

# **Deinking and Kraft Mill Sludge Dewatering Using a Laboratory Sludge Press**

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## ABSTRACT

A laboratory sludge press was used to investigate the dewatering of various wastewater treatment sludges generated in deinking mills and kraft pulp mills. The effects of pressure applied to the sludge, pressing time and temperature were studied in terms of their effects on sludge cake solids consistency, filtrate flowrate and filtrate suspended solids (TSS). The effects of various combinations of primary and secondary sludge were tested. Secondary sludge alone could not be dewatered in the press. Addition of primary sludge made pressing possible and increasing the ratio of primary to secondary sludge improved performance. Sawdust and hog fuel were shown to be effective filter aids. The use of single polymer flocculants was compared to dual polymer coagulant/flocculant additions for various sludges. The use of a coagulant in a dual polymer system didn't improve dewatering compared to using a single polymer flocculant in term of cake solids consistency and filtrate flowrate although it did reduce the total amount of flocculant necessary. But a dual polymer system did reduce filtrate TSS.  $\text{FeCl}_3$  and an inorganic polymer polyaluminum chloride (PAC) could be used to replace organic coagulants.  $\text{FeCl}_3$  is cheaper than organic coagulants, but PAC is more expensive. Measurement of sludge particle charge was shown to be useful in detecting polymer overdosing. The laboratory sludge press was shown to be an effective tool for evaluating the use of polymeric conditioners in sludge dewatering. This press can distinguish the two mechanisms by which sludge dewatering occurs in practice, filtration and cake compression.

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## **Chapter 1**

### **INTRODUCTION**

Both industrial and municipal biological wastewater treatment plants produce large amounts of waste sludge per year. For example, the pulp and paper industry in Canada produced approximately  $1.2 \times 10^6$  oven dry tons (odt) of sludge in 1994 and further increases are anticipated in the future. Typical sludge production is 0.03-0.1 odt of sludge/odt of pulp (Dorica, 1997). Deinking facilities produce two to four times more sludge than virgin pulp mills with the same capacity (Anon., 1995).

Sludge dewatering challenges in activated sludge plants include high operating costs, high loss of suspended solids in filtrates, limitations on ultimate disposal and others. Improvements in sludge dewatering performance should result in significant and beneficial results, such as reductions in the amounts of sludge dewatering chemicals needed and hence the cost thereof and landfill and incineration cost reductions.

Normally sludges from primary and secondary clarifiers have concentrations in the range 3%-5% and 0.5% to 1% respectively (Dorica, 1997). Sludge volume reduction is very desirable when sludge is to be transported somewhere for its final disposal. Dewatered sludge is generally easier to handle than thickened or liquid sludge. So sludge thickening and dewatering are indispensable steps prior to sludge disposal in landfills, in composting and/or by incineration. A sludge dewatering process is expected to produce a cake that has the solids content suitable for whatever final disposal option is chosen. Incineration typically requires sludge to have a minimum of 30% solids to be energy self-sufficient (Saunamaki, 1988). The lower the moisture content, the higher heating value. Landfills often require that a sludge have a regulated, minimum solids content; for example 25 % in Quebec (Dorica, 1997). Low solids content results in greater leachate

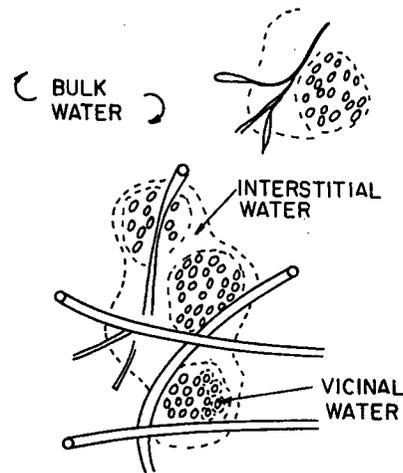
production at the landfill site because of the greater amount of water deposited at the site. For composting, dewatering can reduce the requirements for supplemental bulking agents or amendments. So large amounts of water have to be removed before final sludge disposal.

Primary sludges from pulp and paper industry mainly contain fibres, thus their production is dependent on fibre and other losses from the different departments of the mill. They are easily dewatered due to their high fibre content.

In response to increasingly stringent regulations on water discharges, most pulp and paper mills have had to adopt a biological wastewater treatment system. Unfortunately large volumes of secondary sludge are produced. Secondary sludge mainly contains biomass generated in the mill's wastewater treatment process. Due to its smaller particle size and the greater amounts of vicinal water (see below) associated with these microbial cells, it is more difficult to dewater. Since secondary sludges are more difficult to dewater than primary sludges, they are usually combined with primary sludges to enhance their dewaterability.

Water in sludge appears to exist in four forms (Vesilind, 1989): see Figure 1-1.

