applicable.

3.8.2.1 Recovery Due to the Action of Rollers or Hydraulic Conveyance

Rencken (1992) observed that the cake recovery during the action of the rollers and hydraulic conveyance was a complex function of the filtration pressure, filtration time, the length of the tube through which the cake was conveyed and flow rate used to convey the cake. Filtration pressure had a significant effect on cake recovery, the higher the filtration pressure, the higher the recovery. The filtration time (or mass of dry cake deposited) also had a significant effect, the recovery increased with filtration time, but after a certain *threshold* value remained relatively constant. The tube length and flush flow rate where found to be insignificant.

3.8.2.2 Recovery Function

It seems impossible to develop a theoretical model that will accurately predict cake losses during the cleaning cycle of the Tubular Filter Press. It is proposed that it may however be possible to develop an empirical model describing cake recovery as a function of the principle factors, namely the filtration pressure and the mass of dry cake deposited. The mass of dry cake deposited is in turn dependent on the filtration pressure, feed solids concentration and the filtration time. The recovery function could then be modelled on these three parameters.

Some work was done in this regard, but was abandoned when the empirical expressions got too complex to be of practical use.

4 Experimental Procedures and Techniques

A number of laboratory tests were performed to be able to characterise the sludge, and be able to quantify differences in the performance of the Tubular Filter Press under different operating conditions. This section details the methods and techniques used, and describes the operation of the pilot plants.

4.1 SLUDGE

The sludge used for this investigation was obtained from the clarifiers and filter backwash water at the Wiggins Waterworks. The water is treated using polymeric coagulants and the flocculated particles are allowed to settle in Degremont type *pulsator clarifiers* which produces a sludge concentration of 0,2 to 0,5 % (m/m). Rapid gravity sand filtration captures any further solids present in the water after clarification. The backwash water emanating from these filters is combined with the sludge from the clarifiers and is thickened to a solids concentration of 1,0 to 3,0 % (m/m). The sludge used for the laboratory-scale and pilot-plant tests was sampled from the thickened sludge storage tank at the works.

4.1.1 Determination of Solids Density

The measurement of the solids density is critical in the laboratory tests for obtaining an accurate estimate of the porosity of the compressed cake. Two methods were considered for this determination.

4.1.1.1 Evaporation Method

In order to estimate the density of the solids in a sample of sludge, it is assumed that the solids comprise discrete particles separated only by a film of water. A sample of wet sludge is taken and the volume of the sample is recorded. The sample is placed in a drying oven set at a temperature of 105^{0} Celsius for several hours to allow the water to evaporate. Once the sample has reached constant mass, it is cooled in a dessicator, and the sample of solids remaining is accurately weighed. The volume of the solids is calculated by subtracting the volume of water that was evaporated (assuming the density of water to be 1 g/ml), and hence the density of the solids is determined. Therefore by evaporating the water from between the particles and measuring the mass of the solids for a specific volume of sludge, the volume of the solids is calculated and the density can be obtained.

4.1.1.2 Specific Gravity Bottle Method

Density bottles are small thick walled glass bottles which can be calibrated to determine the volume and mass relationship of a substance. The bottles should be cleaned and rinsed with alcohol or acetone prior to calibration. Distilled water is used to determine the exact volume of the bottle at a specific temperature, as the density of water at temperatures between 20 and 30^0 Celsius is well documented.

The sample of sludge should be dried in an oven at 105^0 Celsius for several hours. The solids should be crushed and sieved to a very fine powder and a homogenous sample is used for the density determination. The density bottle should be cleaned and rinsed again before testing. The mass of the empty bottle is recorded and then an amount of the powdered solids are added to the bottle and reweighed. The mass of the solids in the bottle is calculated.

Distilled water is then carefully added to the density bottle, and the bottle is placed in a water bath at a desired working temperature, between 20 and 26^0 Celsius. The bottle is left to stand in the water bath for at least an hour to allow the solids to completely wet and to release any air bubbles entrapped in the powder. The bottle is removed from the water bath, dried and the combined mass of the bottle, powder and the water is recorded. The difference in the mass of the combined substances and the mass of the bottle with solid powder only, is used to calculate the exact volume of water added to the bottle. The volume of the solid powder in the bottle is therefore the difference between the calibrated volume of the bottle and the volume of water added. The density of the solids can then be determined.

4.2 COMPRESSION - PERMEABILITY CELL TESTS

Schematics of the Compression-Permeability cells (C-P cell) used during this study are shown in **Figure 4.1**. The C-P cell (Cell A) was constructed based on the design given by Rowe et al. (1966) and used by Rencken (1992) and a second cell was manufactured to include a *floating bottom* that enables the transmitted pressure through the cake to be measured (Cell B).

The C-P cells are constructed out of 304 stainless steel which is resistant to corrosion. The base and cover of *Cell A* are attached to the cell body by six pins that screw into the base through the cover, the top diaphragm serves as a gasket, and the bottom is sealed with an o-ring. The base and cover of *Cell B* are bolted to the flanges of the cell body at six positions, with the top rubber diaphragm and bottom rubber membrane serving as gaskets.

A uniform load is applied to the sample by increasing the air pressure in the air filled chamber. A thin plastic plate is placed on top of the top porous plate to prevent the damage to the rubber diaphragm. A grease seal enables the permeate inlet stem to move *frictionlessly* through the cover whilst still maintaining the air pressure in the air filled chamber. The grease seal also serves the secondary function of ensuring the inlet stem remains true. The rubber diaphragm is clamped tightly between two plates which screw onto the inlet stem, an o-ring at the base of the thread completes the seal. Consolidation of the sample is measured at the centre by a dial gauge at the top of the inlet stem. The dial gauge is rigidly supported by an adjustable arm attached to the edge of the cover so that it is free from deflection due to distortion of the cover with changing pressure. A top and bottom filter paper separate the cake sample from the two porous discs.

For *Cell B*, the outlet stem is attached to the bottom rubber membrane by two plates in the same manner as described for the top diaphragm, providing a false bottom that *floats* on a fluid filled chamber that has been machined out of the base of the cell. Any force on the bottom porous plate will be transmitted to the fluid, the associated increase in pressure is measured with a fluid filled pressure gauge. The fluid is incompressible so no movement of the bottom plate should occur. Two o-rings complete the seal in the fluid filled chamber where the outlet stem passes through the base. Once the base of the cell is assembled, air is removed from the fluid filled chamber by inverting the cell and flushing with fluid at a high flow rate, any trapped air is entrained out of the chamber. To facilitate the air clearing, all corners are machined round where possible and an air well is machined into the base at the fluid outlet. Any trapped air will contract as the fluid pressure increases, the associated movement of the bottom plate, if excessive, will disrupt the cake structure which is undesirable. A dial gauge attached to the permeate outlet in the same manner as described above will measure any movement of the bottom plate.

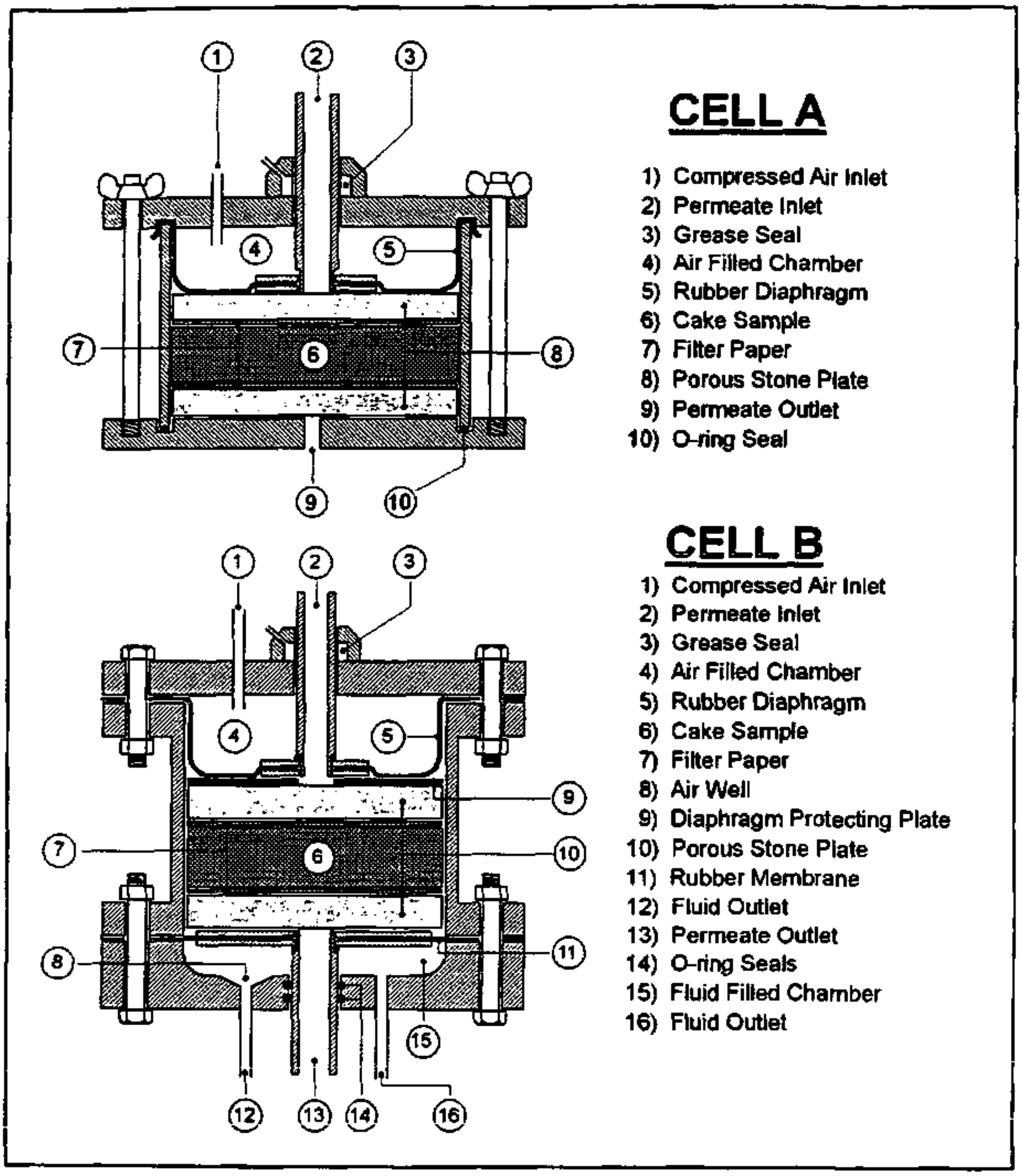

Figure 4.1 **Schematic of Compression-Permeability Cell A and Cell B**

A process schematic for the C-P cells is shown in **Figure 4.2**. Filtrate is stored in a header tank and allowed to flow down the inlet stem through the cake and then through a micropipette to enable the measurement of the volumetric flow rate of the permeate. The level of filtrate in the header tank and hence the hydrostatic head is fixed by an overflow pipe. The cake is compressed by the pressure exerted on the top porous plate by the rubber diaphragm. The air pressure is controlled with a pressure control valve. For *Cell B*, the transmitted pressure is measured by the fluid filled pressure gauge attached to the fluid filled chamber, and fluid inlet line is attached to the mains water.

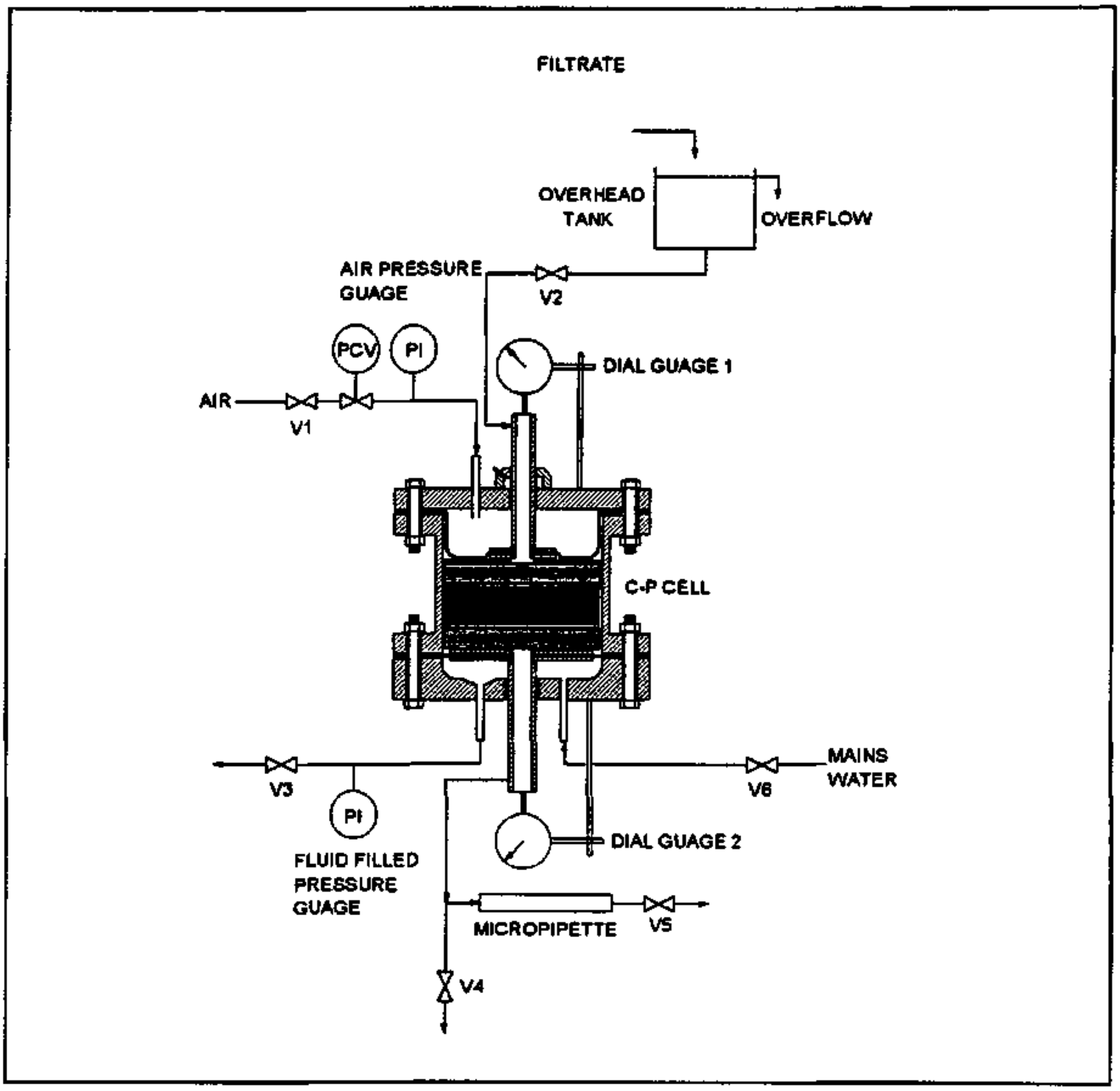

Figure 4.2 **Process Schematic for the Compression-Permeability Cell Equipment**

4.2.1 Assembly of the Compression-Permeability Cell

Cell A - the cell body was placed on the base after the o-ring had been checked for correct alignment, and the locating recess for the cell walls had been filled with water.

Cell B - the base of the cell was bolted tightly to the cell body using the bottom rubber membrane as a gasket, which in turn was tightly clamped between the two bottom plates screwed onto the permeate outlet stem. The assembled base was then inverted and valves V3 and V6 opened. The mains water was then turned on at a low flowrate and the bottom chamber was filled with water and most of the air expelled. The water flowrate was then increased and any remaining air flushed out of the chamber. Valves V6 and then V3 were closed in quick succession so that the fluid pressure inside the chamber was at ambient pressure and not the mains water pressure. The assembled base was then placed upright and the bottom dial gauge attached to the permeate outlet stem.

- The cell was then filled with filtrate and valves V4 and V5 opened so that the air in the feed line to the micro-pippette and the pipette by-pass line was expelled. Valves V4 and V6 where then closed.
- A porous disc which had been soaked in water to remove air and debris from the pores was then allowed to settle into the filtrate onto the bottom of the cell. A glass micro-fibre filter paper which had been cut to a diameter slightly larger than the diameter of the cell and soaked in water to remove any air was carefully placed onto the surface of the filtrate in the cell ensuring that no air was trapped underneath. The filter paper was then placed on top of the bottom porous disc ensuring that a proper seal was formed between the filter paper and the sides of the cell.
- Valve V4 was then opened and the filtrate in the cell allowed to drain to just above the bottom filter paper. The cell was then partially filled with a concentrated slurry, approximately 86 g/l, leaving sufficient room for the top porous disc. Filtrate was then poured gently on top of the slurry through a porous disc held just above the slurry so as not to disturb the slurry. The remainder of the cell was filled with clear filtrate in this manner. Valve V4 was then opened again to ensure that the bottom seal had not been compromised and the permeate was clear, valve V4 was then closed again.
- A second filter paper cut to the same diameter of the cell, which had been soaked in water was then placed on the surface of the filtrate ensuring no air bubbles where trapped underneath and allowed to settle onto the slurry. The second porous disc was then placed in the cell and allowed to settle onto the filter paper. The cell was then left overnight to allow the slurry to gently consolidate under the weight of the top porous plate.
- Valve V2 was then opened and the supply line to the permeate inlet stem cleared of any air. The plastic diaphragm protecting plate was placed on top of the top porous plate. The entire top assembly was then carefully lowered on top of the cell whilst filtrate flowed out of the permeate inlet stem, ensuring that no air was trapped between the diaphragm and the top porous disc.

- The top was then tightly bolted to the cell using the rubber diaphragm as the gasket. The header tank was then filled with filtrate, the level in the header tank was fixed, and valve V4 was opened.

4.2.2 Experimental Procedure

The pressure in the diaphragm was slowly increased in small increments to the first applied pressure of approximately 50 kPa so as not to disturb the cake in the cell, this was important as the cake at this stage was not consolidated and very unstable. The cell was then left for several hours for the cake to consolidate. The top dial gauge was then attached to the top of the permeate inlet stem. Consolidation equilibrium was determined when movement of the permeate inlet stem as observed on the top dial gauge had ceased. In order to measure the permeate flow rate valve V5 was opened and the level in the micro-pipette allowed to drop, valve V4 was then closed and the permeate timed over a measured change in volume in the micro-pipette. Valve V5 was then closed and valve V4 opened. The reading on the top dial gauge was then observed. For *Cell B*, the applied and transmitted pressures and the readings on the bottom dial gauge were also observed. The applied pressure was then increased in increments of approximately 50 kPa, allowing consolidation equilibrium to be reached at each pressure, up to a final pressure of 450 kPa and further readings obtained in the same manner.

At the end of the run, the applied pressure was cut-off and all valves closed. The cell was then carefully disassembled and the highly consolidated cake removed from the cell body and weighed. Any irregularities in the cake, if present, were observed. The cake was then placed in an oven at 120^0 Celsius for at least 5 hours and re-weighed to determine the mass of solids in the cake.

4.3 SETTLING TESTS

Rencken (1992) details the inaccuracy of the C-P Cell tests at low compressive pressures and a metod using settling tests was proposed. The method of determining porosity and permeability at low solids compressive pressures using measuring cylinders is described.

4.3.1 Determination of Porosity at Low Solids Compressive Pressures

Different quantities of homogenised sludge where introduced into various glass and polypropylene measuring cylinders with internal diameters ranging from 60 to 90 mm. It was assumed that the internal diameter and material of the measuring cylinders would not have any significant effects on the settling behaviour of the sludge. The initial heights of the sludge in the cylinders were recorded, and ranged from approximately 60 to 900 mm. The initial solids concentration of the sludge was sufficiently high so as to be in the region where settling occurs due to consolidation and not free or hindered settling. The sludge in the cylinders was left to stand undisturbed in a controlled environment for 3 weeks, by which time settling or consolidation of the sediment had ceased, and the final equilibrium height had been attained. The final equilibrium heights were recorded.

4.3.2 Determination of Permeability at Low Solids Compressive Pressures

Homogenised sludge was introduced into three glass measuring cylinders, at three different initial heights. The glass cylinders had an internal diameter of 50 mm and a volume of 500 ml. The initial heights of the sludges where recorded and height of the interface between the sediment and the supernatant liquid recorded with respect to time to determine the initial settling velocity of the surface of the sediment. This procedure was repeated over a range of initial solids concentrations from approximately 40 to 80 g/l.