- (a) Free (or bulk) water is water that is not attached to sludge solids and that can be separated by simple gravitational settling.
- (b) Interstitial water is water that is trapped within the floc structure and travels with floc. This water can be released when the floc is broken up. Some interstitial water might be removed by mechanical dewatering devices.
- (c) Vicinal water is water that is associated with the solid sludge particles. This water is held on particles surfaces by virtue of the molecular structure of the water molecules and cannot be removed by mechanical means.
- (d) Water of hydration is water that is chemically bound to the particle and can be released only by the application of heat as in drying of the particles. This is sort of water found in inorganic hydrates, eg.  $\text{Al}_2\text{SO}_4 \cdot 6\text{H}_2\text{O}$ .



**Figure 1-1. Representation of a sludge floc ( Vesilind, 1989)**

Polymer addition (see below) reduces the amount of interstitial water by compressing the flocs, but does not change the surface area, and therefore does not change the amount of vicinal water (Vesilind 1994). Thus mechanical sludge dewatering methods have limits in that they cannot remove any more than bulk and interstitial water from the sludge matrix. Neither vicinal water nor water of hydration would be affected by these processes.

The relative amounts of these forms of sludge-water are most difficult to measure and not well understood. Such information is more likely to be used in research rather than in practice. But it still can help to explain some experimental results.

Some form of chemical conditioning is always implemented before combined sludge dewatering to achieve a satisfactory sludge dryness. Usually this is done via the use of polymers. The cost of these polymers is a large part of the total cost of the operation of the wastewater treatment plant. Polymer costs often exceed more than half of the operating cost of sludge handling (Abu-Orf, 1994, Vesilind 1979). Therefore, it is desirable to optimize the use of the polymers. Surveys of Canadian pulp and paper mills sludge dewatering practices (Dorica, 1999) indicate that some

mills utilize a two-stage polymer system, involving the addition of organic polymeric coagulants and flocculants, particularly mills with waste activated sludge. Some mills apply a single stage polymer system, involving only the addition of flocculant.

Chemicals selection is another problem in the chemical conditioning of the sludge. There are many kinds of chemicals available for sludge dewatering now, especially the synthetic organic polymers. Variations in the monomers used, molecular weight and charge density can result in an innumerable variety of polymers. It is difficult to determine which type of polymer is most suitable for a specific sludge. The best way of determining what to choose is to do a full scale test on site using the mill equipment and sludge, but this is costly and not practical if a lot of testing has to be done. So it is useful to develop a laboratory method to be used as a screening technique to select the best polymer(s) and determine its(their) optimal dosage(s). Then take the polymers and doses so selected to a mill trial.

Two parameters: specific resistance to filtration (SRF) and capillary suction time (CST) are commonly utilized in the evaluation of sludge filterability or dewaterability as well as optimizing polymer utilization (Pere, 1993; Lotito, 1993). But these two parameters have their own limits. One of them is whether or not they can really represent practical sludge dewatering procedures in which high pressure and high solids concentrations are utilized such as in the belt presses and screw presses that are typically used in industry. Another is that measurement of these parameters does not provide information on the final cake solids content nor on the pick-up and release characteristics from the filter medium. From these considerations it was decided to build a laboratory sludge press that would be able to quickly and accurately determine the dewaterability of a sludge with varied polymer types and doses (Groves, 1995). Since it can simulate sludge dewatering procedures in practice such as those found in belt presses and screw presses, its results should be reliable (Rehmat et al., 1997), however further information on its performance was thought to be desirable hence this thesis.

**OBJECTIVES**

1. To investigate the behaviour of the lab scale sludge press in terms of measuring the SRF, cake solids, filtrate solids and filtrate flow rate.
2. To use the lab press in studying the optimization of polymer utilization, including mixing conditions, dosage, etc.
3. To study the effects of pressure and pressing time on the sludge dewatering procedure.
4. To study the effect of temperature on the sludge dewatering procedure.
5. To study the effect of dewatering aids other than polymer, e.g. sawdust and hog fuel on the sludge dewatering procedure.
6. To study the dewatering behavior of sludges from different pulp and paper mills.

## **Chapter 2**

### **BACKGROUND AND LITERATURE REVIEW**

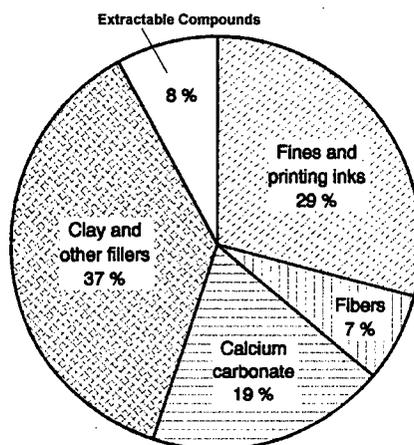
#### **2.1 Sludge Dewatering in Canadian Pulp and Paper Mills**

##### **2.1.1 Sludge Characteristics**

In the pulp and paper industry there are several types of sludge generated including primary, secondary and deinking sludge. The amount of difficult to dewater, secondary sludge has been steadily increasing over the past few years because more activated sludge plants have been installed to meet the revised regulatory limits concerning liquid discharges.

Pulp and paper primary sludges contain mainly fibres and are easily dewatered even without chemical additives. These sludges characteristically dewater well because the fibres provide low resistance channels for water to flow through. Mechanical pulping processes (TMP/CTMP/ground wood, etc.) generate sludges containing shorter fibres and large proportions of fines than chemical pulping processes. This makes mechanical pulping primary sludge more difficult to dewater (Dorica, 1997) than chemical pulping primary sludge.

Deinking sludge is generated from pulp mills which process recycled paper. The majority of the primary sludge in this process comes from the flotation skimmings and the whitewater clarifier rejects (Gilbert, 1991). Primary deinking sludges consist of fillers and coating pigments, fibers, fiber fines and printing inks, and adhesive components. Figure 2-1 shows that the average contents of a deinking sludge from the production of wood-containing graphic papers (such as newsprint and magazine grades). The compounds extractable with methylene chloride contain wood components from the fibers such as resins, fats and resin acids, soluble printing ink and adhesive components (Gottsching, 2000). Fines in sludges make dewatering difficult as they reduce filterability by plugging the filter media. The primary to secondary sludge ratio is usually high for deinking mills, due to the presence of fillers and coatings released from the waste paper furnish.



**Figure 2-1. Composition of deinking sludges from the production of wood-containing graphic papers related to dry substances**

Printing inks contain a liquid vehicle or solvent, polymers, resins, binders and solid pigment particles (TAPPI, 1997). These components may get into the primary sludge. In the deinking process, at the flotation clarification step, some coagulants such as polyamine, bentonite clay, alum, and flocculants such as polyacrylamide are added. So there may be a residue of them in the sludge. In fact, since they are similar to the polymers added to sludge to improve its dewaterability, any such residual polymers could be helpful for sludge dewatering. Deinking mill sludges normally have high primary to secondary ratio. As already said primary sludge is much easier to dewater than secondary sludge. This is why the deinking sludge dewatering process often achieves higher cake solids consistency and needs less polymer dosage than conventional pulp mill, combined primary and secondary sludge.

The activated sludge wastewater treatment process is primarily used to remove biochemical oxygen demand (BOD) from mill wastewater. This process relies on microbial oxidation of the oxidizable matter that causes BOD. In so doing the microbial population increases its mass as it consumes BOD. Some of this has to be wasted from the process and this constitutes secondary sludge, which consists mainly of flocculated microorganisms which have settled out in the secondary clarifier. See the flow sheet in Figure 2-2 below. The quantities of secondary sludge produced depend on the metabolic growth rates of the microorganisms present in the treatment system and on the level of dissolved oxygen available to the microorganisms. Secondary sludge is known to be more difficult to dewater than primary sludge due to its smaller particle sizes,

more vicinal water and dissolved high molecular weight material which tend to decrease the permeability of the pore system and to increase the consumption of conditioning agents (Eriksson, 1993). Another reason for this difficulty in dewatering is that the compressibility of the secondary sludge is so high that a thin layer, which accounts for nearly all the hydraulic pressure drop across the filter cake, is likely to be formed just above the filter medium (Sorensen, 1993). In the pulp and paper industry this disadvantage can be somewhat overcome by combining the secondary sludge with primary sludge which contains fibres that make for a filter cake that is less resistant to the flow of filtrate.

### **2.1.2 Sludge Dewatering Practice**

Although the specific features of the sludge dewatering processes might differ from one mill to another, the general lay-out is similar at most mills. A typical dewatering sequence for mills with activated sludge plants is illustrated in Fig. 2-2 (Dorica, 1999).

Primary and biological sludges are typically mixed in a sludge blend tank prior to dewatering. As the combined sludge is withdrawn from the blend tank, a coagulant is added to the pipeline. The high turbulence at this location assures a fast distribution of coagulants, to allow for a uniform neutralization of surface charges on the sludge particles. Then the flocculant is added in front of a static mixer to cause the neutralized sludge particles to bridge together into large flocs. Floc development and stabilization are completed in the flocculation tank. Some mills use only flocculant (no coagulant), achieving charge neutralization and bridging in one step. Before pressing, the conditioned sludge is thickened to 3 % to 10 % solids content, using rotating screen thickeners or gravity tables. Screw presses and belt presses are typically used in the last stage of the dewatering process to produce sludge cake suitable for disposal.