4.4 SINGLE VERTICAL TUBE PILOT PLANT TESTS

A single tube pilot plant was erected at the Umgeni Water Process Evaluation Facility to assess the performance of the vertical tube configuration **Figure 4.3**.

4.4.1 Experimental System

The pilot plant comprises a feed tank for the sludge, a permeate collection tank and a single woven fabric tube (60 mm tube diameter).

A metering pump (feed pump) pumps sludge from the feed tank into the top of a single tube suspended vertically above the feed tank. A discharge valve on the bottom end of the tube, when closed during the filtration cycle, allows the pressure to build up inside the tube and filtration through the woven fabric is achieved. The pressure in the tube increases and is controlled at the operating pressure by a pneumatic pressure sustaining valve which directs sludge back into the feed tank. The permeate resulting from filtration through the woven fabric is collected in a PVC trough attached to the bottom of the woven tube, and is directed into the permeate tank.

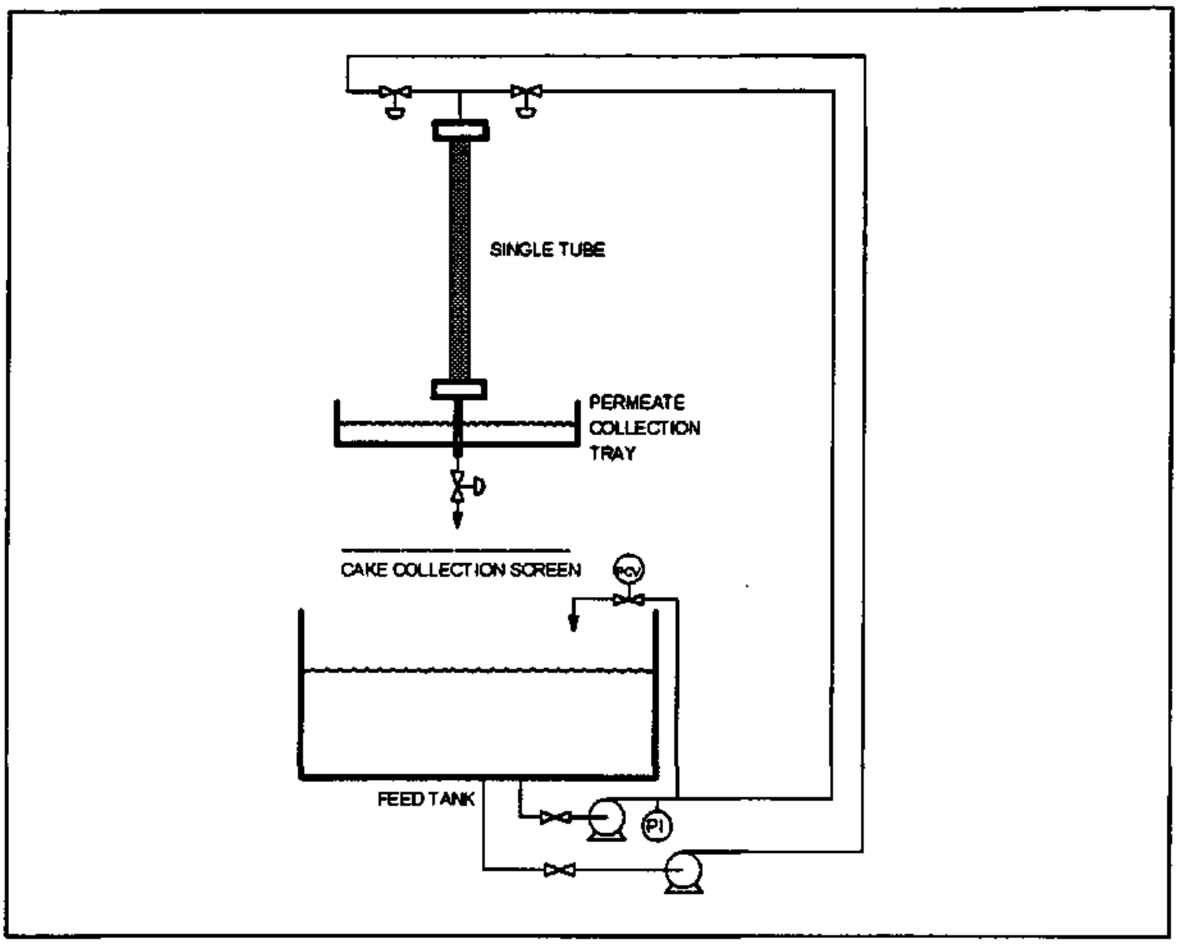

Figure 4.3 **Single-Tube Pilot Plant Flow Diagram**

A centrifugal pump is provided to assist with the removal of solids from the tube when the discharge valve is opened during the flush cycle. Feed sludge is used as a flushing fluid. The flowrate is measured and can be regulated to optimise the flushing velocity through the tube. A manual isolation ball valve is also positioned in the pipework between the flush pump and the filter. A compressor is installed to provide compressed air which is regulated at 1 bar above the required operating pressure of the system.

Initially the pilot plant was built in a fixed position and the tube was supported next to a board. A hand roller was provided to restrict the flow of flushing fluid in the tube and to simulate the action of a roller on a full-scale plant. During the progress on the project, the pilot plant became an extremely useful apparatus to assess whether the process is viable for a particular application. To enable the unit to be moved from one site to another the single tube was constructed on a road trailer. The plant has therefore been designed so that the vertical tube and associated pipework can be dismantled and transported easily.

4.4.2 Assembly of the Pilot Plant on the Trailer

Before start up of a test or a series of tests it is critical that certain aspects of the operation are checked. When the plant is assembled it is necessary to ensure that all the rubber gaskets are installed correctly to alleviate any problems due to leaks during pressurised operation.

The tube should be checked visually for damage and possible blockages to ensure that optimal operation is obtained. Adequate thread tape must be applied to the galvanised fittings and these should be tightened well, before start up of the plant. A clear tube is provided which fits over the woven tube and acts as a spray guard. Should the woven fabric rupture or develop a hole the fluid can be retained. In some applications, an acid sludge may be tested and an acid spray should be avoided wherever possible to ensure a safe working environment for the investigation. When connecting the discharge valve and permeate collection tray to the bottom of the tube it is necessary to ensure that all o-rings are placed in the PVC unions and threaded valves.

The electrical cable must be connected to an adequate power supply (380V, 20A) and the rotation of the metering pump must be checked before operation of the unit for the first time at a different location. The centrifugal pump and the compressor are supplied by a 220V supply and will operate adequately at any location.

4.4.3 Sludge Preparation

A representative sample of sludge should be collected for the tests. It can be assumed that the sample of sludge is typical for the particular application. In some instances a number of tests should be performed over a period of time to establish the repeatability of the results in order to predict the ultimate performance of the process and to avoid problems associated with variability in sludge characteristics.

The sludge should be mixed well and poured into the feed tank immediately prior to the commencement of the test. Should any period of time lapse between the filling of the feed tank and the operation of the unit, the sludge should be mixed well in the feed tank prior to the operation of the unit. During every test the sludge should be intermittently stirred to ensure that a homogeneous mixture of sludge is being pumped out of the feed tank. Variation in the feed solids concentration could impact on the results of the test.

4.4.4 Filtration Cycle

Each test is performed on a batch basis, and a homogeneous sample of the feed sludge is required. The initial volume in the feed tank is recorded to determine the mass of solids in the feed tank at the beginning of the test. The isolation valves between the feed tank and the metering pump must be opened and the feed pump started. As the woven fabric tube fills with sludge, filtration will commence and the pressure will slowly increase to the operating pressure. It is necessary to observe the increase in pressure to ensure that the pressure does not exceed the desired pressure of operation. It has been reported in WRC project No 386 *The development of Crossflow Microfiltration* and shown by VL Pillay (1992) that, should the solids be exposed to a high operating pressure in the early stages of a filtration cycle, the cake exposed to the higher pressure will have a higher specific cake resistance and will adversely affect the performance of the filter.

During the filtration cycle the particles in the sludge form a resistive layer (filter cake) on the inside of the tube, and by maintaining the pressure at a constant value the flowrate slowly decreases with time. The filtrate flows down the outside of the woven fabric and is collected in a permeate trough and is directed into the permeate tank. The flowrate of the permeate (flux) and the filtrate quality is continually monitored. Once the flux decreases to a predetermined value, the feed pump is stopped and the discharge valve opened to release the pressure in the tube. The discharge is directed over a screen into the feed tank thereby collecting any filter cake that has been formed.

4.4.5 Flush Cycle

The manual valves between the feed tank and the centrifugal (flush) pump should be opened. Sludge from the feed tank is used as a flushing fluid, and the flowrate is regulated using a manual control valve. A large flowrate of sludge is pumped through the tube and then stopped after a short period by closing the manual ball valve. This creates a pulsing action which creates a collapsing action in the woven fabric and assists with the removal of the filter cake. Initial tests were performed to establish the benefit of the pulsing action and optimise the number of pulses required as opposed to the use of a roller mechanism **Section 5.6.2.** The pulses are performed at least four times in an attempt to remove all the filter cake. After four flushes, the flush pump is allowed to continue running and the tube is compressed to a width of approximately 5 mm using a hand roller. The roller is moved slowly down the tube effectively removing any remaining filter cake.

4.4.6 Mass Balance and Calculation of Cake Recovery

The cake collected on the screen is weighed and a sample of the cake is placed in a drying oven at 105^0 Celsius. The mass of dry solids removed during each of the flush sequences as well as when a roller is applied to the tube was measured. The total mass of solids removed during the batch experiment can be calculated.

4.5 FULL-SCALE VERTICAL TUBULAR FILTER PRESS

The Vertical Tubular Filter Press was commissioned at the Umgeni Water Wiggins Waterworks as a demonstration unit. The design, construction and modifications to the plant is described in **Chapter 2**, and the flow diagram of the plant is shown in **Figure 2.2**.

4.5.1 Control of the Vertical Tubular Filter Press

A Programmable Logic Controller (PLC) was used to control the operation of the process. During start-up, thickened sludge from the Wiggins Waterworks is fed into a continually mixed pre-feed tank where provision is made for the addition of lime to the sludge. Level switches in the feed tank are used to ensure that sufficient sludge is available for operation of the process.

The operation of the Vertical Tubular Filter Press is programmed as a sequence of cycles. The plant was designed to accommodate three modules each with a double curtain and a roller. A control panel was designed that by selecting specific switches on the panel, the operation of one or more modules independantly of the other, can be selected.

Instrument air was provided from a compressor unit at the Wiggins Waterworks, and by switching solenoids on the air distribution, the pneumatic valves on the plant are controlled. Outputs from the PLC are also wired to relays in the electrical panel to control the operation of stirrer motors, conveyor belt, roller mechanism and pumps in the process.

4.5.2 Filtration Cycle

Once the correct switches have been selected on the control panel, the sequence is started by depressing the start button on the control panel. The PLC program is started, and provided the level of sludge in the feed tank is sufficiently high (as detected by level switches), the automatic valve between the feed pump and the vertical curtain (V2, **Figure 2.2**) will open, the automatic valve between the flush pump and the curtain (V1, **Figure 2.2**) will close and the feed pump will start automatically.

A pressure regulating valve (V6, **Figure 2.2**) maintains the pressure in the vertical tubes to a pre-set operating pressure (normally 300 to 400 kPa). As the woven fabric tube fills with sludge, filtration will commence and the pressure will slowly increase, the tubes will inflate **Figure 2.4a** until the operating pressure is reached. It has been reported in WRC project No 386 *The development of Crossflow Microfiltration* and shown by VL Pillay (1992) that, should the solids be exposed to a high operating pressure in the early stages of a filtration cycle, the cake exposed to the higher pressure will have a higher specific cake resistance and will adversely affect the performance of the filter. The pressure regulation although automated should be monitored to prevent excessive pressurisation of the solids and poorer operating performance.

During the filtration cycle the particles in the sludge form a resistive layer (filter cake) on the inside of the tube, and by maintaining the pressure at a constant value the flowrate slowly decreases with time. The filtrate flows down the outside of the woven fabric and is collected in a permeate collection tray and is directed into the permeate tank. The flowrate of the sludge pumped into the vertical curtain is measured by an electromagnetic flowmeter which provides an analogue output (pulse) to the PLC. The time between successive pulses is measured by the PLC and once this becomes greater than a preset value the filtration cycle is terminated. The feed pump is stopped and the discharge valve (V4, **Figure 2.2**)at the bottom of the vertical curtains is opened to release the pressure in the tubes. The discharge is directed over a conveyor belt into the feed tank thereby collecting any filter cake that has been formed.

4.5.3 Flush Cycle

Once the vertical curtain has drained the PLC starts the flush pump. The flush pump is started against a closed valve, and a series of timers then opens and closes the valve between the flush pump and the vertical curtain (V1, **Figure 2.2**). Sludge from the feed tank is used as a flushing fluid, and the flowrate is regulated using a manual control valve. During each flushing pulse, a large flowrate of sludge is pumped through the curtain, creating a collapsing action in the woven fabric and assisting with the removal of the filter cake. Initial tests were performed to establish the benefit of the pulsing action and optimise the number of pulses required as opposed to the use of a roller mechanism **Section 5.6.2**. The pulses are performed at least four times in an attempt to remove all the filter cake.

4.5.4 Roller Movement and Tube Cleaning

Whilst the flush pump is still operating after the sequence of flushes is complete, a pair of rollers are engaged and are driven slowly down the length of the vertical curtain **Figure 2.4b**. A restriction is created between the rollers and the velocity of the flushing fluid is increased as it is pumped through the tubes. The shearing effect of the increased velocity ensures that all the filter cake that has been deposited during the filtration cycle is removed.

Once the rollers have reached the bottom of the curtain a limit switch signals to the PLC that the cleaning of the curtains is complete, and the sequence of filtration is initiated.

4.5.4 Cake Collection and Mass Balance

During the flush cycle and the tube cleaning using the rollers, the conveyor belt is automatically started, and filter cake that is formed in the tubes is collected and transported from the curtain discharge to a sludge hopper. The cake collected in the hopper is weighed and a sample of the cake is placed in a drying oven at 105° Celsius. The mass of dry solids removed during each of the flush sequences as well as when a roller is applied to the curtains was measured. The total mass of solids removed during the batch experiment can be calculated.

The electromagnetic flowmeter provides a digital display of the total volume of sludge pumped to the vertical curtains. A sample of the feed sludge is also dried at 105° Celsius, and the total mass of solids pumped into the vertical curtains can be calculated.

The volume of feed sludge from the Wiggins Waterworks was not measured. By measuring the mass of solids fed to the Vertical Tubular Filter Press, the overall recovery of the process could be determined. This was however not possible duringthe project, but is recommended for future installation.

5 Results and Discussion

The Wiggins Waterworks is currently undergoing an upgrade in capacity, and significant modifications were made to the sludge treatment process at the works resulting in periods where there was no power or sludge available for the operation of the single tube pilot plant or the Vertical Tubular Filter Press.

During the operation of the Vertical Tubular Filter Press, some mechanical modifications were required. This resulted in large periods of time when the plant was not or could not be operated. Additional equipment had to be acquired and added to the unit to achieve continuous (24 hour) operation. In an attempt to transfer the technology during the project, Umgeni Water considered the use of the Vertical Tubular Filter Press at one of its waterworks. The results of the laboratory scale tests are presented. The performance of the single tube pilot plant as well as the Vertical Tubular Filter Press are discussed, and aspects of modelling of the process are presented.

5.1 NATURE OF THE SLUDGE AND WIGGINS RAW WATER QUALITY

Water from the Inanda Dam (KwaZulu Natal) is treated at the Wiggins Waterworks. Following pre-treatment of the water with either chlorine or ozone, the suspended and colloidal solids in the water are coagulated using a blended polymeric coagulant. The use of ozone at Wiggins Waterworks during the period of the project was limited due to an upgrade of the process at the works. The coagulant used during the period of the project was a blended poly-aluminum-chloride (PAC) and di-methyl-di-allyl- ammonium-chloride (DMDAAC) which has been found to be efficient and cost effective in treating the water from the Inanda Dam.

Once the solids have been coagulated, the flocs formed are allowed to settle in *Degremont type* pulsator clarifier, thereby concentrating the solids to between 0,2 and 0,5 % (m/m) in the form of a sludge. The water is filtered through rapid gravity sand filters, and the backwash water emanating from the filters is combined with the sludge from the clarifiers and further thickened before solids dewatering. The following chemicals and water quality parameters may affect the nature of the sludge and the efficiency of dewatering.

- ***Turbidity / Suspended Solids*** - During periods where the turbidity of the water is low, the solid particles are very small and colloidal in nature. The flocs formed during coagulation are a collection of fine particles bound together by a polymer. When the turbidity is higher, normally after periods of above average rainfall, the solids in the water are more particulate and the resulting sludge is easier to dewater.

- *Lime Addition* - The pH of the raw water is adjusted prior to coagulation. The particles of lime which are not totally dissolved in the water will be collected in the sludge. Although these particles will tend to improve the filterability of the sludge the effect of a small percentage of particles remaining in the sludge will be minimal.
- *Polymeric Coagulant Addition* - The polymeric coagulant has an effect of destabilisation during coagulation and results in particles being bound together as flocs. A large presence of organics may result in a more difficult sludge to dewater, especially during periods of low turbidity (i.e. majority colloidal particles). These effects could not be measured during this project.
- *Bentonite Addition* - Bentonite is added as a coagulant aid when the turbidity is low and problems are experienced in coagulating the particles in the raw water. **Figure 5.1** shows the bentonite dose applied at Wiggins Waterworks, as well as the raw water turbidity during the period of the project.

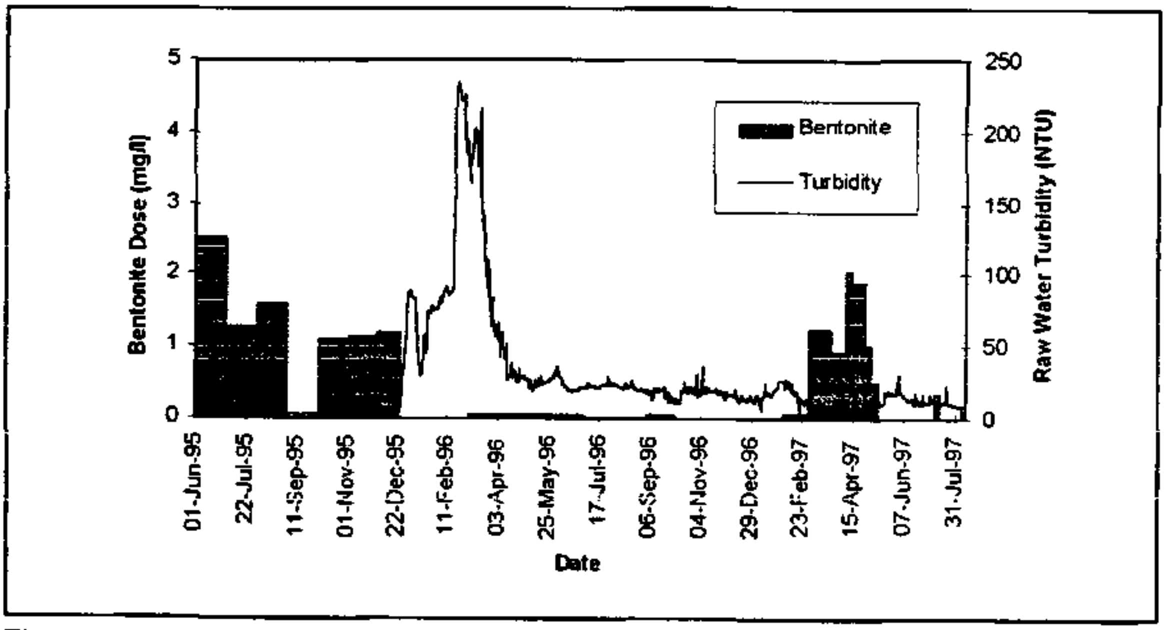

Figure 5.1 **Raw Water Turbidity and Bentonite Dose Applied at the Umgeni Water Wiggins Waterworks**

During the period prior to the operation of the Vertical Tubular Filter Press on a continuous basis (March to May 1997) bentonite was added to the raw water. This had a detrimental affect on the operability of the plant and the performance of the filter for the dewatering of the sludge at Wiggins Section 5.7.