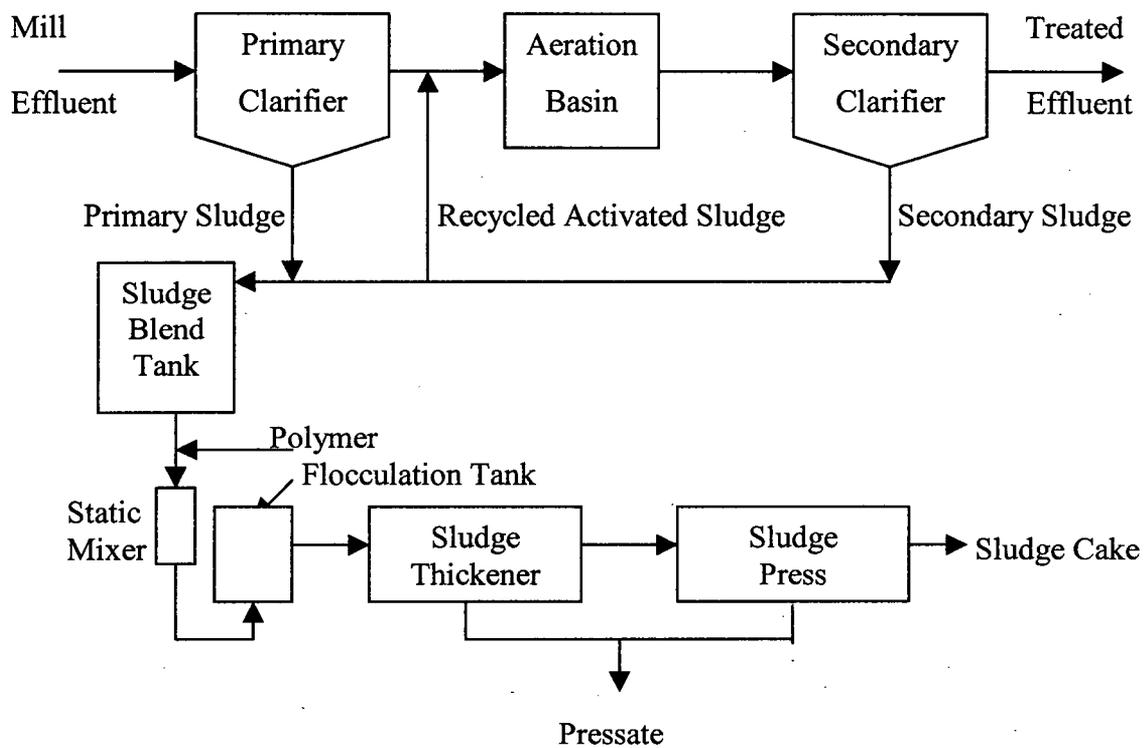


Figure 2-2 Typical combined sludge treatment system in pulp and paper mills

### 2.1.3. Thickening Equipment

#### 2.1.3.1. Rotary Screen Thickeners (RST)

RSTs are widely used for thickening fibrous sludges. The force of gravity combined with filtration, are the principal separation mechanisms of an RST. An RST is capable of increasing sludge consistency to 4-10% depending upon the percentage of secondary sludge and the percentage of solids from the secondary and primary clarifier (Kenny, 1997). Chemical conditioning of the sludge before passage through the RST is normally required.

### **2.1.3.2.Gravity Thickeners**

A gravity thickener is a radial clarifier that is normally 11-12 metres in diameter and 3 to 4 metres in depth. The advantages of a gravity thickener are its simplicity, low operating costs, low level of operator attention and a degree of sludge storage. Conditioning chemicals are not normally required and there is minimal power consumption. The disadvantages are septicity, odour generation, low dewatering capacity and large space requirements.

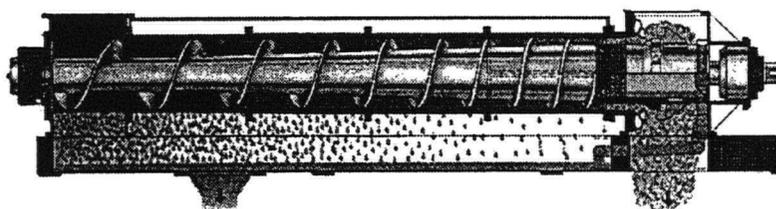
### **2.3.1.3. Gravity Belt Thickener**

The gravity belt thickener is also called a gravity table. It is based on the principle of separation of free water from the sludge through gravity drainage along a porous horizontal belt. As with the RST, chemical conditioning of the sludge is normally required to create a separation of solids from the free water. Plows evenly distribute the sludge onto the belt and create furrows to allow water drainage through the belt. After the thickened sludge is removed, the belt travels through a wash cycle. Gravity tables produce similar sludge consistencies to those of an RST.

There are other types of pre-dewatering technologies in occasional use. They are dissolved air flotation (DAF) clarifiers and belt presses. DAFs are effective with secondary sludge.

### **2.1.4.Dewatering Equipment**

#### **2.1.4.1 Screw Press**



**Figure 2-3 Screw press**

Screw presses have been used to dewater pulp and paper mill sludges having primary to secondary ratios of from 100:0 to 25:75 (Dorica, 1997). In combination with a thickener, screw presses can produce cakes having 30 % - 67 % solids. The press consists of a screw rotating in a stationary perforated screen barrel. The screw may have a fixed or varying pitch angle, and a worm channel is formed by the screw thread. The channel is tapered so that there is a gradual reduction in channel depth towards the discharge end. See Figure 2-3 (Egenes, 1992). In a screw press the sludge is progressively pushed through the stationary barrel of the press by the screw. Filtrate is forced out through the circumferential screens mounted on the barrel. Typically, the perforated screen has larger holes at the inlet end and smaller holes at the discharge end. Most screw presses are equipped with pressure cones to hold back sludge and increase its final dryness.

The variables that determine the efficiency of a screw press in dewatering sludge are screw speed, headbox level, pressure cone setting, heat addition and the consistency of the feed stream. The screw speed is directly related to the throughput, and inversely related to the discharge cake consistency, with all other factors remaining constant. The faster it turns, the more material it will convey, resulting in a reduced retention time for the sludge in the press, and therefore less dewatering.

The headbox level is directly related to both throughput and discharge cake consistency with all other factors constant. If the headbox level is increased (i.e. the sludge flow to the press is increased), throughput, cake solids content, and motor amps will all increase. The higher head level promotes increased gravity driven dewatering in the "free drainage" section at the inlet end of the press. This results in higher consistency material being fed into the press, increasing the throughput, and the increased loading results in greater dewatering.

The purpose of the pressure cone is to apply a back pressure on the cake in the high pressure section of the press. The pressure can be adjusted by changing the spring force on the pressure cone. As the back pressure is increased, the discharge cake consistency increases and the throughput either remains constant or is decreased slightly. Some adverse effects of increasing the back pressure include, accelerated wear on the press, a higher motor loading, increased chance of

plugging, decreased capture efficiency for low fibre sludges, and the possibility of rupturing the screen.

The addition of heat by injecting steam into the hollow screw shaft of the press can increase the cake discharge by making the sludge easier to dewater. Heat addition can rupture the biological cell membranes, releasing some of the water that would not otherwise be removed. It also lowers the viscosity of the filtrate causing it to flow more freely under certain circumstances. The addition of heat has the added benefit of reducing the friction between the screw and the sludge, which reduces wear and the energy consumption of the press (Badar, 1987).

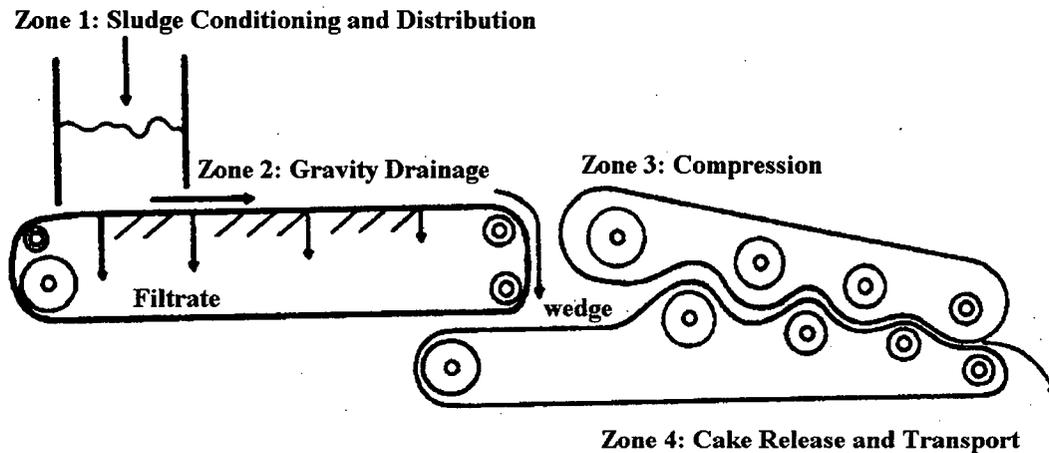
The inlet consistency to the screw press is directly related to throughput capacity, sludge cake discharge, and capture rate efficiency.

#### **2.1.4.2 Belt Press (Belt Filter)**

Another kind of press often used in mills for sludge dewatering is the belt press. Belt presses typically produce dewatered sludge with a lower percentage of solids than screw presses. In mills using gravity thickeners in combination with belt presses, the cake solids content is usually under 25% (Kenny, 1997).

Sludge dewatering by belt press filtration utilizes a combination of gravity and applied mechanical pressure as the driving force for the process. There are four zones depicted in Figure 2-4 (Severin, 1998). Belt presses may be of the two-belt variety in which the lower pressure belt is used as the gravity drainage section, or three-belt presses where an independent belt is used for gravity drainage. In Zone 1, sludge is conditioned to produce a drainable floc. The conditioned sludge is then applied to a gravity drainage belt (zone 2) where water is removed from the forming cake. Drained cake then enters the press zone (zone 3) where a top belt is contacted with the bottom belt, forming a low pressure 'wedge'. A series of rollers with decreasing diameters then contact the belts. This has the effect of increasing the contact pressure as the sludge is moved through the process. The action of moving the belt around the rollers creates shear stress on the sludge cake by differential movement of the belts. The shear forces are such that some of

the bound water and possibly some intercellular water is released in the high pressure section (WPCF, 1983). Finished cake is released from the belt in Zone 4.



**Figure 2-4. Three belt press**

#### **2.1.4.3 Screw Press vs. Belt Filter**

Belt presses and screw presses have some fundamental similarities. Both operate on a continuous basis (not batch) to make a rough separation of liquids and solids. Both of them have a filtration stage followed by a solids compression stage. However, each machine has its own strengths and weaknesses.