5.2 DENSITY MEASUREMENT

Two methods of determining the solids density were evaluated. The *Evaporation Method* **Section 4.2.1.1** which relies on an accurate measurement of the volume of sludge as well as the assumption that all the water is evaporated from between the solids particles. The *Specific Gravity Bottle Method* **Section 4.2.1.2** measures the volume of water and solids by difference of mass in a calibrated bottle. The mass of the bottle, solids and water can be accurately measured to one tenth of a milligram.

Table 5.1 shows that the results obtained using the *Evaporation Method* are not repeatable, probably due to the inaccuracy in measuring the volume of the sludge as opposed to the measurement of mass in the *Specific Gravity Bottle Method*. The results from the latter method are reasonably consistent, and this method is therefore recommended for any future investigations of this nature.

Table 5.1 Solids Density Measurement

	SOLIDS DENSITY VALUES (kg/m^3)	
	Evaporation Method	SG Bottles Method
1	2800	2289.7
2	3210	2380.7
3	2330	2333.9
4		2275.6
5		2358.6
Average	2780	2310

When performing experiments with the addition of lime to improve the filtration it is necessary to adjust the solids density for the mixture of sludge and lime. The density of lime was determined to be 2240 kg/m^3 and the density of the solids in the sludge is 2310 kg/m^3. In this experiment the difference in the densities is small and a weighted average of the individual solid densities was used. For example :

Mixture 95 % solids and 5 % lime : $2310 \times 0.95 + 2240 \times 0.05 = 2306.5$ kg/m^3

5.3 COMPRESSION-PERMEABILITY CELL TESTS

Two series of C-P Cell experiments were performed on sludge from the Wiggins Waterworks. A standard C-P Cell which ignores wall friction effects was set up at the Process Evaluation Facility. A second C-P Cell was constructed at the University of Natal Chemical Engineering Department to investigate the effects of wall friction as part of a MSc (Eng.) project.

5.3.1 Compression-Permeability Cell Tests - Cell A

The objective of the C-P Cell tests using the design given by Rowe et al (1966) and used by Rencken (1992) is to determine the sludge characteristics and to assess the benefits that can be achieved by adding lime to the sludge. When bentonite was used as a coagulant aid in water treatment at HD Hill waterworks, the filterability of the sludge was found to deteriorate. By adding lime to the sludge, the filtration improved and the plant was operable.

5.3.1.1 Experimental Technique

During the C-P Cell experiments, pressure is exerted on a sample of sludge, and the resulting cake is allowed to stabilise, before the flow of permeate through the compressed sludge is measured.

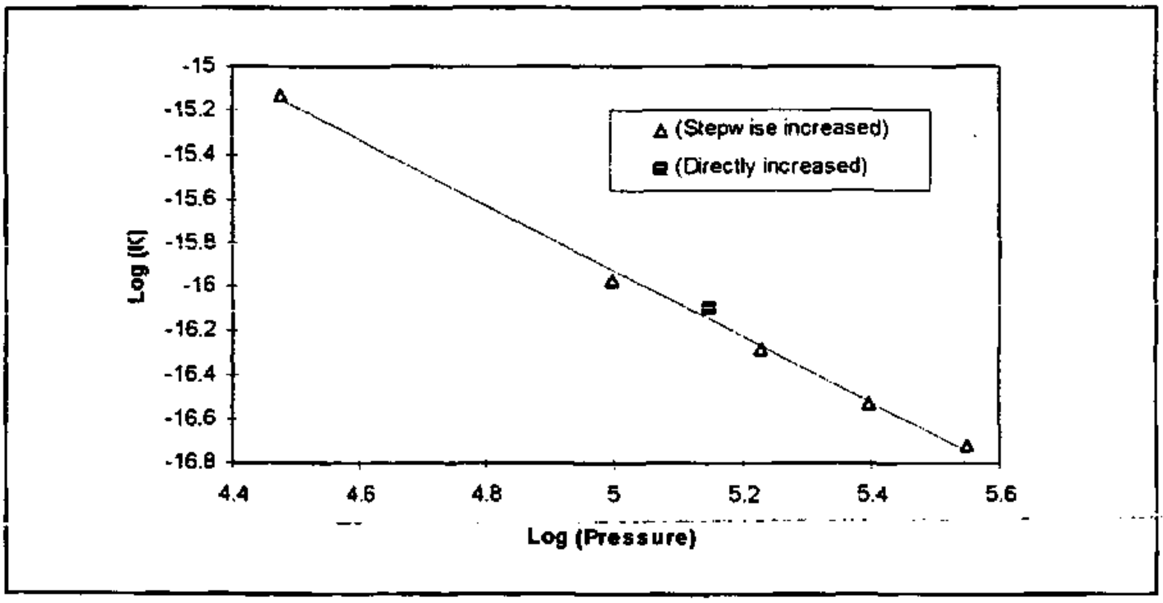

Figure 5.2 **Comparison of Experimental Technique by Increasing *the Pressure Incrementally or Directly***

The pressure is incrementally increased using the same cake sample to obtain data points throughout the working pressure range of the filter. An experiment was performed to determine the effects of the methodology of the testing procedure by immediately exposing the sludge to a much larger pressure at the beginning of the test. It was found that the log of the permeability vs. log of the solids compressive pressure relationship **Figure 5.2** was not significantly affected by a change in the experimental technique. This test also served to confirm the previous test results and indicates that although the test procedure differed there is repeatability in the results.

Two sets of experiments were conducted in an attempt to establish whether repeatability is obtained in different tests. The experiments used the technique of incrementally increasing the pressure. The incremental pressure setting were however different in each test.

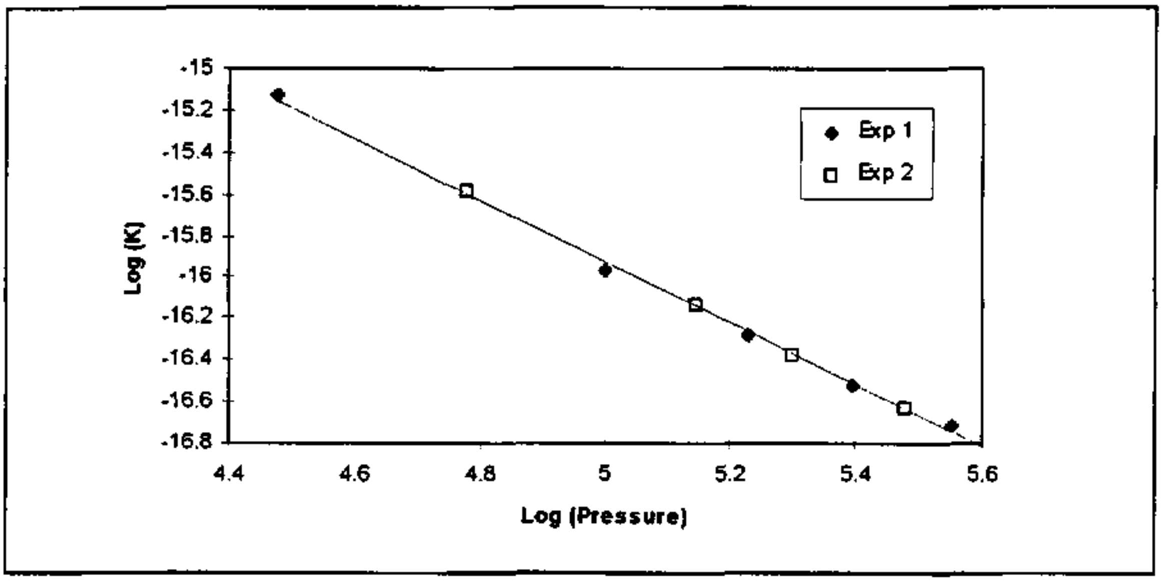

Figure 5.3 **Comparison of Two Separate Sets of Experiments to Establish Repeatability**

Figure 5.3 shows that a repeatability between different tests can be achieved as the log of the permeability is linearly proportional to the log of the solids compressive pressure for both experiments. Two samples of sludge were used for the separate tests.

5.3.1.2 Filter Medium Resistance

In the calculation of permeability it is assumed that the filtration resistance is due to the compressed cake only and that the resistance of the filter media (porous disc, filter papers, pipe friction, etc.) is negligible. To verify this assumption an experiment was conducted to determine the combined resistance of the medium, tubing and fittings of the apparatus.

Table 5.2 **Filter Medium Resistance**

Calculated resistance (filter papers, fittings etc. only)	8.551×10^{10} m^{-1}
Calculated resistance (filter papers and cake @ 140 kPa)	6.465×10^{13} m^{-1}
Percentage of total resistance due to the filter medium	0.132 %

Table 5.2 shows that the percentage of the resistance due to the medium only, in comparison to the total filtration resistance (cake + filter media) is 0.132 %, which is very low and can be considered to be negligible. The assumption that the resistance is negligible is therefore verified.

5.3.1.3 The Effect of Lime Addition on Filter Cake Characteristics

The effect of dosing lime in the sludge was investigated. An attempt was made to establish if the effect of dosing lime in the sludge would improve the permeability and porosity, and hence the filterability of the sludge. A series of C-P Cell experiments were conducted to determine the effects of lime addition on the permeability and porosity with solids compressive pressure. Lime was added to the sludge prior to the experiment and the sample was thoroughly mixed to ensure a homogeneous sludge. Lime concentrations of 5, 10 and 15 % (mass of lime /mass of solids) were investigated.

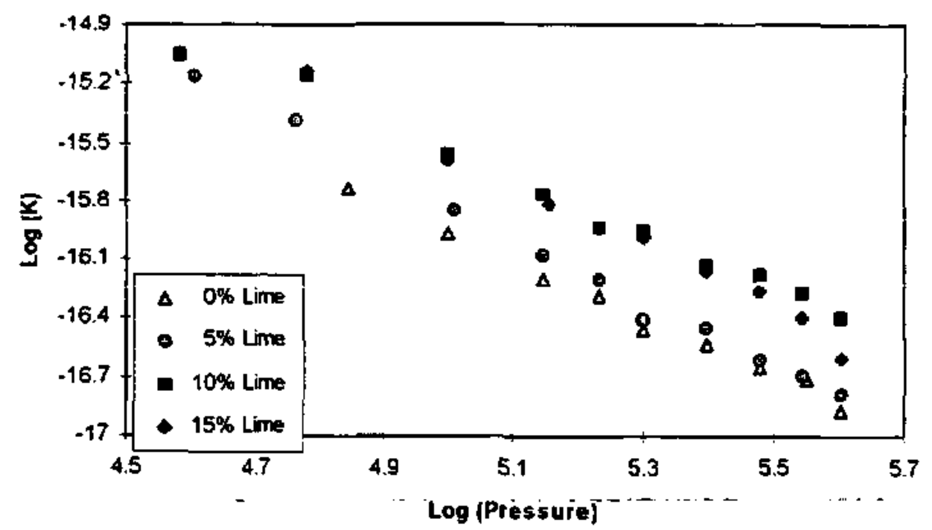

Figure 5.4 **Solids Compressive Pressure versus Permeability**

Figure 5.4 shows that the permeability increases with an increase in lime concentration. The graph shows that a lime concentration of 5% may not improve the permeability significantly, but when the concentration of lime was increased to 10 and 15 %, a marked improvement in the permeability was evident.

Figure 5.5 shows again that the addition of 5 % lime does not appreciably alter the sludge characteristics. By adding 10 and 15 % lime to the sludge the porosity of the sludge is increased, particularly at lower solids compressive pressures. These results are consistent as a more porous cake structure results in a more permeable cake and an improved filterability.

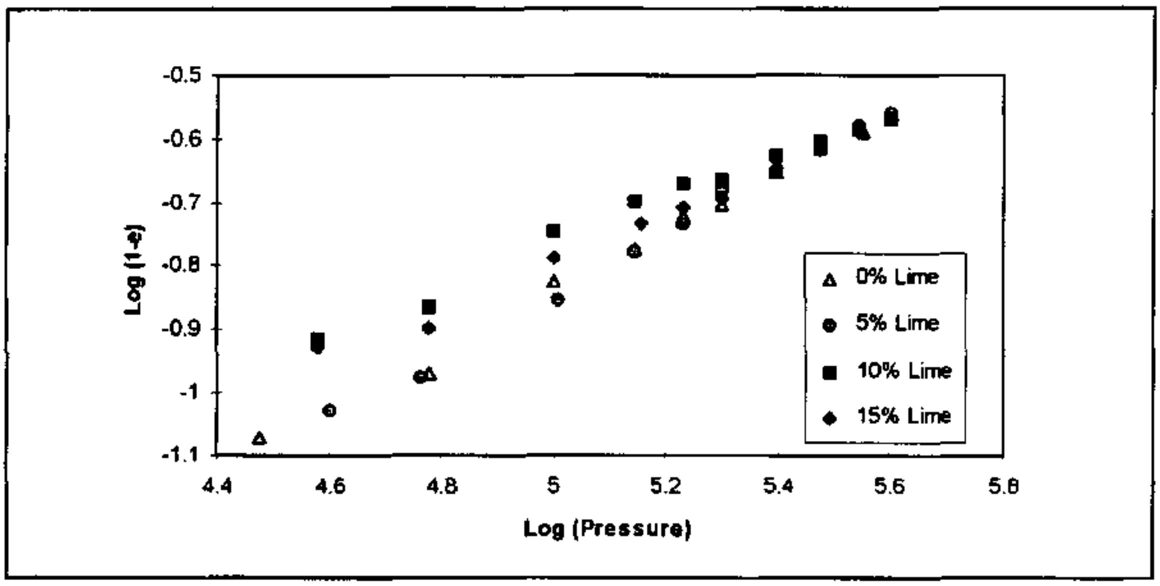

Figure 5.5 **Solids Compressive Pressure versus Solids Volume Fraction**

The specific cake resistance (α) is a better parameter to observe the effects of adding lime to the sludge as it combines both the permeability (K) and the porosity (ε) in its definition.

$$\alpha = \frac{1}{(1-\varepsilon)\,K\,\rho_s} \qquad (5.1)$$

The benefit of reducing the specific resistance of the cake is that filtration rates are increased and a subsequent increase in the mass of sludge that can be filtered by the Vertical Tubular Filter Press per day.

Figure 5.6 shows clearly that the addition of lime into the sludge assists in reducing the specific cake resistance as had been observed by Rencken (1992). The lime concentration of 5% did not have a noticeable effect while the concentration of 10 and 15% reduced the resistance significantly. Problems experienced with performing the C-P Cell experiments, especially with lime resulted in a limited amount of experimental data. Had more experiments been performed, the effect of the addition of 5 % lime may have been more noticeable. During the full-scale plant operation **Section 5.7.2** the addition of between 3 and 5 % lime significantly improved the recovery of solids of the Vertical Tubular Filter Press, and resulted in a much smoother and manageable operation.

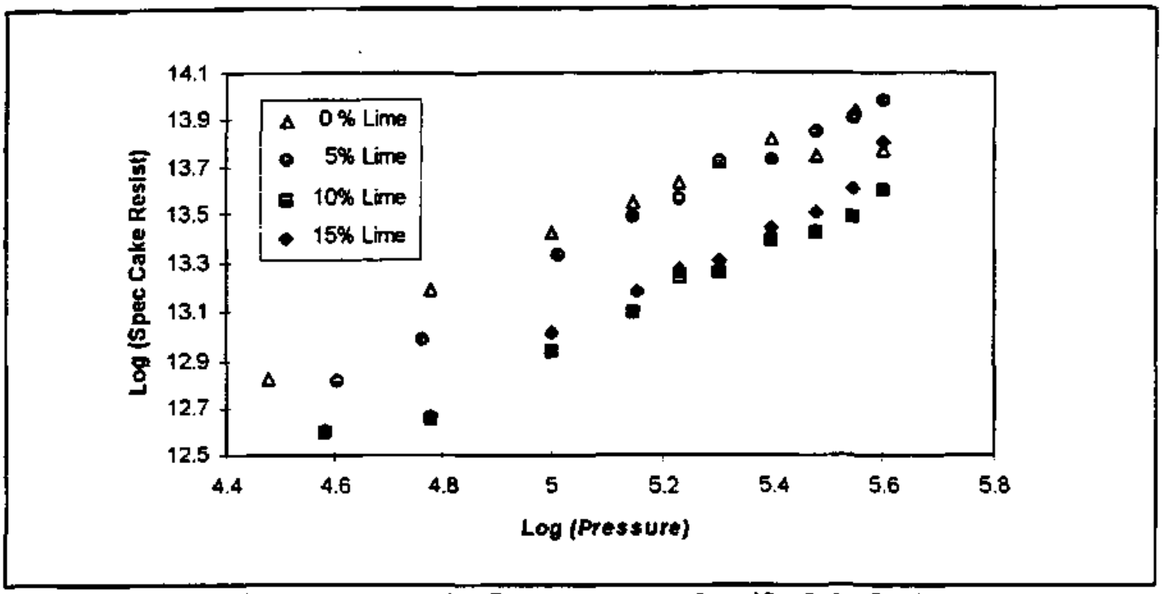

Figure 5.6 **Solids Compressive Pressure versus Specific Cake Resistance**

The empirical constants for the permeability and porosity correlations **Section 3.1** for each of the C-P Cell tests conducted are tabulated in **Table 5.3**.

Table 5.3 **Empirical Constants Calculated from Experimental Data**

Exp. No	Lime Conc. (% m/m)	Permeability Empirical Constants		Porosity Empirical Constants	
		$F \times 10^8$	δ	$B \times 10^3$	β
1A	0	4.17	1.72	0.5	0.5
1B	0	0.28	1.49	1.19	0.42
2A	0	0.33	1.49	0.82	0.45
2B	0	0.01	1.22	0.47	0.5
3	5	2.91	1.66	0.5	0.49
4	10	0.18	1.36	3.19	0.35
5	15	0.98	1.51	2.35	0.37

The compressibility of the cake, given by the exponents δ and β, is relatively unaffected by the addition of lime in the concentration range investigated. Although the addition of lime has been shown to improve the characteristics of the sludge its effect on the 'structural stability of the cake' and hence the recovery of the cake needs to be determined.

5.3.2 Compression-Permeability Cell Tests - Cell B

The purpose of these experiments was to determine the effects of wall friction on the accuracy of the C-P cell for obtaining sludge permeability and porosity characteristics. The analysis of the results is divided into two sections. The results are initially presented assuming no wall friction effects (standard analysis), and then corrected for the effects of side wall friction.

5.3.2.1 Standard Analysis

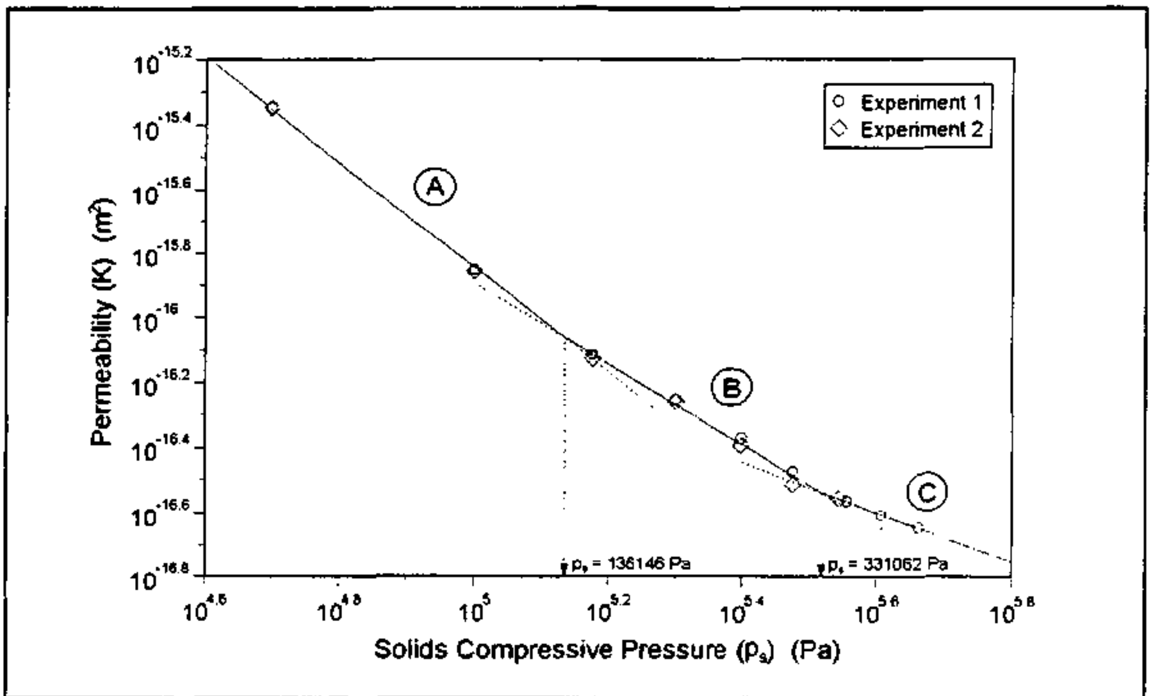

Figure 5.7 **Permeability versus Solids compressive Pressure for the Compression-Permeability Cell Experiments**

Plots of permeability and solids volume fraction versus solids compressive pressure for the two tests are shown in **Figure 5.7** and **Figure 5.8** respectively. Over the range of solids compressive pressure for the tests, the data was subdivided into regions that showed greater individual linearity.

Linear regression analysis using the permeability versus solids compressive pressure data in each of the regions A, B and C **Figure 5.7**, and over the entire data range ABC, for the two tests yielded values for F and δ in each of these regions **Equations 3.3**. The results of the regression analysis are shown in **Table 5.4**.

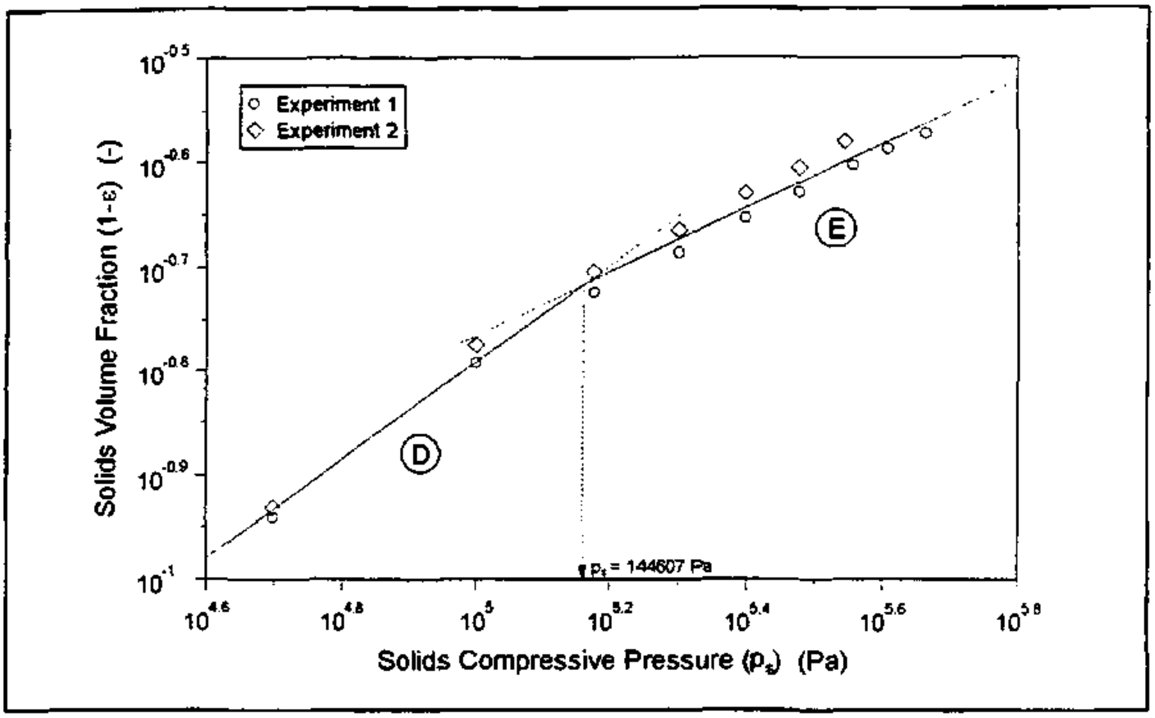

Figure 5.8 **Solids Volume Fraction versus Solids Compressive Pressure for Compression-Permeability Cell Experiments**

Table 5.4 **Regression Analysis on Permeability Data for Compression-Permeability Cell Experiments 1 and 2 Combined**

	Linear Region			
	A	B	C	ABC
F	2.008×10^{-8}	2.063×10^{-10}	4.495×10^{-13}	1.069×10^{-9}
δ	1.629	1.242	0.759	1.37
r^2	0.999	0.989	0.998	0.989

Where r^2 = correlation coefficient for the linear regression analysis.

The intersections of each of the linear regions was calculated from the regression data and found to be at compressive pressures p_s = 136146 Pa (region A-B) and p_s = 331062 Pa (region B-C).

Linear regression analysis using the porosity versus solids compressive pressure data in each of the regions D and E in **Figure 5.8**, and over the entire data range DE, for the two tests yielded values for B and β in each of these regions **Equations 3.3**. The results of the regression analysis are shown in **Table 5.5**.

Table 5.5 **Regression Analysis on Porosity Data for Compression-Permeability Cell Experiments 1 and 2 Combined**

	Linear Region		
	D	E	DE
B	7.337×10^{-4}	5.036×10^{-3}	2×10^{-3}
β	0.469	0.306	0.38
r^2	0.989	0.943	0.976

The intersection of the two linear regions was calculated from the regression data and is p_s = 144607 Pa (region D-E).

5.3.2.2 Approximate Correction for Side Wall Friction

A computer program (written in the C programming language), was written to analyse the C-P cell data and to determine the parameters $k_0 f$ and c in the wall friction model described in **Section 3.5.2** such that the objective function given by **Equation 3.69** was minimised. The program employed a numerical direct search optimisation technique known as the *Simplex method*.

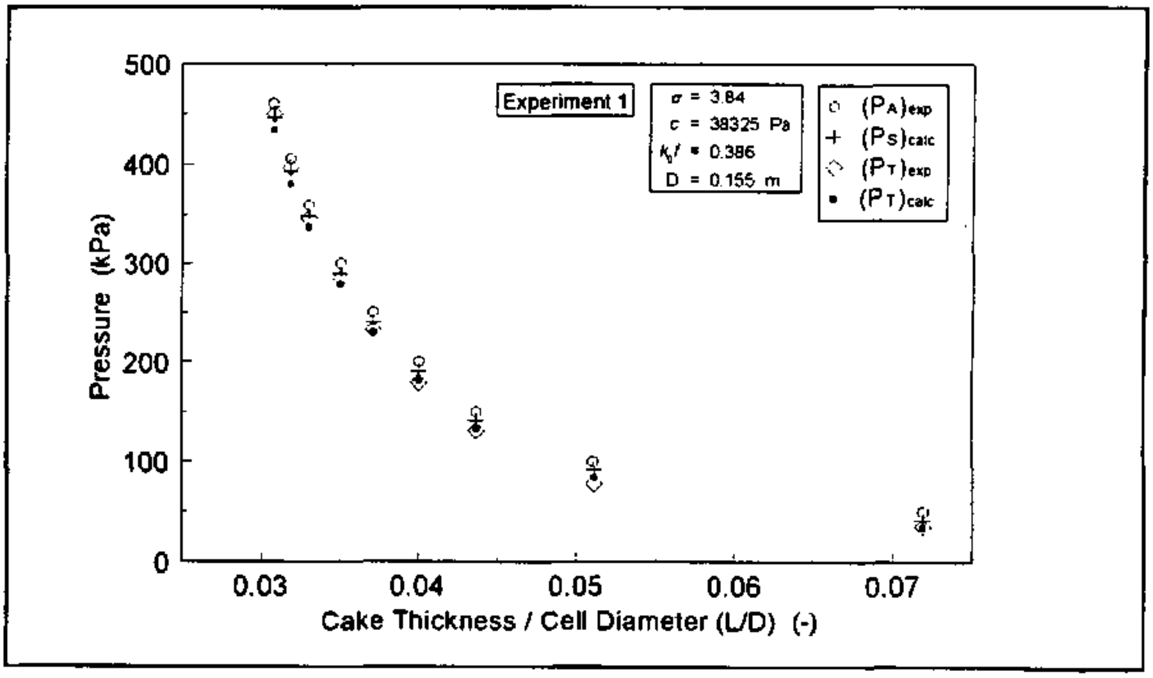

Figure 5.9 **Results of Wall Friction Analysis for Experiment 1**

The results of the analysis for the C-P cell experiments 1 and 2 is shown in **Figures 5.9 and 5.10** respectively and given in **Table 5.6**.

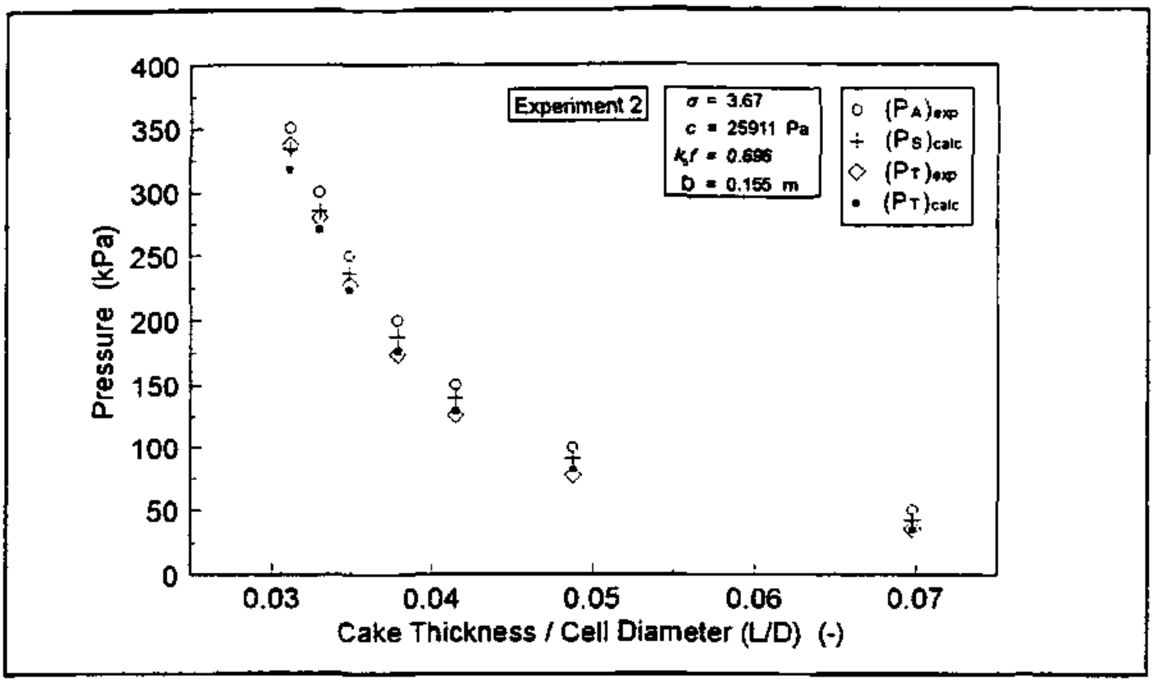

Figure 5.10 **Results of Wall Friction Analysis for Experiment 2**

Table 5.6 **Wall Friction Model Parameters for Compression-Permeability Cell Exeriments 1 and 2**

	Test 1	Test 2
$k_o f$	0.386	0.696
c	38325 Pa	25911 Pa
σ	3.842	3.671

The accuracy of the wall friction model for predicting the transmitted pressure through the cake decreases for applied pressures greater than 250 kPa. In some cases the average solids compressive pressure through the cake as determined by **Equation 3.65** is less than the experimentally measured transmitted pressure. For this reason, the corrected solids compressive pressure for the C-P cell tests for applied pressures greater than 250 kPa was determined by taking the average of the experimentally measured applied and transmitted pressures. For applied pressures of 250 kPa and less the corrected solids compressive pressure was determined from the wall friction model using **Equation 3.68**.

A plot of the permeability and solids volume fraction versus the *corrected* solids compressive pressure for the two experiments are shown in **Figures 5.11** and **5.12** respectively. Over the range of solids compressive pressure for the tests, the data was subdivided into regions that showed greater individual linearity.

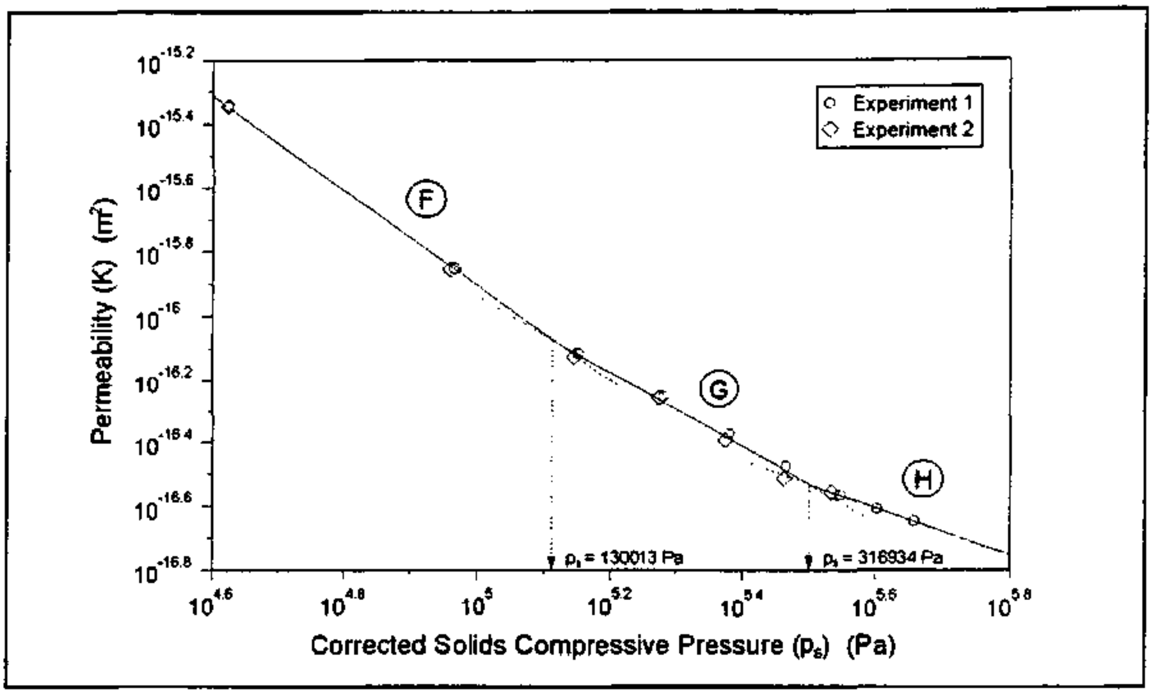

Figure 5.11 **Permeability versus Corrected Solids Compressive Pressure for Compression-Permeability Cell Experiments**

Linear regression analysis using the permeability versus *corrected* solids compressive pressure data in each of the regions F, G and H **Figure 5.11**, and over the entire data range FGH, for the two tests yielded values for F and δ in each of these regions **Equations 3.3**. The results of the regression analysis are shown in **Table 5.7.**

Table 5.7 **Linear Regression Analysis on Permeability Data for Compression-Permeability Cell Experiments 1 and 2 Combined and the Corrected Solids Compressive Pressure**

	Linear Region			
	F	G	H	FGH
F	3.035×10^{-9}	9.641×10^{-11}	3.283×10^{-13}	3.355×10^{-10}
δ	1.477	1.184	0.736	1.281
r^2	0.999	0.987	0.999	0.991

The intersections of each of the linear regions was calculated from the regression data and found to be p_s = 130013 Pa (region F-G) and p_s = 316934 Pa (region G-H).

Linear regression analysis using the porosity versus *corrected* solids compressive pressure data in each of the regions I and J **Figure 5.12**, and over the entire data range IJ, for the two tests yielded values for B and β in each of these regions **Equations 3.3**. The results of the regression analysis are shown in **Table 5.8**.

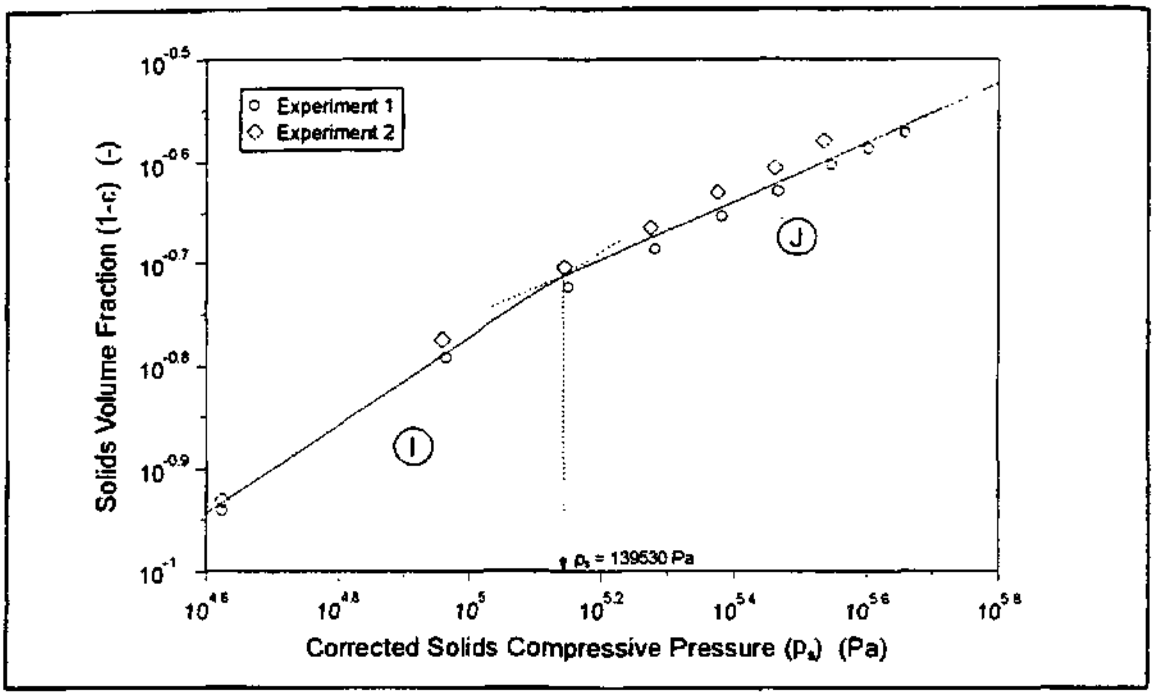

Figure 5.12 **Solids Volume Fraction versus Corrected Solids Compressive Pressure for Compression-Permeability Cell Experiments**

The intersection of the two linear regions was calculated from the regression data and is $p_s = 139530$ Pa (region I-J).

Table 5.8 **Linear Regression Analysis on Porosity Data for Compression-Permeability Cell Experiments 1 and 2 Combined and the Corrected Solids Compressive Pressure**

	Linear Region		
	I	J	IJ
B	1.265×10^{-3}	6.257×10^{-3}	2.758×10^{-3}
β	0.425	0.29	0.355
r^2	0.988	0.937	0.978

5.3.3 Common Experimental problems

C-P cell tests are time consuming to setup and a single experiment can last in excess of 12 hours, during the course of the experiment the test is prone to various operational failures and as a result experiments often have to be aborted.

Some of the more common experimental problems are as follows :

- the top rubber membrane deteriorates with time and often develops pin-hole leaks, allowing air to enter the cake sample,
- if an inadequate seal is formed between the bottom filter paper or the cake sample and the cell wall, or if the cake develops small cracks, the permeate bypasses the cake sample and results in exaggerated permeate flow rates, and
- in the case of *cell B*, if the fluid filled chamber is not adequatley sealed, it will slowly drain under pressure, the resultant movement of the cell base deforms the cake sample and results in erraneous dial guage readings.