Because a belt press works with a filter made of cloth, it tends to retain more of the finer particles than is possible with a screw press. The metal screen (perforated metal or profile bar) in a screw press results in far more suspended particles in the press liquor than is characteristic of a belt press. For this reason a screw press is not acceptable, operating by itself, for dewatering wastewater sludge. Filtrate from a screw press is usually recycled back to the inlet of the wastewater treatment process.

On the other hand, a screw press can squeeze a lot harder than a belt press. The all-metal construction of a screw press allows higher compression and higher applied torque. The result is that a screw press can press sludges of lower moisture content than is possible with a belt filter.

## 2.2 Parameters for Measuring Sludge Dewaterability

### 2.2.1. Specific Resistance to Filtration (SRF) and Filtration Theory

When dewatering sludge, by means of centrifugal filters, belt presses or screw presses, filtration is one of the major mechanisms by which water removal takes place, the other is compression of the wet cake formed in the filtration stage (Egenes 1992). This has led to the use of specific resistance to filtration (SRF) as a lab parameter to characterize the dewaterability of a sludge (Coakely et al. 1956). It is based on an analysis of pressure drop for flow through a porous medium using Darcy's law. In 1856, Darcy discovered that the pressure loss through a sand filter was directly proportional to the flow rate of the fluid.

$$\frac{\Delta P}{L} = \frac{\mu dV}{kdt} \frac{1}{A} \quad (2-1)$$

Where  $\Delta P$  is the pressure drop,  $L$  is the cake depth,  $\mu$  is the liquid viscosity,  $dV$  is volume of filtrate flowing in time  $dt$  and  $A$  is the cross-sectional area of the filter bed.  $k$  is permeability.

For cake filtration, during filtration the cake depth increases due to deposition of solids at the filter cake surface. The change in cake depth is accompanied by changes in fluid flow rate and pressure differential, as filtration time increases. A material displaying constant cake concentration is *incompressible*, and the type of filtration is known as incompressible cake filtration. But most filtration processes involve *compressible* cake filtration, which requires some modification of the basic theory (Rushton et al. 1996).

For incompressible filtration the cake concentration remains constant throughout the depth of the cake, thus for each unit volume of suspension filtered the filter cake volume increases by a uniform and constant amount. So a constant of proportionality  $\beta$  can be used to give an equation for cake depth at any instant in time:

$$L = \frac{\beta V}{A} \quad (2-2)$$

this can be substituted into equation (2-1) to give:

$$\frac{dV}{dt} = \frac{A^2 \Delta P k}{\beta V \mu} \quad (2-3)$$

Multiply the right hand side of equation (2-3), both numerator and denominator by the cake solid concentration  $C$  and density  $\rho_s$ , and rearrange the resulting equation to give:

$$\frac{dV}{dt} = \left(\frac{A}{\beta V C \rho_s}\right) \left(\frac{k C \rho_s}{1}\right) \left(\frac{A \Delta P}{\mu}\right) \quad (2-4)$$

Let:

$$w = \frac{\beta V C \rho_s}{A} = L C \rho_s \text{ (from equation 2-2) = the mass of dry solids per unit area of filter.}$$

and  $\alpha = \frac{1}{k C \rho_s}$  = a constant called the “specific resistance to filtration (SRF)” which has units of  $m \text{ kg}^{-1}$ .

So given the equation:

$$\frac{dV}{dt} = \frac{A \Delta P}{\mu w \alpha} \quad (2-5)$$

if the term  $c$  represents the mass of dry cake deposited per unit volume of filtrate, and  $V$  is cumulative filtrate volume, then:

$$w = c \frac{V}{A} \quad (2-6)$$

Substituting equation (2-6) into equation (2-5) gives:

$$\frac{dV}{dt} = \frac{A^2 \Delta P}{\mu c V \alpha} \quad (2-7)$$

For constant pressure filtration, the overall pressure drop can be considered as the sum of the pressure drop in the medium  $\Delta P_m$  and the pressure drop over the filter cake  $\Delta P_c$ :

$$\Delta P = \Delta P_c + \Delta P_m \quad (2-8)$$

Darcy’s law can be applied to both terms:

$$\Delta P = \frac{\mu c \alpha}{A^2} V \frac{dV}{dt} + \frac{\mu L_m}{A K_m} \frac{dV}{dt} \quad (2-9)$$

where  $L_m$  and  $K_m$  are the medium depth and permeability. If they remain constant during filtration, they can be replaced by  $R_m = \frac{L_m}{K_m}$ , which is called the medium resistance with units of  $m^{-1}$ .