5.4 SETTLING TESTS

5.4.1 Determination of Porosity at Low Solids Compressive Pressures

Settling tests were performed using different volumes of sludge in measuring cylinders and allowing the sludge to settle over a period of 3 weeks. A plot of the final equilibrium height of the sediment versus the volume dry solids per unit cross-sectional area is shown in **Figure 5.13**.

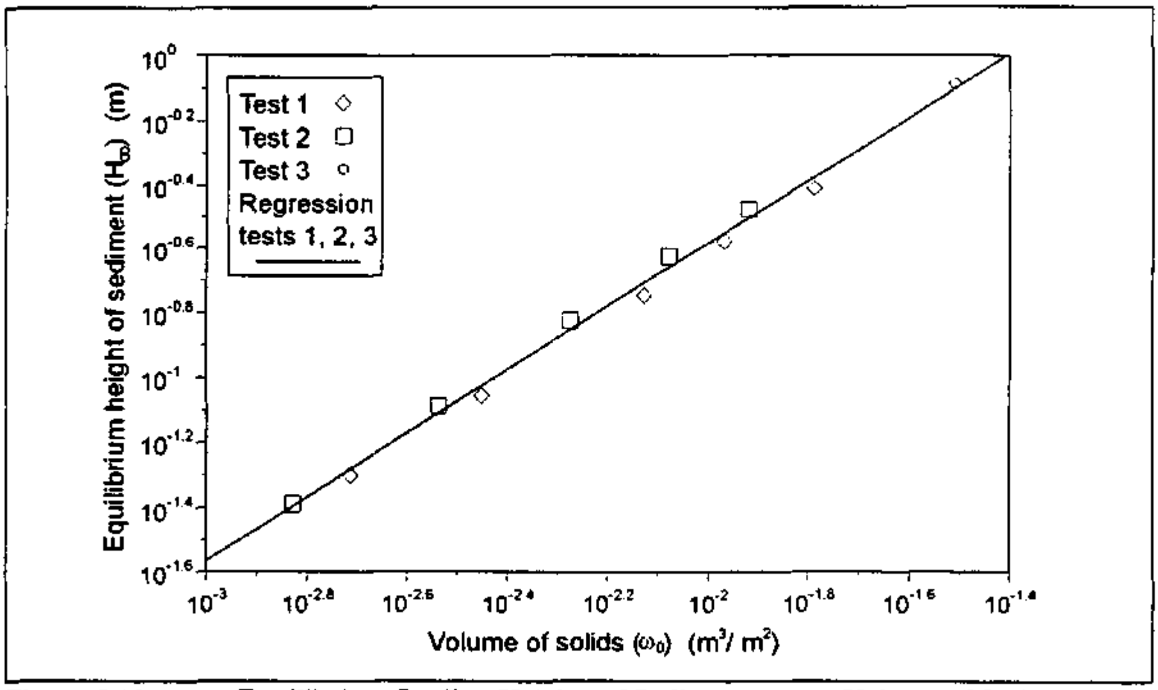

Figure 5.13 Equilibrium Settling Heights of Sediment versus Volume of Solids

A linear regression analysis on the data yields values for the parameters a and b in accordance with **Equation 3.77 of Section 3.61**.

$$a = 23.86$$

$$b = 0.9809$$

$$r^2 = 0.995$$

From **Equations 3.75 and 3.76** the parameters B and β respectively, where determined **Equations 3.3**. The solids compressive pressure range for these experiments was, $19.1 \le p_s \le 399.1$ Pa.

$$B = 0.03565$$

$$\beta = 0.01915$$

5.4.2 Determination of Permeability at Low Solids Compressive Pressures

Settling tests were performed using different initial concentrations of sludge. A plot of the initial settling velocity of the surface of the sediment versus the initial porosity of the suspension is shown in **Figure 5.14**. The dilute, intermediate and consolidation settling regions where identified.

$$\varepsilon_L = 0.977$$

$$\varepsilon_{if} = 0.974$$

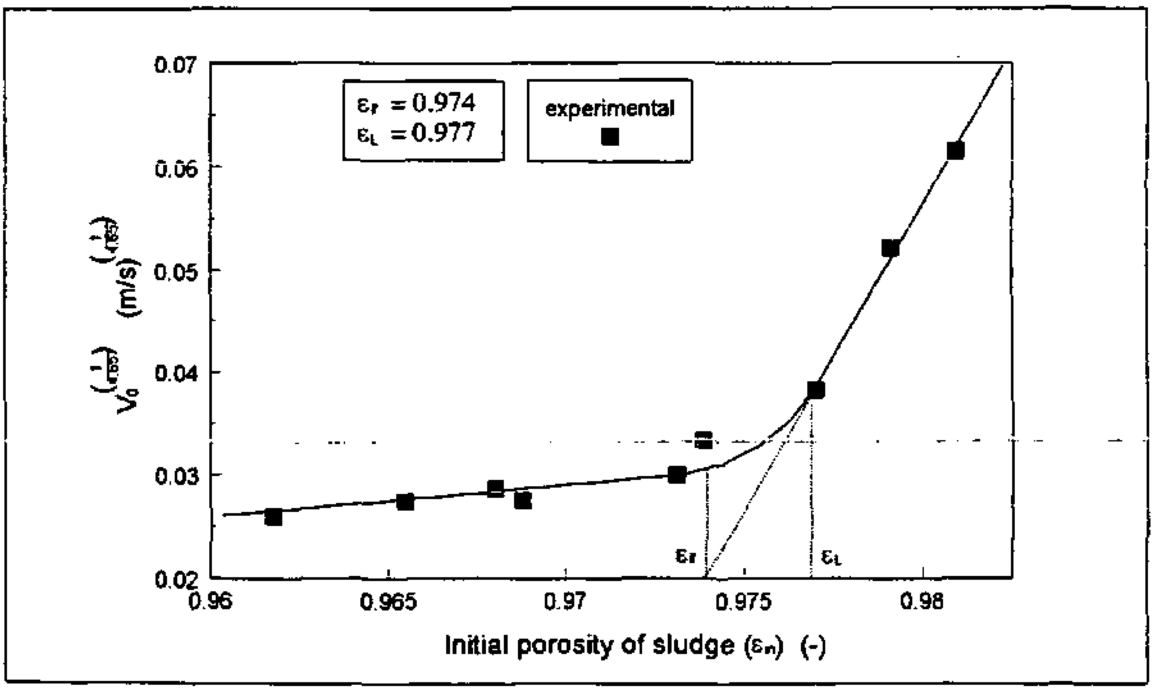

Figure 5.14 **Initial Settling Velocity of Sediment Surface versus Initial Porosity of the Suspension**

For each point in the consolidation settling region, $\varepsilon_m < \varepsilon_{if}$, the permeability was determined from **Equation 3.84** and the solids compressive pressure from the initial porosity of the suspension and the porosity correlation parameters determined in **Section 5.4.1** above, in accordance with **Equation 3.74**. A plot of the permeability versus solids compressive pressure is shown in **Figure 5.15**. A linear regression analysis on the data yielded values for F and δ, **Equations 3.3**. The solids compressive pressure range for these experiments was $4 \times 10^{-7} \le p_s \le 40.4$ Pa.

$$F = 1.081 \times 10^{-13}$$

$$\delta = 0.05381$$

$$r^2 = 0.945$$

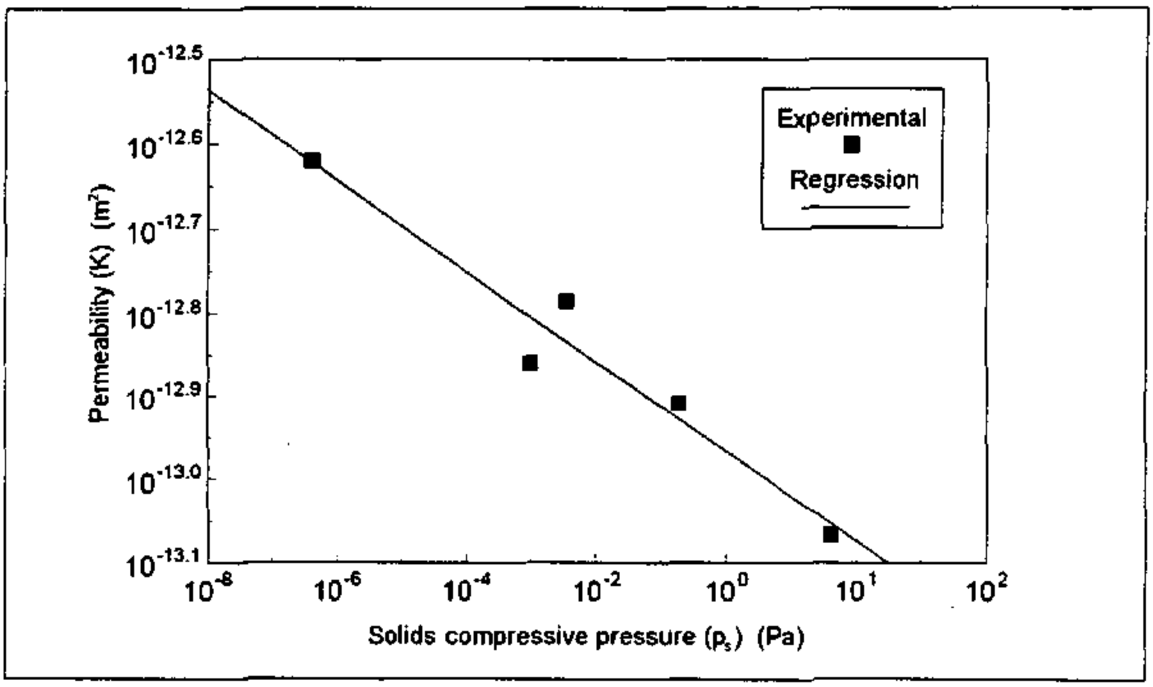

Figure 5.15 **Permeability versus Solids Compressive Pressure for Settling Experiments**

5.5 PERMEABILITY AND POROSITY CORRELATIONS

The results of the C-P cell and settling tests are fitted to the permeability and porosity correlations for multiple parameter sets, **Equations 3.3**.

Equations (3.3) require that a solids compressive pressure, p_{si}, be identified where the porosity and permeability are assumed constant for solids compressive pressures below this value. This is achieved by assigning to p_{si} an arbitrarily determined value or by rearranging **Equation (3.3.d)** and solving for the solids compressive pressure where the porosity is given by the porosity of the feed sludge.

$$p_{sf} = p_{si} = \left(\frac{1-\varepsilon_f}{B}\right)^{\frac{1}{\beta}} = \left(\frac{1-\varepsilon_f}{0.03565}\right)^{\frac{1}{0.01915}} \qquad (5.2)$$

where p_{sf} = solids compressive pressure corresponding to the porosity of the feed sludge, (Pa)

ε_f = porosity of feed sludge, (-)

The multiple correlation parameter sets are given below. The cake compressibility varies with solids compressive pressure, for very low solids compressive pressures the cake is mildly compressible, generally the cake is super-compressible with the compressibility of the cake decreasing slightly with solids compressive pressure in the higher solids compressive pressure range. The data corrected for the effects of wall friction show a slightly decreased compressibility (indicated by the magnitude of the exponents). The effects of correcting the data for wall friction will only be able to be fully assessed once the correlation data has been incorporated into the predictive filtration model and the results compared to filtration data.

5.5.1 Wall Friction in Compression-Permeability Cell Tests Neglected

$K = 1.081 \times 10^{-13} p_{sf}^{-0.05381}$	$0 \le p_s \le p_{sf}$	(5.3.a)
$K = 1.081 \times 10^{-13} p_s^{-0.05381}$	$p_{sf} \le p_s \le 2214$ Pa	(5.3.b)
$K = 2.008 \times 10^{-8} p_s^{-1.639}$	$2214 \le p_s \le 136146$ Pa	(5.3.c)
$K = 2.063 \times 10^{-10} p_s^{-1.242}$	$136146 \le p_s \le 331062$ Pa	(5.3.d)
$K = 4.495 \times 10^{-13} p_s^{-0.759}$	$p_s \ge 331062$ Pa	(5.3.e)

$(1-\varepsilon) = 0.03565 p_{sf}^{0.01915}$	$0 \le p_s \le p_{sf}$	(5.3.f)
$(1-\varepsilon) = 0.03565 p_s^{0.01915}$	$p_{sf} \le p_s \le 5664$ Pa	(5.3.g)
$(1-\varepsilon) = 7.337 \times 10^{-4} p_s^{0.4695}$	$5664 \le p_s \le 144607$ Pa	(5.3.h)
$(1-\varepsilon) = 5.036 \times 10^{-3} p_s^{0.3064}$	$p_s \ge 144607$ Pa	(5.3.i)

5.5.2 Wall Friction in Compression-Permeability Cell Tests Accounted

$K = 1.081 \times 10^{-13} p_{sf}^{-0.05381}$	$0 \le p_s \le p_{sf}$	(5.4.a)
$K = 1.081 \times 10^{-13} p_s^{-0.05381}$	$p_{sf} \le p_s \le 1133$ Pa	(5.4.b)
$K = 3.035 \times 10^{-9} p_s^{-1.477}$	$1133 \le p_s \le 130013$ Pa	(5.4.c)
$K = 9.641 \times 10^{-11} p_s^{-1.184}$	$130013 \le p_s \le 316934$ Pa	(5.4.d)
$K = 3.283 \times 10^{-13} p_s^{-0.7358}$	$p_s \ge 316934$ Pa	(5.4.e)

$(1-\varepsilon)=0.03565p_{sf}^{0.01915}$	$0 \le p_s \le p_{sf}$	(5.4.f)
$(1-\varepsilon)=0.03565p_s^{0.01915}$	$p_{sf} \le p_s \le 3751$ Pa	(5.4.g)
$(1-\varepsilon)=1.265\times10^{-3}p_s^{0.4248}$	$3751 \le p_s \le 139530$ Pa	(5.4.h)
$(1-\varepsilon)=6.257\times10^{-3}p_s^{0.2899}$	$p_s \ge 139530$ Pa	(5.4.i)

5.6 SINGLE TUBE PILOT PLANT TESTS

When the concept of the vertical tubular filter was suggested, the design sub-committee recognised the need to perform pilot-scale studies to establish whether the idea and configuration of the filter was feasible. A single tube pilot plant was constructed at the Umgeni Water Process Evaluation Facility to perform the initial investigation. Single tube material brand named *Swiss Silk* was obtained for the initial tests. Once a more durable fabric, manufactured by Gelvanor Textiles was available, a section of a woven curtain was installed on the single tube pilot plant.

5.6.1 Feasibility of Vertical Tubular Filtration

The pilot plant was operated using a *Swiss Silk* material (tube diameter of 60 mm). This type of material had previously been used in studies on the Tubular Filter Press, and was found to be susceptible to splitting under pressure after a period of operation. In an attempt to strengthen the material two tubes were assembled inside each other. During this period there was unfortunately very little sludge available from the Wiggins Waterworks and a mixture of this sludge and sludge from the Durban Heights Waterworks was used for the pilot-plant tests. The following observations were made :

- Water works sludge can be dewatered using a vertical tube.
- The solids concentration can be improved to between 16 and 30 % m/m.
- Most of the sludge can be removed without a roller (This was variable depending on the type of sludge used, and the amount of solids removed during the run).
- In certain instances and for more difficult sludges, a roller is required to achieve better cleaning.

The tube made of *Swiss Silk* eventually ruptured under excess pressure, which was expected.

5.6.2 Process Investigation Using the Single Tube Pilot Plant

A sample of new curtain material was obtained from Gelvenor, and used for subsequent pilot-plant tests. Since the commissioning of the Inanda Dam - Wiggins tunnel in August 1994, the raw water turbidity at Wiggins Waterworks has been consistently below 10 NTU (except during

December, January and February 1996 where excessive rainfall was experienced), and bentonite was occasionally added as a coagulant aid. Previous operation of the Tubular Filter Press at the H.D Hill Waterworks showed that the dewatering of sludges is more difficult when bentonite is used in the plant.

Experience gained in the operation of the Tubular Filter Press indicated that the main process parameters which affect the formation of cake and the cleaning of the tubes are the feed solids concentration, the filtration pressure, and the limit to which the permeate flux is allowed to decline during constant pressure filtration. Waterworks sludge is highly compressible, and these process parameters together with the characteristics of the sludge affect the filterability and recovery of the solids.

5.6.2.1 Effect of Feed Solids Concentration

Using the new fabric, supplied by Gelvanor, the filter was operated at a constant feed pressure of 300 kPa, initiating a wash when the permeate flux decreased to 45 or 60 l/m²h. The cake was collected by flushing four times and then during the fifth flush a roller was used to ensure complete recovery of the solids. The percentage of sludge recovered by each subsequent flush was then calculated.

Table 5.9 shows that by increasing the feed solids concentration, the mass of the cake collected during the cleaning increased, and that this occurred for a shorter run time. It was also noticed that better cleaning of the curtain can be obtained without the use of a roller when the feed solids concentration increases.

Table 5.9 The Effect of Feed Concentration on Filtration Performance (Pressure 300 kPa, final flux 50 L/m².h)

Feed Sludge Conc. (g/l)	Ave. Run Time (min)	Total Mass of Wet Cake (g)	Cake Removed During Cleaning (% of the Cake Recovered)					Solid Conc. in Cake (%m/m)	Ave. Cake Recovery (%)
			Flush 1	Flush 2	Flush 3	Flush 4	Roller		
2 - 3	120	350	24	17	10	3	46	18,5	25
6 - 8	70	610	32	64	2	0	2	17,9	31
10	60	900	50	42	6	0	2	17,7	38

5.6.2.2 Effect of Filtration Pressure

By increasing the feed pressure to 400 kPa, and operating the filter until the same final permeate flux of 45 l/m^2h was obtained, the filter run time decreased **Table 5.10**. By operating at a higher pressure the solids are deposited and compressed more rapidly and to a greater degree, resulting in a higher average cake resistance. As less cake was deposited before the final flux was obtained, and as the recovery is proportional to the mass of solids deposited in the tubes during the filtration, correspondingly lower recoveries were obtained.

Although at a higher operating pressure, a higher recovery is expected for the same mass of solids deposited in the tubes, in this case the same limiting flux results in a lower mass of solids in the tubes and therefore a poorer removal by flushing and a lower recovery. A roller was required to remove 13% of the filtered solids.

Table 5.10 The Effect of Filtration Pressure on Filtration Performance (Final flux 45 - 50 L/m2.h)

Filtration Pressure (kPa)	Feed Sludge Conc. (g/l)	Ave. Run Time (min)	Total Mass of Wet Cake (g)	Cake Removed During Cleaning (% of the Cake Recovered)					Solid Conc. in Cake (%m/m)	Ave. Cake Recovery (%)
				Flush 1	Flush 2	Flush 3	Flush 4	Roller		
300	115	60	853	84	10	4	0	2	169	40
400	134	50	630	44	40	2	1	13	155	27

5.6.2.3 Effect of Final Permeate Flowrate Before Cleaning

In an attempt to improve the cake recovery, the pilot plant was operated until a permeate flux of 27 l/m^2h was obtained. The tests were performed at an operation pressure of 400 kPa. As the run time is longer the mass of solids filtered increased. The increased operation in the cake consolidation phase results in a more stable cake, and the cake recovery improved.

Table 5.11 shows that under these operating conditions the cake was predominantly removed by flushing and the benefit of using a roller was minimal.

Table 5.11 The Effect of Reducing the Final Permeate Flowrate before Cleaning

Permeate Flux (l/m²hr)	Feed Sludge Conc. (g/l)	Ave. Run Time (min)	Total Mass of Wet Cake (g)	Cake Removed During Cleaning (% of the Cake Recovered)					Solid Conc. in Cake (%m/m)	Ave. Cake Recovery (%)
				Flush 1	Flush 2	Flush 3	Flush 4	Roller		
45	134	50	630	44	40	2	1	13	155	27
28	146	100	1,280	72	21	2	4	1	176	50
27	141	110	1,360	3	93	1	3	0	169	53

5.6.2.4 Mass of Solids Removed per Unit Filtration Area

All the tests performed on the single tube pilot plant indicated that the amount of solids deposited inside the tube during a filter run, may be critical to improving the recovery of solids and the removal of the dewatered solids during flushing. The mass of solids recovered during each test (by flushing and the use of a roller) is compared to the total mass of cake removed during flushing only. (i.e. without the use of the roller). **Figure 5.16** shows clearly that the roller is not necessary provided more than 3 kg of solids are recovered per square meter during each filtration cycle.