Substitute  $R_m$  into equation (2-9) and integrate:

$$\int_0^t dt = \frac{\mu c \alpha}{A^2 \Delta P} \int_0^v V dV + \frac{\mu R_m}{A} \int_0^v dV$$

After integration and rearrangement the following equation, known as the linearised parabolic rate law results:

$$\frac{t}{V} = \frac{\mu c \alpha}{2A^2 \Delta P} V + \frac{\mu R_m}{A \Delta P} \quad (2-10)$$

Many materials will give a filter cake of roughly constant concentration, when filtering under conditions of constant pressure, as would be expected in an incompressible filtration. Increasing the filtration pressure results in another cake of roughly uniform concentration but higher than the first. Thus compressibility is displayed between filtrations at different pressures, but not necessarily within a filtration. This observation has led to the almost universal adoption of the concept of *average* values of specific resistance. So equation (2-10) can be used for both the incompressible and compressible cake. For compressible cake  $\alpha$  represents the average specific filtration resistance. The major difference between compressible and incompressible cakes is that  $\alpha_{av}$  becomes a function of pressure. The following equation (Grace, 1953) expresses the relationship between specific resistance and pressure drop over the cake:

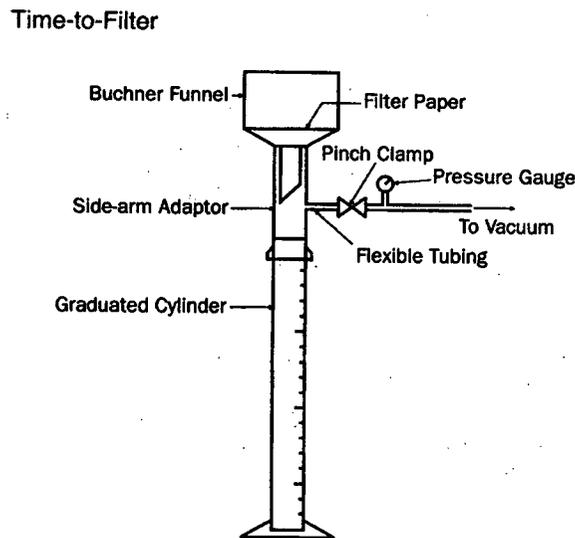
$$\alpha = \alpha_0 \Delta P^n \quad (2-11)$$

$\alpha_0$  and  $n$  are empirical constants,  $n$  is related to compressibility. For an incompressible cake,  $n = 0$ ; for a compressible cake,  $n > 0$ . According to Leu (1981), degree of compressibility: slight,  $n = 0.20$ ; moderate,  $n = 0.60$ ; high,  $n = 1.20$ .

Equation (2-10) plots as a straight line, where  $t/V$  is the dependent and  $V$  is the independent variable. Thus its slope =  $\frac{\mu c \alpha}{2A^2 \Delta P}$  and its intercept =  $\frac{\mu R_m}{A \Delta P}$ . If the liquid viscosity, filter area, filtration pressure and mass of dry cake per unit volume of filtrate are known, the values of the slope and intercept of the plot can be used to calculate the cake specific resistance and filter medium resistance.

### 2.2.2 Time to Filter (TTF)

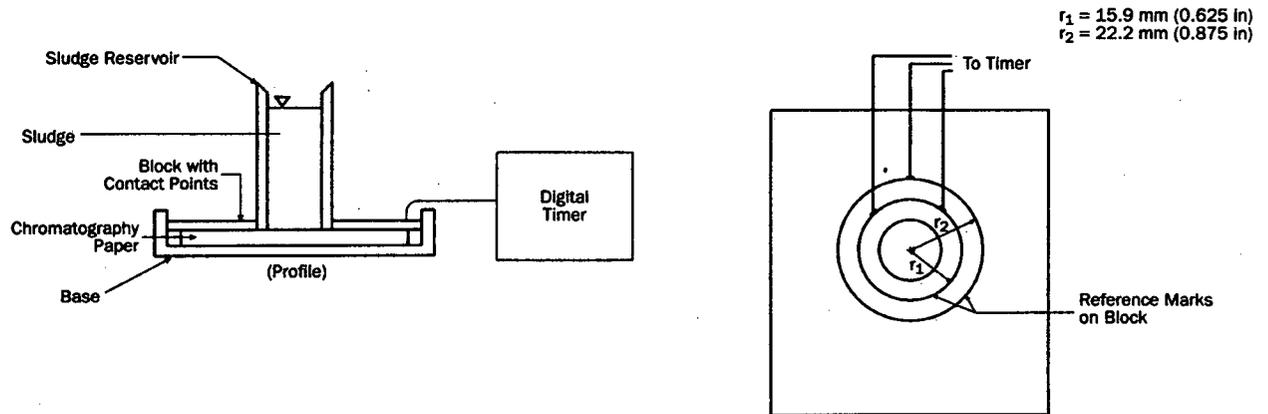
Time to filter is similar to the specific resistance to filtration (SRF) if sludge solids content and filtrate viscosity do not vary among compared samples. The test consists of placing a sludge sample of 200 ml in a Buchner funnel with a paper support filter, applying vacuum, and measuring the time required for 100 ml filtrate (50% of original sample) to collect. While similar to the specific resistance to filtration test, the time to filter test is superior in terms of its ease of use and simplicity.