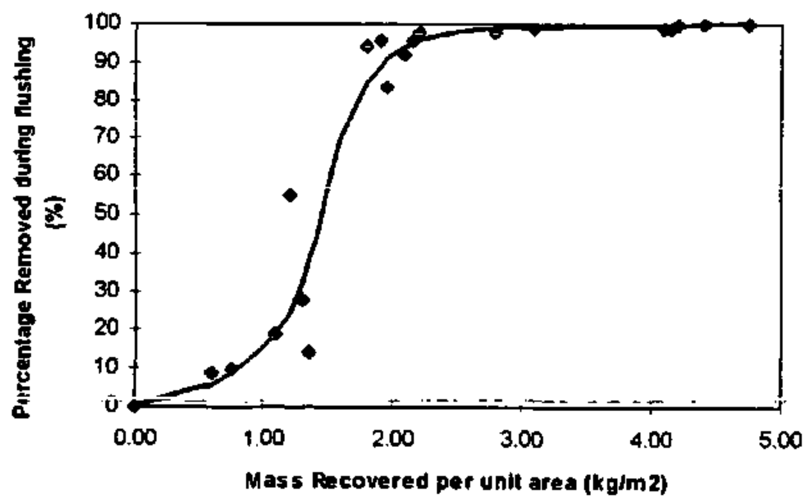

Figure 5.16 Removal of Solids Using Flushing Without Using a Roller to Assist Tube Cleaning

The process parameters investigated therefore require optimisation to provide a reliable process where the cleaning of the tubes can be achieved using a flush sequence only. **Tables 5.9, 5.10,** and **5.11** indicate that by increasing feed solids concentration, operating the plant at a slightly lower pressure so as to reduce the compression of the cake, and by increasing the filtration time and allowing the permeate flux to drop to lower values before cleaning, the solids deposition can be increased resulting in an increase in recovery.

5.7 OPERATION OF THE VERTICAL TUBULAR FILTER PRESS

Following final commissioning and the installation of the magnetic flowmeter, monitoring of the plant performance commenced on 12 January 1996. During the period January to May the plant was operated on an intermittent basis, and the total volume of sludge filtered was 137 m^3. The feed concentration of sludge fluctuated between 7 and 30 g/l.

5.7.1 Batch Operation to Decide on Operating Parameters

The unit was operated in a batch mode whereby feed sludge is pumped into the filter tubes in *a dead-end* mode of operation. During the filtration cycle a cake is formed on the inside of the filter tubes and is compressed under constant pressure delivered from the feed pump. The filtration rate drops as the cake thickness increases until the feed flowrate decreases to a predetermined value measured by the control system. At this point a wash cycle is initiated and the cake is discharged onto a conveyor and removed into a hopper.

There are three main factors (operating pressure, final filtration rate before cleaning, and feed solids concentration) which affect the performance of the unit for sludge dewatering, and have a direct impact on the final cake solids concentration and the cake recovery. The effect of varying the operating pressure and final flux was assessed. It was unfortunately not possible to control the feed solids concentration during this period. The Wiggins Waterworks was at the time undergoing major changes in the design and operation of the sludge treatment plant, and variations in the sludge concentrations were common.

5.7.1.1 The Effect of Filtration Pressure

The effect of filtration pressure was investigated by comparing the results obtained from plant operation at different operating pressures. Selected batch experiments were selected for comparison where the final flux and feed concentration were approximately the same.

Table 5.12 **The Effect of Filtration Pressure on Plant Performance**

Pressure (kPa)	Feed Conc. (g/l)	Final Flux (l/m^2h)	Cake Conc. (% m/m)	Cake Recovery (%)	Production Rate (kg/m^2.day)
210	11.5	50	19.8	52.3	18.3
210	14.3	70	20.2	38.8	22.5
300	12.5	50	20.3	48.4	18.3
300	13.7	70	26.6	36.1	19.8
300	10.7	50	27.4	73.3	19.2
300	12	50	25.3	70.6	21.7
350	8.7	50	31.2	69.1	14.9
350	10.8	50	31.8	73.1	17.9

Table 5.12 shows that by increasing the operating pressure from 210 to 300 kPa and from 300 to 350 kPa, an improvement in final cake concentration is observed. During the operation however, the increase in cake concentration did not result in a significant increase in cake recovery or production rate provided the limiting flux and feed solids concentration remained constant. If anything the production rate decreased slightly with an increase in pressure.

During the operation of the unit, there were no significant trends in cake concentration at different operating parameters. This is because the cake concentration can be affected directly by the pressure of operation (compressible characteristics of the sludge), or indirectly by cake consolidation equilibrium effects which in turn is affected by the feed solids concentration and the final flux. The cake concentration did however appear to generally increase with an increase in operating pressure. The nature of the sludge at Wiggins Waterworks is very dependant on the turbidity of the raw water, the amount of chemical addition (bentonite, polymeric coagulant, lime, sodium hypochlorite etc.), and the age of the sludge and thus will impact on the final cake concentration. In general the cake removed from the filter ranged between 20 and 32 % for a range of feed solids concentrations between 8 and 30 g/l, and limiting fluxes between 30 and 90 l/m^2.h.

5.7.1.2 The Effect of Feed Solids Concentration

The effect of feed solids concentration was compared from data obtained during operation at a pressure of 300 kPa. **Figures 5.17 and 5.18** show how the production rate and cake recovery are influenced by the feed solids concentration.

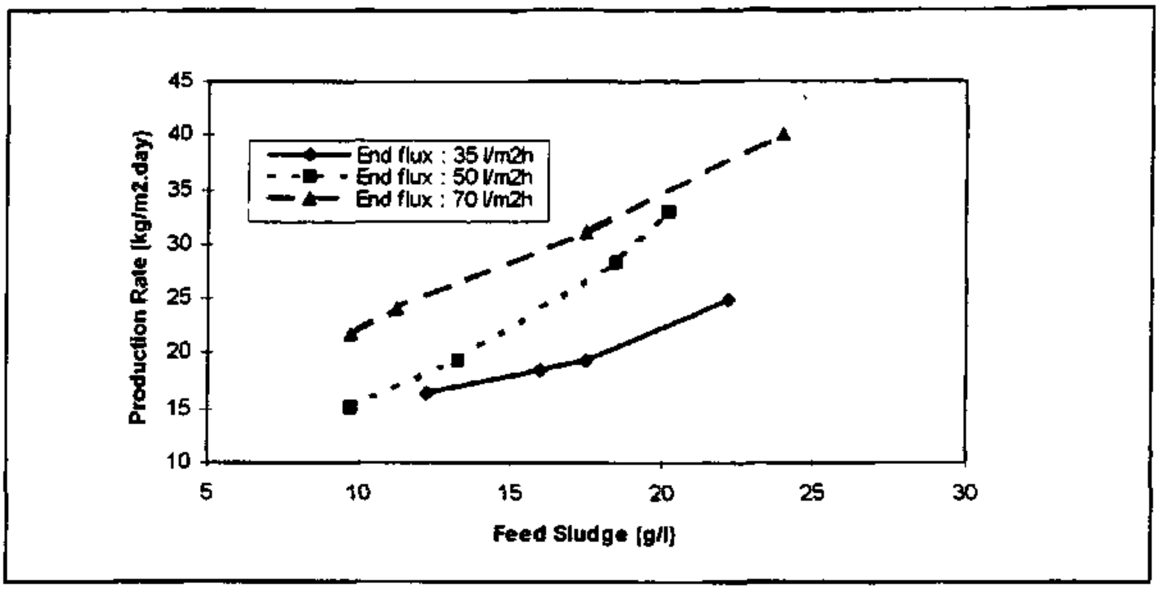

Figure 5.17 **Effect of Feed Concentration on the Dewatering Capability of the Tubular Filter Press**

It is reasonable that the dewatering capability of the filter (production rate) will increase with an increase in feed solids concentration as less permeate needs to be removed per mass of solids deposited in the tubes.

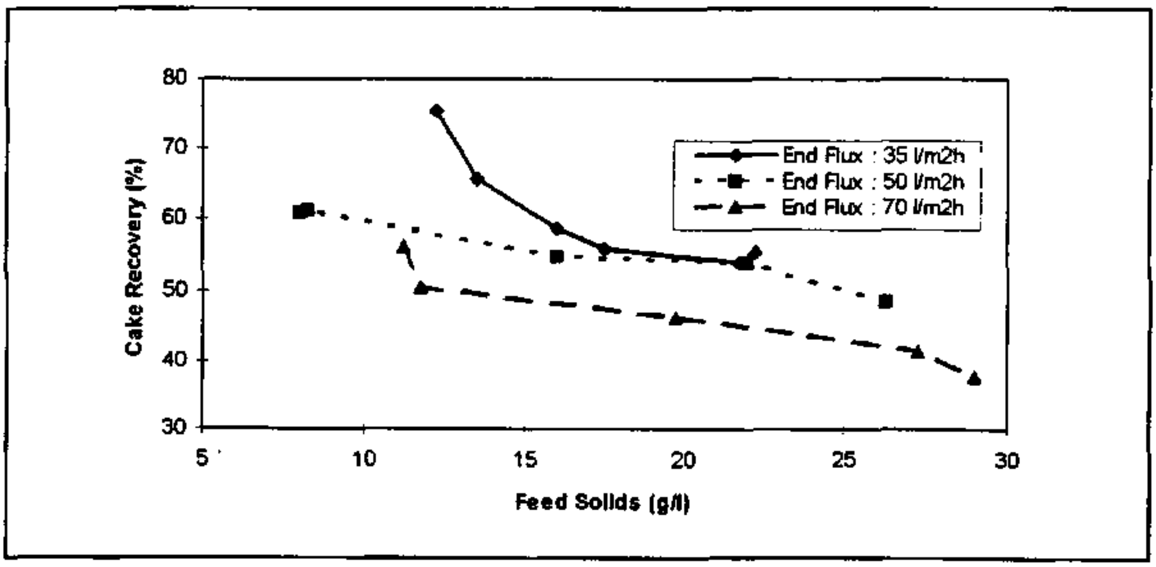

Figure 5.18 **Effect of Feed Concentration on the Cake Recovery**

During filtration the flush sequence is initiated at a specific final flux. Should the filtration be terminated at a higher final flux, the average filtration rate will increase for the continuous operation of the filter and the dewatering rate will increase.

The operation of the filter at a specific feed solids concentration, the recovery is expected to increase with a decreasing final flux as the filtration cycle will increase, a larger amount of solids will be deposited in the tube, and the cake will consolidate to a greater degree.

At the same end flux, theoretically the recovery should be constant as the mass of solids deposited should be the same. If equilibrium effects are significant, then as the feed solids concentration increases, the cake will be formed more rapidly and the mass of solids deposited will increase before the end flux is attained, but the filtration time will be reduced. Rencken (1992) found that the recovery increases with an increasing filtration time (and hence with an increase in solids deposited in the tubes). In this case however, the decrease in the filtration time is more significant than the increase in solids concentration or the mass of solids deposited, due to equilibrium effects and the time required for adequate consolidation of the cake, and the recovery was found to decrease.

5.7.1.3 The Effect of Final Permeate Flux Before Cleaning

The effect of the limiting flux on the performance of the unit is most significant. **Figures 5.17 and 5.18** show that a decrease in the final flux can significantly improve the cake recovery and negatively affect the production rate of the unit.

Figure 5.19 shows the decrease of flux during a run. During the initial stages of a run, the filtration rate is relatively high and as a cake is formed on the inside of the tubes in a gradual decrease in flux is observed. It was noticed that the longer the cake is exposed to pressure the easier it becomes to remove the cake, resulting in higher cake recoveries. If the filtration cycle is stopped when the flux is still relatively high, the sludge inside the tubes will not have effectively formed a cake, and will be washed back into the feed tank, adversely affecting the recovery.

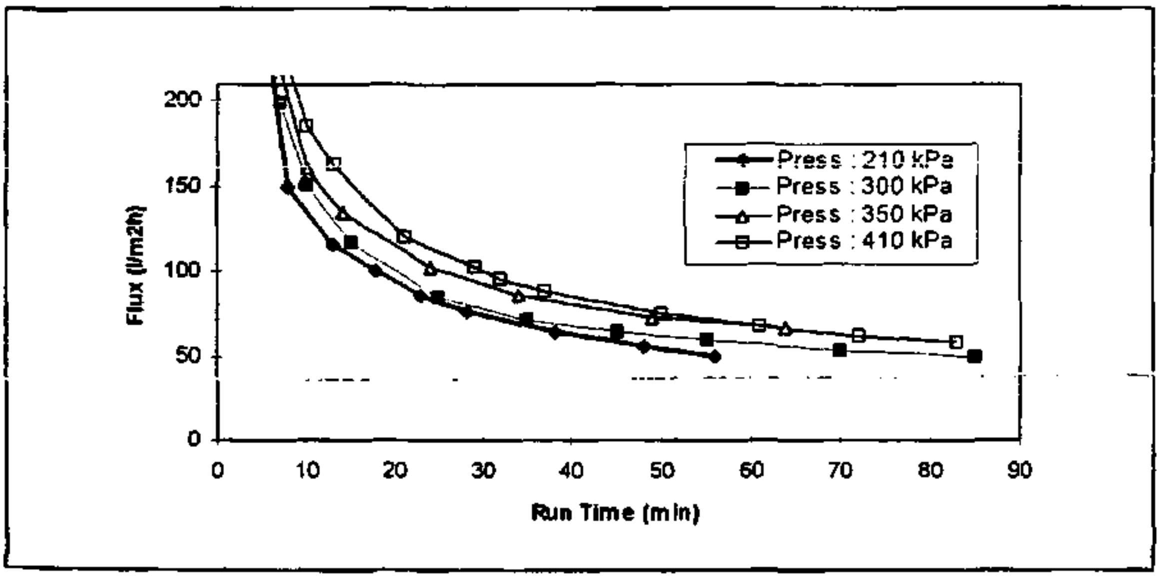

Figure 5.19 **Flux versus Time Curve for Sludge Filtration**

5.7.1.4 **Improving the Dewatering Rate of the Filter**

Figure 5.20 demonstrates how (at 300 kPa) the production rate can be more than doubled by reducing the run time between washes and increasing the end flux. Although this has a negative impact on the cake recovery per cycle, an increase in the feed solids concentration can compensate for any reduction in efficiency.

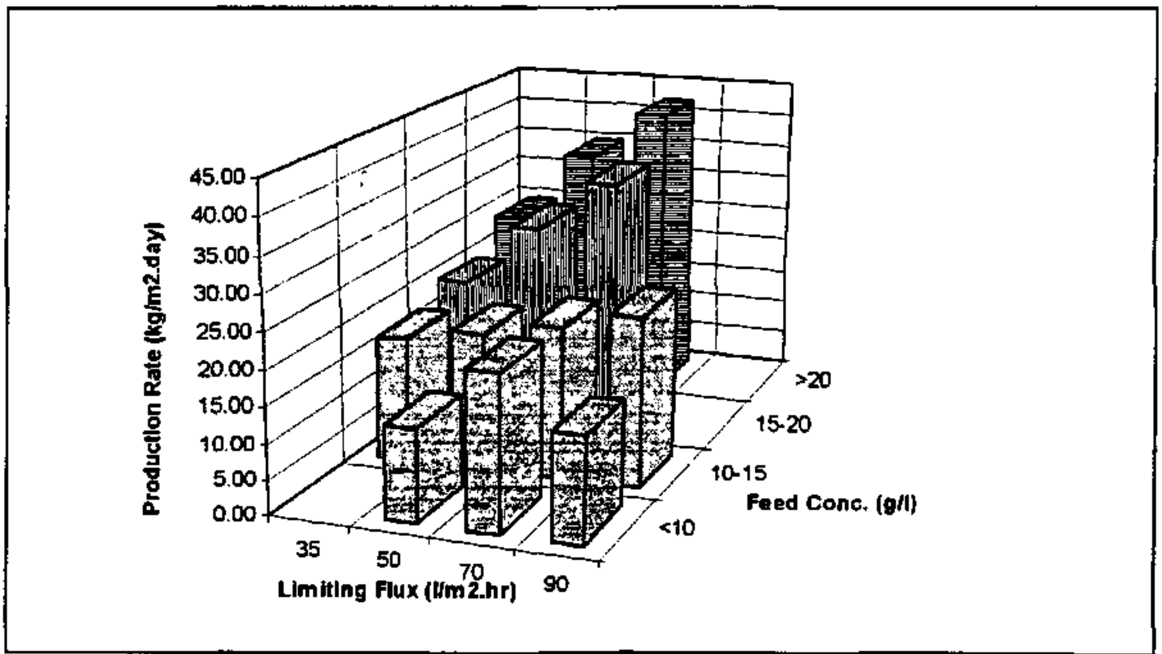

Figure 5.20 **Effect of Final Flux and Feed Concentration on the Dewatering Capability of the Filter (Production Rate)**

5.7.1.5 **The Efficiency of the Cleaning Sequence**

The effectiveness of the cleaning sequence was investigated. It was observed during the initial single tube experiments that most of the solids were removed from the tubes during flushing, and the roller was only removing a small fraction.

During the initial period of operation the washing sequence included two flushes followed by the action of rollers. The performance of the plant was estimated as it was at that stage difficult to collect all the cake removed from the tubes. Following modifications to the unit whereby the filtrate collection was separated from the sludge collecting bin, all the sludge could be weighed and plant recoveries calculated more accurately. In this way the sludge removed during flushing was separated from the sludge removed during the action of the rollers and the efficiency of the flushes was evaluated as a stand alone cleaning mechanism.

The filtration time (final flux) was found to be critical to the effectiveness of the flushing as a stand alone cleaning mechanism. At lower final fluxes, longer filtration times are common, the cake is pressurised for a longer period of time and greater consolidation is allowed to take place, the cake thickness is greater and hence a better removal is obtained. Under certain operating conditions, the use of a roller will still be required to effectively remove all the cake from the tubes.

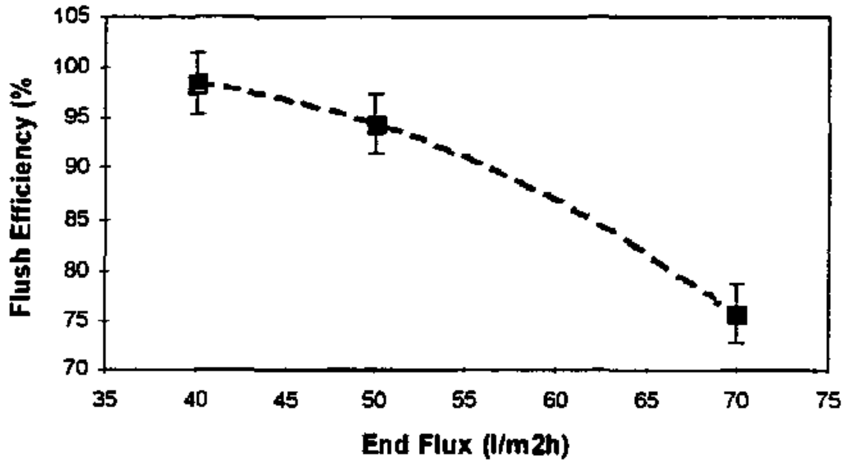

Figure 5.21 **Efficiency of Flushing (as opposed to the use of a Roller) versus Final Flux**

5.7.1.6 Cake Solids Concentration

During continual operation of the filter the final cake concentration was not constant. After each batch operation of the Vertical Tubular Filter Press, the solids removed on the conveyor was sampled.

The variability can be attributed to any of the following :

- Changes in the cake and sludge characteristics with time.
- Equilibrium effects due to changes in operating parameters (pressure, final flux).
- External factors such as flush cleaning velocity (which may remove thin layers of cake to varying degrees), the use of rollers or the number of flushes.
- Condition of the conveyor belt. It was found that after a period of operation the conveyor blocked, and the size of the aperture had to be changed to allow sufficient dewatering. Similarly a spray had to be installed to wash the conveyor during operation.

During December and January 1996, excessive rainfall was experienced in the Durban area, and the turbidity of the raw water at Wiggins Waterworks increased significantly during the period December to March 1996 **Figure 5.1**.

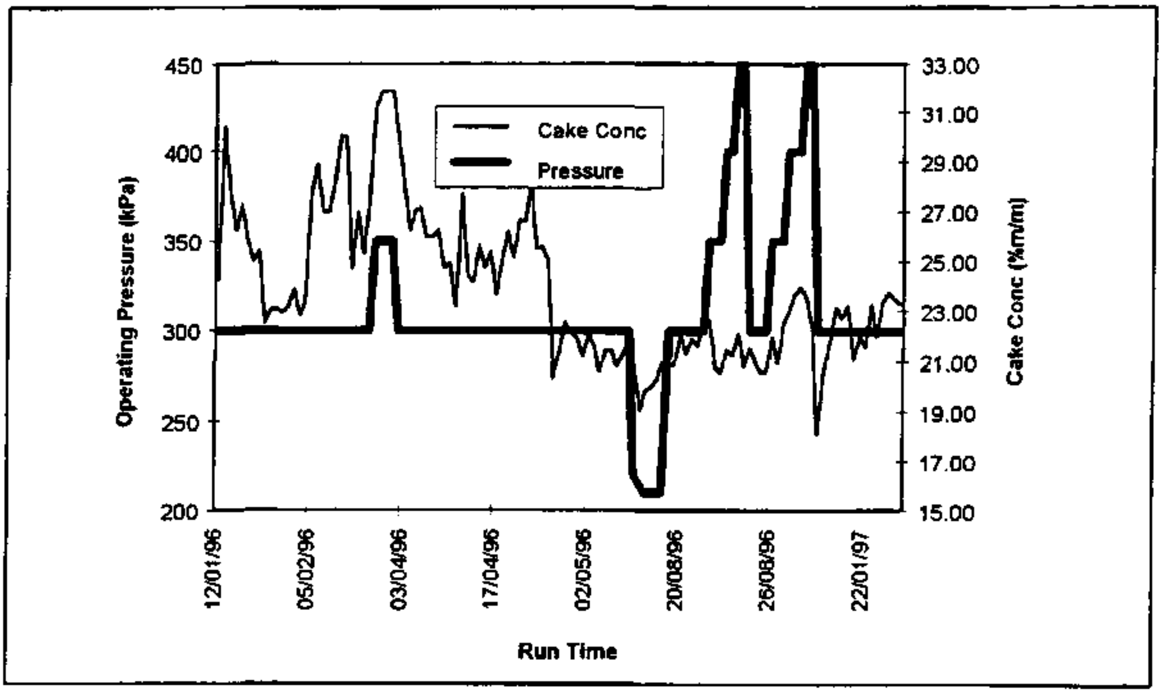

Figure 5.22 Variability of Final Cake Concentration and Operating Pressure During the Operation of the Vertical Tubular Filter Press

As discussed in **Section 5.1**, the nature of the solids in the raw water will significantly affect the performance of the tubular filter press. In April 1996 the turbidity of the raw water decreased to below 20 NTU, and the nature of the suspended particles changed and became more colloidal in nature. The performance of the tubular filter deteriorated at this time. This was also coincidentally the period when the problems with the conveyor blockages occurred. Nonetheless the Vertical Tubular Filter Press successfully produced cake concentration above 20 % (m/m).

5.7.1.7 Power Consumption of the Vertical Tubular Filter Press

During the operation of the filter, the power consumed per mass of solids was monitored. The power consumed is particularly dependant on the length of the filtration cycle, as the flush pump consumes a large amount of power. As discussed in previous sections, the feed solids concentration, operating pressure and other factors all contribute to the length of the filtration cycle. These also contribute to the recovery of the process. His is an extremely difficult parameter to assess and optimise, and is reported for the process operation irrespective of the operating parameters. **Figure 5.23** shows a typical power consumption of 50 kWh/ton of dry solids.

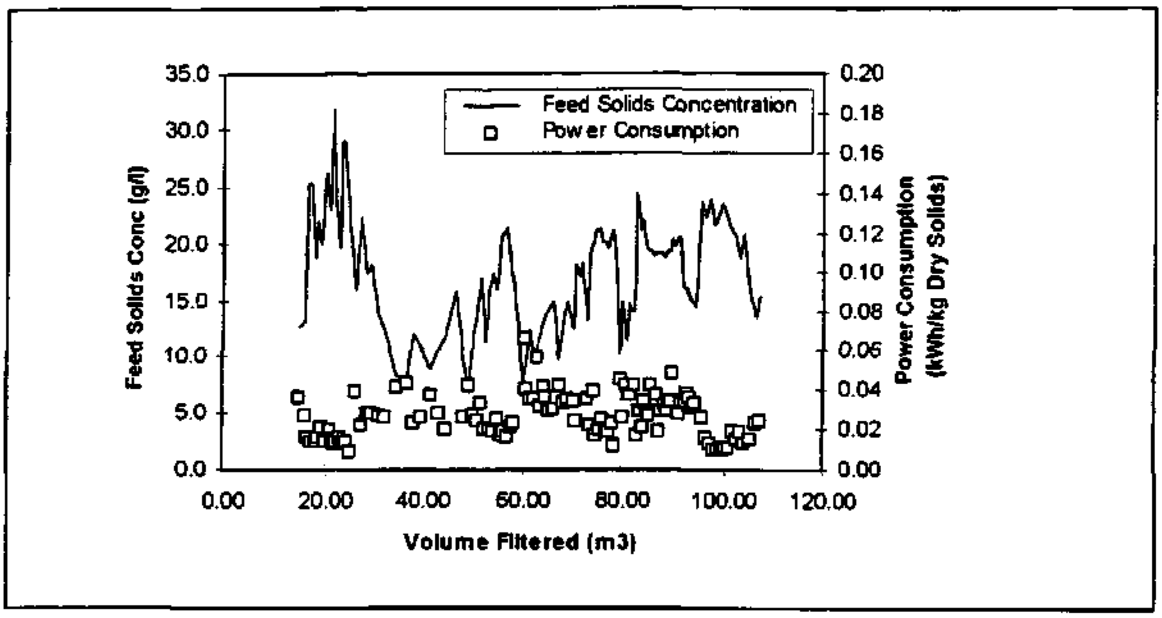

Figure 5.23 **Power consumption and Feed Solids Concentration for the Vertical Tubular Filter Press**

5.7.2 Continuous Operation of the Vertical Tubular Filter Press

The Vertical Tubular Filter Press was operated on a continuous basis to establish the efficiency and reliability of the process over a longer period of time. The plant had up until this stage been operated on a batch basis, and had shown that it was feasible to dewater sludge from the Wiggins Waterworks using the vertical tubular configuration.

The method by which the PLC program determines when a filtration cycle should be terminated and the tubes cleaned by flushing and the action of a roller is described in **Section 4.5**. Operating parameters (pressure, and final permeate flux) were initially set based on previous batch tests (300 kPa and 50 L/m^{2}h respectively). As with all continuous processes, plant operating conditions are never stable, and variations in the feed sludge concentration, as well as fluctuations in solids recovery will affect the operation of the filter.

During the filtration cycle, the level in the feed tank is maintained by a batch filling controlled by level probes. The concentration in the feed tank is typically higher than the concentration of feed sludge from the waterworks, and the concentration in the feed tank therefore fluctuates as the level in the feed tank is maintained.

(Feed Sludge Concentration 16 g/l, Pressure 300 kPa, Final flux 50 L/m²h)

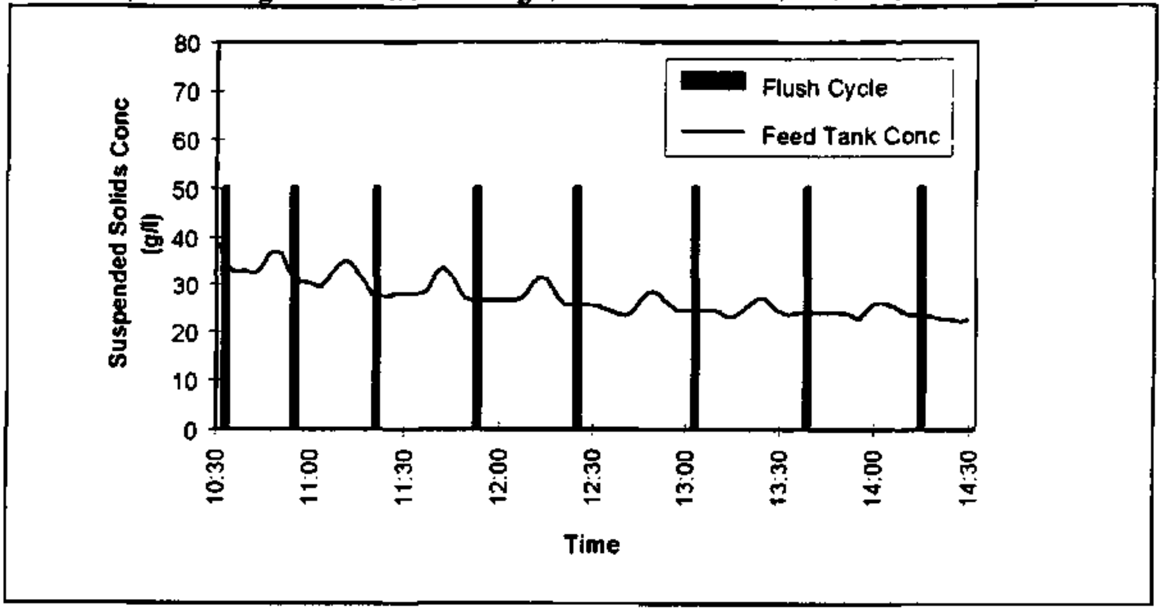

Figure 5.24 **Feed Tank Concentration and Flush Interval for Continuous Operation**

Figure 5.24 shows a slow dilution of the feed tank solids concentration during continuous operation as the feed sludge concentration was at that point lower than the concentration in the feed tank. The length of the filtration cycle time (i.e. the time it takes during each filtration cycle to reach the desired final permeate flux of 50 L/m2h), clearly increased during this period as the feed solids concentration decreased.

Due to cake losses during the cake removal cycle the concentration of the feed tank will increase after the cleaning cycle. If the recovery is poor, the feed tank concentration will continue to rise or decrease until steady state is reached where dilution on filling is offset by concentration of the tank during cleaning.

5.7.2.1 The Effects of Feed Solids Concentration on the Length of the Filtration Cycle

During the operation of the plant, dewatering difficulties were experienced after prolonged periods of operation. The recovery of solids appeared to decrease and the length of the cycle time was noticeably shorter. The concentration in the feed tank had increased during this period to levels in excess of 4% (m/m) as a result of the lower recoveries. In the period prior to the operation of the unit, bentonite had been used at Wiggins Waterworks, and this was believed to be the cause of the poor performance of the filter. It should be noted that as the sludge characteristics change, so do the optimal operating parameters, and the plant should be controlled in a stable operating regime characterised by good recoveries and longer filtration times. Previous operation of the Tubular Filter Press at H D Hill Waterworks indicated that dewatering problems could be solved by the addition of lime to the sludge.

Compression permeability cell tests showed that the addition of lime does improve the permeability and that the average specific cake resistance is reduced. A lime feeder was installed and lime was added at a rate of 5% lime per mass of solids in the sludge. The improvement in the performance of the filter was remarkable with immediate recovery of solids and longer filtration cycle times. The filter was operated using lime for a period of a day; followed by a period when the lime was not added, to establish whether intermittent use of lime could be used to correct operational difficulties.

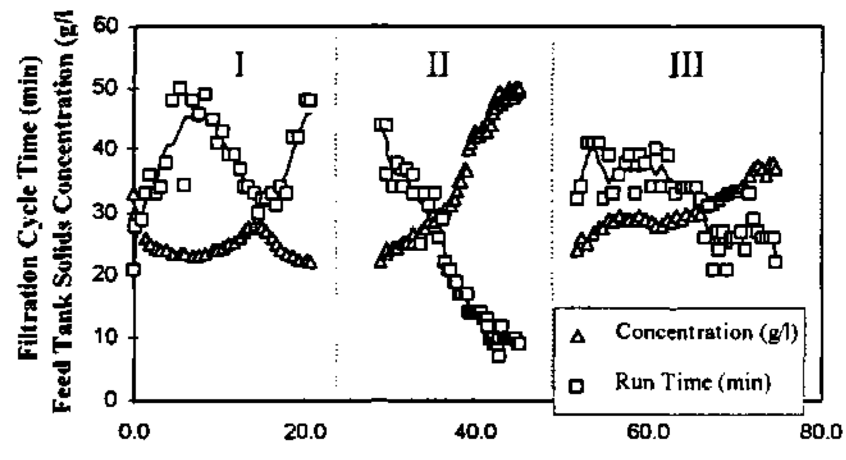

Figure 5.25 **Effect of Feed Solids Concentration on the Filtration Cycle Time**

Figure 5.25 shows how the filtration cycle time varies with a change in the feed tanks solids concentration. During *Periods I and III* **Figure 5.25**, lime was added and clearly the filtration cycle time was manageable and was maintained between 25 and 50 minutes allowing sufficient time for compression of the cake and thereby improving the recovery of the solids. When the lime addition was stopped (*Period II*) the recovery dropped significantly causing an increase in the feed tank solids concentration, and a subsequent shortening of the filtration cycle time to below 10 minutes. During these periods a highly resistive thin layer of cake is formed and due to a shortened filtration cycle the cake is not adequately compressed and the cleaning of the tubes was not effective.

The pH of the sludge increases to above pH 11 when lime is added, and the quality of the permeate tends to decrease during this period. It is therefore concluded that although the filtration is improved and higher recoveries are obtained, the precise influence of the lime has not been established.

Although the high pH and lime addition modifies the chemical nature of the bentonite to improve the filtration, problems with scaling and precipitation of calcium carbonate in the fabric may be resulting in poor initial fluxes. This type of fouling cannot be reduced merely by improving the internal cleaning of the fabric, and other methods such as acid washing or removal of the curtains may need to be considered.

The pH of the permeate obtained during filtration using lime was consistently above 10, which during normal solids dewatering would be returned to the head of the works for reprocessing. This would have a severe impact on the pH of the raw water, and although the amount of pre-lime could be reduced, acid may be required to maintain stable pH for water treatment. This complication is a definite disadvantage, and the use of lime should be avoided.

During the process operation a number of methods were tested to prevent the solids concentration in the feed tank from increasing to such an extent that the performance of the filter is affected. Regular emptying of the feed tank proved useful, but once the run time was less than 10 minutes this did not significantly improve the performance. It was noticed however that when the tubes had been left to dry over a weekend and then flushed again on a start-up of the plant, the performance was improved and much longer filtration cycle times were evident. This indicated a thin resistive layer of cake is remaining inside the tubes and is not being removed by conventional flushing.

5.8 QUALITY OF THE FILTRATE

During the operation of the single tube pilot plant as well as the Vertical Tubular Filter Press, the filtrate was of an excellent quality. In most instances the turbidity of the water was below 0,5 NTU, and only during initial cake formation or when pin holing of the filter (fabric failure) occurred, did the turbidity exceed 1 NTU.

The excellent turbidity reduction by the filter while dewatering sludges is due mainly to the compressed solids layer inside the woven fabric forming an extremely fine membrane for particle rejection. When filtering low turbidity raw water the reduction was not as consistent.

5.8.1 Production of Potable Water

The single tube pilot plant was operated in *dead-end* mode for the production of potable water. The procedure of operation was similar to that of the Crossflow Microfilter. A precoat of approximately 100 to 150 g/m^2 of limestone was applied to the vertical tube. The precoat solution was allowed to circulate for 10 minutes before the introduction of raw water. **Figure 5.26** shows a decline in the filtration flux with time at an operating pressure of 300 kPa. This is compared to a similar operation of the Crossflow Microfilter operating at 400 kPa. The filtered water turbidity decreased with time of operation, but was in most cases below 1 NTU.

An attempt was made to repeat the test at a higher operating pressure, but small holes developed in the filter cloth material along the seam of the tube. This resulted in poorer filtered water turbidities and higher flowrates. The holes were not observed during sludge dewatering as a filter cake inside the tubes reduces the risk of pinhole development.

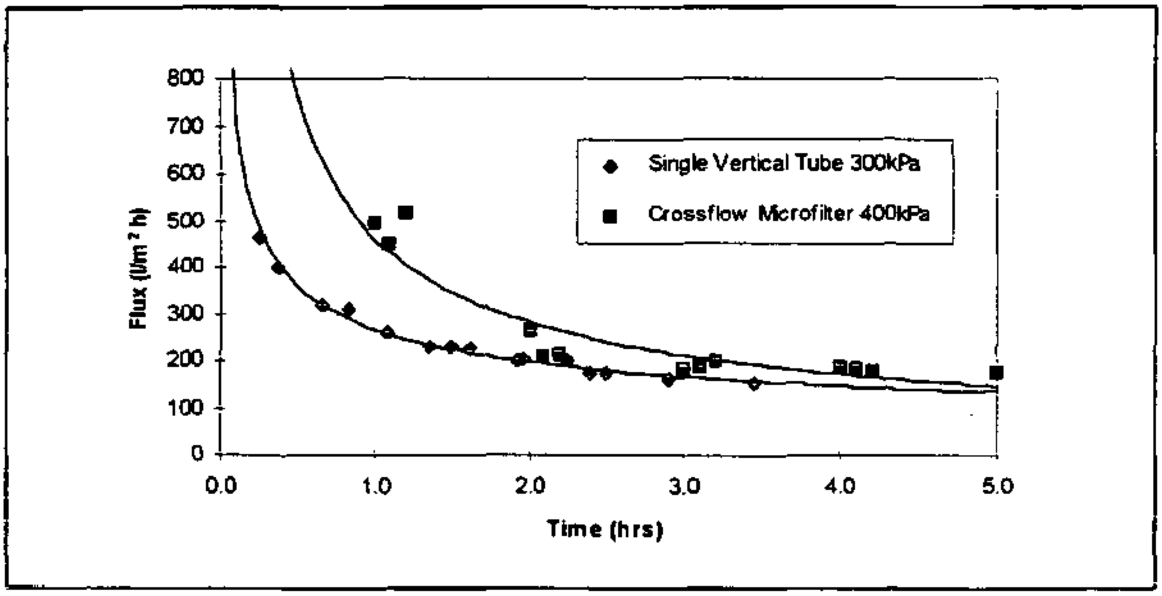

Figure 5.26 **Comparison of Flux Decline for the Crossflow Microfilter and the Vertical Tubular Filter Press**

Further experiments operating the single-tube pilot plant at a pressure of 200 kPa were conducted to establish the turbidity of the filtered water and the microbiological rejection of the filter for potable water production. **Figure 5.27** shows the raw water and filtered water turbidities during the operation of the filtration cycle treating raw waters between 8 and 20 NTU. When a pre-coat of limestone was used the permeate turbidity was reduced to between 1 and 2 NTU. The turbidity at the beginning of the filtration cycle after the completion of the pre-coat was at times significantly higher than the raw water turbidity, and are significantly higher than turbidities obtained using crossflow microfiltration, as the effectiveness of pre-coating the filter durting *dead-end* *filtration* is not as effective as the deposition of a controlled layer of pre-coat during crossflow operation.

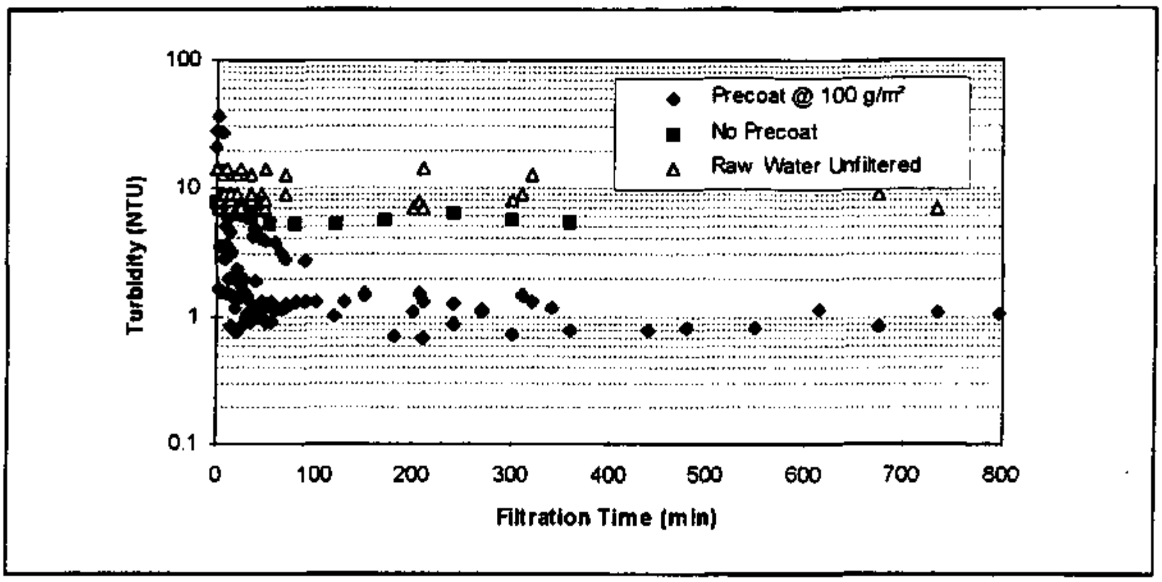

Figure 5.27 **Raw and Filtered Water Turbidities Using the Vertical Tubular Filter Press With and Without Precoat of Limestone**

During the operation of the filter no solids were collected and removed by this method of operation. Should the unit be used for raw water treatment, the use of crossflow microfiltration is likely to be more efficient for the production of potable water.

5.9 COMPUTER SOFTWARE DEVELOPMENT

A comprehensive mathematical model for constant pressure compressible cake filtration has been developed. In this form it is inaccessible and therefore cannot be utilised. The calculation framework required for the model is complex and rigorous and has to be performed by computer. Although it may be possible to perform the calculations on a spreadsheet, this would be cumbersome and tedious. The mathematics of the model are too complex and iterative to be able to solve effectively, and specific computer software has to be developed to enable these methods to be used effectively in practise. The model is to be made accessible in the form of user-friendly, Windows based programme that can be used by anyone without understanding the complexities of the model or calculation framework employed.

The development of the software has not been completed, and motivation to complete this work was submitted to the Water Research Commission. A WRC project K8/298, *The development of computer software to model constant pressure, compressible cake filtration* was approved and the software will be developed in two parts.

- A predictive computer program which can be used to facilitate the design and optimisation of constant pressure compressible cake filtration plants once the characteristics of the slurry have been determined, and
- A regressive computer program which can be used to regress on actual filtration data to obtain characteristics of the slurry which would otherwise be obtained by standard test procedures.

5.10 TECHNOLOGY TRANSFER

Umgeni Water has been working very closely with Explochem throughout the project, and together with the University of Natal the technology has been promoted. The single tube pilot plant has been extremely useful as a tool for performing trials on site, and has been beneficial in demonstrating the operation of the technology to prospective clients.

Explochem and Umgeni Water have held numerous discussions with regard to the installation of a large scale plant at the new Midmar Waterworks in Pietermaritzburg. The proposal was however not approved, but recommended for a smaller waterworks within Umgeni Water. A number of pilot scale trials were performed at the Hazelmere Waterworks (**Section 5.10.2**).

Dr G. Rencken and D. Mullan presented a poster at the WISA conference in Port Elizabeth in May 1996 entitled *Modelling of the Tubular Filter Press*, and once the modelling program has been completed, this will form part of a unique package for determining the characteristics of sludges requiring dewatering.

A model of the Vertical Tubular Filter Press was constructed and displayed during the trade exhibition at the AWWA conference on Toronto (June 1996). As the Vertical Tubular Filter Press is significantly different to the previous horizontal design, application has been made to patent the technology. A South African patent application was registered in August 1995, and prior to the AWWA conference in the USA, application was made in the United States to patent the new technology. A third patent application was also submitted in Korea to protect the novelty of the technology while negotiations were progressing for the application of this technology in filtering a calcium fluoride precipitate from an industrial effluent.

5.10.1 Sizing of a Vertical Tubular Filter Press for Umgeni Water

It was initially proposed that a tubular filter press be installed at the new Midmar Waterworks and operated as a trial unit by Explochem, where there would be additional dewatering capacity to manage the full solids loading of the works. A decision was made by the design team however not to install a prototype unit at the Midmar Waterworks but at an alternative location within Umgeni Water. The Hazelmere Waterworks was proposed as a suitable alternative, and an investigation was initiated to propose a suitable treatment system for the works.

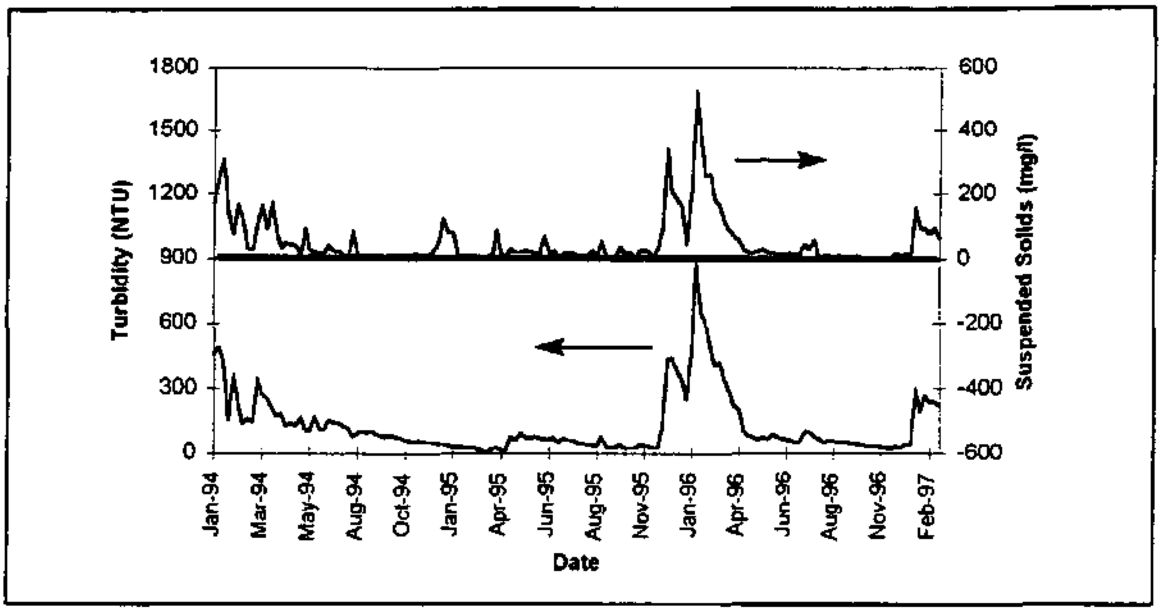

Figure 5.28 **Turbidity and Suspended Solids Analyses of the Hazelmere Waterworks Raw Water**

The single tube pilot plant was moved to the works, and a series of experiments were planned to obtain design and operating data for the sizing of the tubular filtration dewatering plant. The turbidity and suspended solids in the raw water **Figure 5.28** from Hazelmere dam has been monitored since 1994, and shows some periods in the year where a high solids loading can be expected. This is an advantage for using the vertical tubular filter as bentonite is not required at the works to assist with coagulation.

5.10.1.1 Single Tube Pilot Plant Trials

A single tube pilot plant was operated during April and May 1997 at Hazelmere waterworks to determine the dewatering capabilities of the process on sludge at Hazelmere. Batches of solids were thickened to different solids concentrations and filtered through a single tube. The solids concentration represents the steady state sump concentration which is normally higher than the concentration of solids in the feed to the plant. In each batch the recovery was calculated as being the amount of solids removed as dry cake expressed as a proportion of the solids pumped into the tubes during a filtration cycle. By mass balance the feed solids concentration was calculated based on the sump concentration and the measured recovery. This information is shown in **Figure 5.29**.

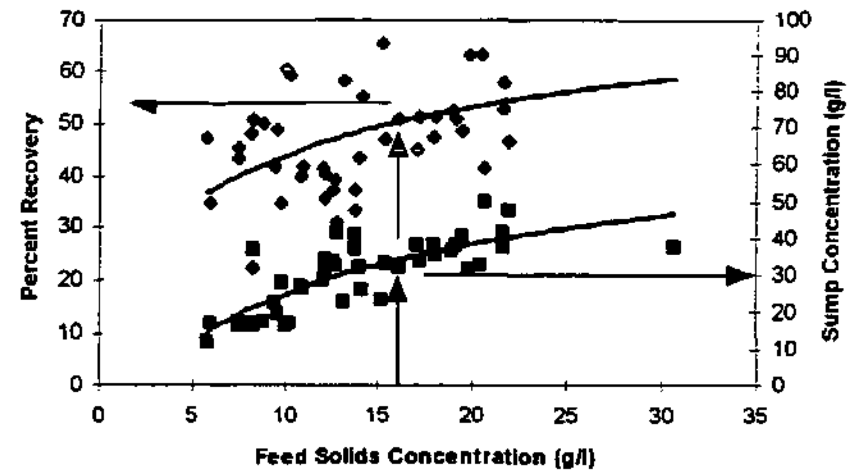

Figure 5.29 **Steady State Relationship Between Feed Solids Concentration, Sump Concentration and the Recovery**

The unit was operated at 300 kPa and at different limiting fluxes, and a dewatering rate was calculated per square meter of tube area. This is expressed as kilograms of solids per day. A conservative design region is shown for a feed solids concentration of 1.2% (12 g/l) when the plant is operated at a limiting flux of 50 $l/m^2.hr$, a dewatering rate of 40 $kg/m^2.day$ can be achieved.

Figure 5.30 shows that by changing the operating conditions and initiating the cleaning cycle at a limiting flux of 70 $l/m^2.hr$, the dewatering rate can be increased by approx. 50%.

During the pilot plant batch trials the final dewatered solids concentration was consistent in the range 25 to 30% dry solids (m/m). The cleaning efficiency was found to be excellent without the use of rollers, and dewatering problems at Hazelmere are likely to be few as no bentonite is used for water treatment.

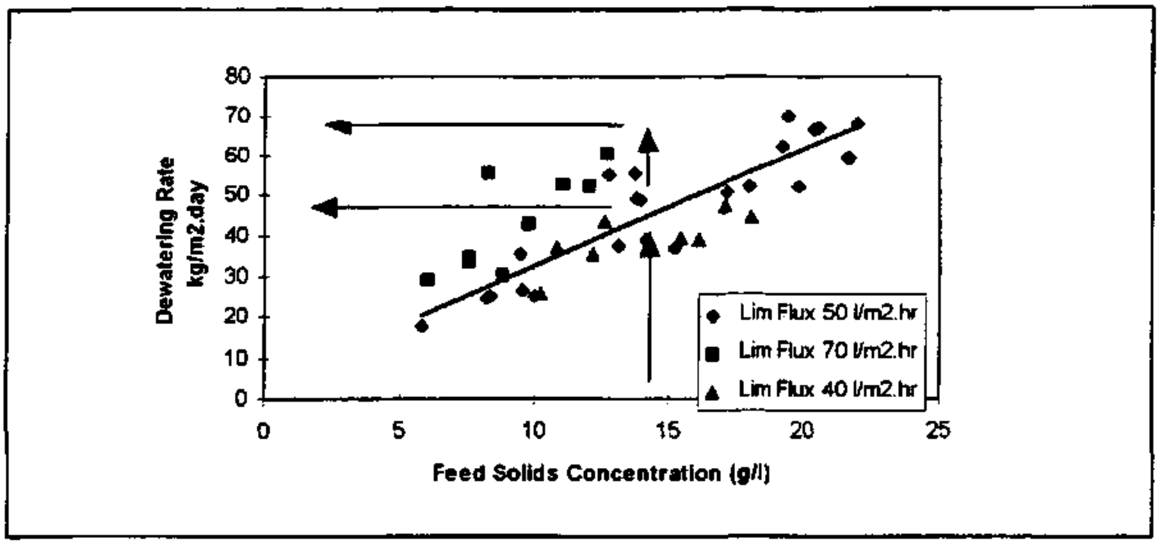

Figure 5.30 **Improvement in the Dewatering Rate of the Vertical Tubular Filter Press by Changing the Final Flux**

5.10.1.2 Sizing and Recommendations

The suspended solids concentration of the raw water to Hazelmere waterworks has been used for the design and sizing of the tubular filtration dewatering plant. It is common to use a 95 percentile (5% exceedence limit) as a design criteria for water treatment installations at Umgeni Water. In this case in particular it has been noted that by changing the operating parameters, the sludge dewatering rate can be increased significantly. A 90 percentile (10 % exceedence) criteria could be used for the design sizing.

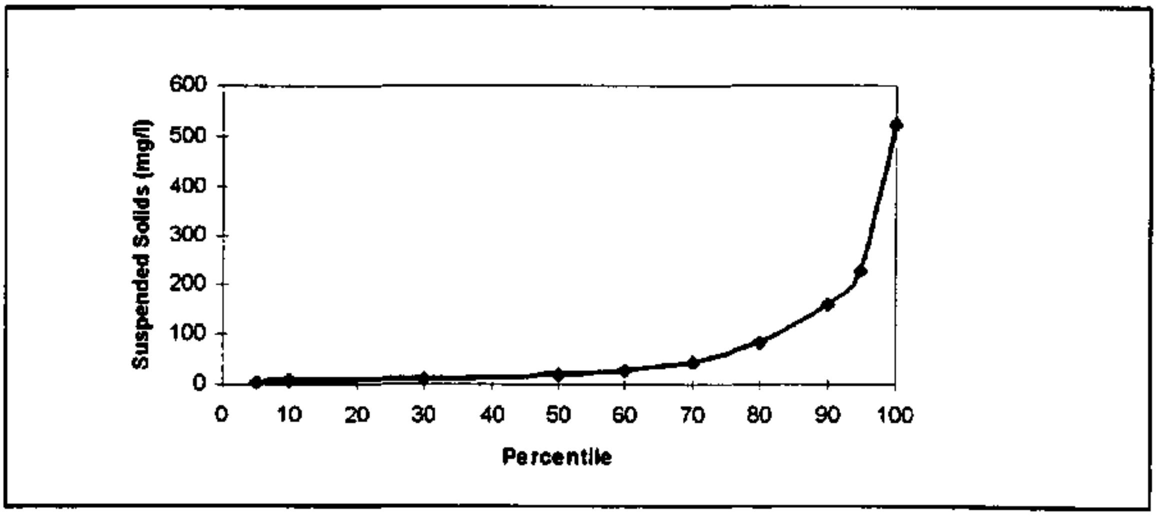

Figure 5.31 **Statistical Representation of Suspended Solids Concentration**

The design of the Vertical Tubular Filter Press proposed was the same configuration as the unit installed at David Whitehead in Tongaat, comprising 9 curtains per module. The table shows the sizing calculations for both a 5 and 10 % exceedence probability.

Table 5.13 Sizing of a Vertical Tubular Filter Press for Hazelmere Waterworks

	5 % exceedence	**10 % exceedence**
Design water flowrate at Hazelmere	45 Ml per day	45 Ml per day
90 percentile suspended solids	200 mg/l	160 mg/l
Design solids dewatering capacity	9 tons per day	7,2 tons per day
Pilot scale dewatering rate	40 kg/m^2.day	40 kg/m^2.day
Tube area required	225 m^2	180 m^2
Tube diameter	60 mm	60 mm
Tube length	2,5 m	2,5 m
No of tubes	477	382
No of curtains (14 tubes per curtain)	34	28
No of Modules (9 curtains per module)	4	4

The number of curtains in four modules (36 curtains) exceeds the required number based on a 5 % exceedence criteria, but the 28 curtains required based on 10 % exceedence is more than the number of curtains in only three modules. A design of four modules is therefore recommended which allows for additional solids contribution from the addition of lime and coagulant on the waterworks.

The benefit of installing the plant as a modular unit is that while the solids loading at the works is low, fewer units can be operated, and, while the works is operating below its design capacity, the installation can be programmed in phases to cater for the current solids loading, thereby reducing the initial capital cost.

6 Conclusions

1. The deficiencies in the operation of the the Tubular Filter Press at the Umgeni Water H.D. Hill Waterworks were identified and modifications to the design were proposed. A single-tube pilot plant was constructed and successfully operated in a new vertical configuration, demonstrating that the proposed design was feasible. The single-tube pilot plant was found to accurately represent the filtration and cake removal mechnisms, and can be used effectively to obtain design information for the sizing of dewatering applications.

2. The performance of the filter was reasonable and produced cake concentrations between 20 and 32 percent solids (m/m), at cake recoveries up to 75 %. The production rate was found to be dependant on filtration pressure, feed solids concentration and final permeate flux allowed before flushing.

3. The plant operation was found to be highly dependant on the sludge characteristics, not only with regard to the filtration cycle (cake formation) but also the flush cycle (cake removal). Cake recovery is a complex function of operating pressure, final permeate flux and feed solids concentration. In order to optimally operate a Tubular Filter Press not only should the filtration characteristics be known, but the recovery of solids *characteristic function* should also be determined.

4. It was found that under certain operating conditions, flushing without the use of a roller may be sufficient to effectively remove the cake from the tubes. This may not always be the case as the sludge characteristics were found to vary considerably during the operation of the plant.

5. Tube blockages (previously experienced at the H.D. Hill Waterworks) were completely eliminated by increasing the tube diameter to 60 mm and decreasing the tube length. The increased tube diameter did not result in any occurrences of tube splitting or failure using the fabric manufactured by Gelvanor Textiles. The removal of dewatered solids and conveyance out of the tubes improved, as the vertical orientation of the tubes assisted this by collapsing during the flush cycle.

6. The addition of lime to a waterworks sludge was found to improve the filterability of the sludge by altering the sludge characteristics. This was determined by compression-permeability cell tests, and was also evident during the continuous operation of the Vertical Tubular Filter Press. The addition of lime can have a negative impact on the plant operation as the pH of sludge and permeate increased significantly, and fouling of the woven tube fabric may occur due to the precipitation of calcium carbonate.

7. Compression-permeability cell (C-P cell) tests were performed to determine whether the standard C-P cell test was accurate in determining sludge characteristics or whether wall friction was significant for waterworks sludges. Although wall friction was observed the difference in the sludge characteristics obtained did not appear to be significant. This can only be fully ascertained once these parameters are incorporated into the model and the model prediction is compared to an operating system.

8. A new generalised *Area Contact Model* has been proposed for the constant pressure compressible cake filtration. Solution methodologies have been developed to regress for cake characteristics from operating plant data and to account for the period of pressurisation at the start of a filtration cycle. Once the software has been totally developed the accuracy of the *Area Contact Model* can be determined.

9. A single-tube pilot plant was operated in *dead-end* filtration to assess the use of the new vertical design in raw water filtration for the production of potable water. It was shown that a precoat of limestone is required to reduce the turbidity of the raw water to below 1 NTU, but the efficiency of pre-coating in *dead-end* mode (in a vertical tube) was poor. This process was clearly not adequate when compared to the Crossflow Microfiltration Process for potable water production.

7 Recommendations

1. It is recommended that this technology be actively marketed for dewatering of sludges where few problems are likely to occur. Areas to be avoided are organic effluents and highly variable industrial effluents where problems may arise as a result of inadequate information. To achieve this purchasing agreements need to be set up with the curtain supplier, and a small business could be initiated for the manufacture of curtains.

2. The Vertical Tubular Filter Press at the Umgeni Water Wiggins Waterworks should be operated on site, or moved to another site where it can be optimised and then operated on a continuous basis for an extended period of time in order to fully demonstrate and market the technology. The application should be specifically chosen where the sludge supply and characteristics are more consistent (more particulate in nature). Sufficient instrumentation must be installed (solids concentration meters, feed and permeate flowmeters and a measurement of the flushing velocity) to adequately monitor the continuous operation.

3. A limitation of the single tube pilot plant is that it can only be operated in a batch mode. It is also recommended that this be automated such that it can be utilised for process investigations of new applications and together with the computer model form a total design package.

4. Further investigations into the aspects of cake removal are recommended with the objective to develop an empirical recovery function based on the parameters that have been identified.

5. The software should be completed and fully evaluated before being developed into a marketable product that will compliment the technology.

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