**Figure 2- 5. TTF equipment**

### 2.2.3. Capillary Suction Time (CST)

Since measurement of the specific resistance to filtration is cumbersome and time consuming, Gale and Baskerville (1968) developed what they called the capillary suction time (CST) apparatus. It is a quick and easy method for assessing sludge filterability. The CST is the time required for the liquid fraction of a sludge to travel a given distance in an adsorbant paper. Although CST is a purely empirical test and is not based on any theoretical analysis of sludge dewaterability, some authors have given the method a theoretical basis (Vesilind, 1988; Frank, 1990).



**Figure 2 - 6. CST apparatus**

#### **2.2.4 Limitations of SRF and CST**

Although SRF and CST are widely used to characterize sludge dewaterability, some workers have found that there are some deficiencies in these two parameters.

SRF and CST do not take the structural elements of sludge flocs into consideration, they measure only the filterability of a sludge and this need not correspond with the moisture content (final cake solids) in the dewatered sludge (Olboter and Vogelphohl, 1993).

Karr and Keinath (1978) found two limitations of SRF. First, SRF values are superficially influenced by the total solids concentration of a sludge when fines are present to the extent that they cause blinding of the passageways through the sludge or the filter medium, this is contrary to specific resistance theory. Second, the specific resistance test does not account for those properties (such as pick-up, release, scrolling) that are important for dewatering sludge using real full-scale equipment.

According to Tosun et al (1993), the average specific cake resistance comes from the classical filtration theory which may not explain the mechanism of the filtration very well because of the inconsistencies and contradictions in its development. They thought that the controlling factor in

the filtration was not the resistance of the cake but the resistance of the cake – septum interface. This point came from their work showing the effect of the filter medium on the filtrate volume-time data. The slope of the  $t/V$  versus  $V$  was strongly affected by the filter medium. For the same type filter medium, there was change in the slope for up-flow and down-flow experiments.

CST values are dependent on initial sludge solids concentration and cannot give information about the dewatered cake or the filtrate suspended solids. Karr and Keinath (1978) observed that the CST test seemed to be more sensitive to changes in concentration of fragile settleable solids than the SRF test, even though the total settleable solids concentration was maintained constant. For the CST test, only a small volume of filtrate is removed from the sludge, CST results, therefore, are more sensitive to pore size and initial packing characteristics (Christensen, J.R. and Christensen, G.L.,1997). The CST test seems less appropriate as a measure of sludge conditioning performance than SRF and TTF (time to drain a fixed volume) due to its lack of sensitivity (Christensen, J.R. and Christensen, G.L.,1995).

### **2.2.5 Pressure Filtration Time (PFT)**

Pressure Filtration Time is also a measurable filtration parameter, developed in the UK (Hoyland, 1986). It can be measured by a specific PFT-meter which is commercially available (Plaisier, 1997).

PFT is the time measured between the times at which filtrate volumes of 5 and 15 cm<sup>3</sup> appear with a filtration area of 12 cm<sup>2</sup> when 30 cm<sup>3</sup> of sludge are filtered at constant pressure. PFT is inversely proportion of the SRF when the pressure drop, solids content and filtrate viscosity are constant. If these variables don't change too much, PFT is also comparable among samples. PFT is similar to TTF, but it can be measured automatically. The device also can provide values of the sludge cake solids at certain press time.

## **2.3 Chemical Conditioning**

Thickening and dewatering of sludge, particularly those containing sludges from biological treatment systems, generally are not practical without chemical conditioning. Conditioning

chemicals reduce the sludge particle surface charge, aggregate them together to form big flocs, and thus significantly increase the sludge dewaterability. There are two types of effect: coagulation and flocculation. Coagulation is the adhesion of particles by forces of molecular and atomic origin and the presence or absence of coagulation depends upon the balance between the attractive van der Waals' forces and the repulsive electrical double layer forces (Rushton, et al. 1996). Flocculation refers to processes where certain types of long-chain polymers or polyelectrolytes cause the particle to aggregate by forming bridges between them ( Rushton, et al. 1996).

### **2.3.1. Chemicals Used in Sludge Conditioning**

Various types and combinations of coagulants and flocculants are used in sludge conditioning. Chemicals used in sludge conditioning include both inorganic and organic chemicals. Typical inorganic chemicals are ferric, ferrous, and aluminum salts, and calcium oxide or hydroxide (WPCF, 1988). Organic chemicals generally used for sludge conditioning include the large group of organic polyelectrolytes (polymers). Inorganic chemicals are usually used as coagulants, organic polymers can be used as both coagulants and flocculants. Typical organic coagulants are polyamines, polyacrylamides and diethylamines (Kenny, 1995).

### **2.3.2 Inorganic Chemicals vs. Organic Polymers**

Inorganic chemicals have been used in sludge conditioning for many years, e.g. as far back as 1740 in France (WPCF, 1988). Although ferric chloride and lime are still widely used sludge conditioning chemicals with vacuum filtration and recessed plate filtration, organic polymers have replaced inorganic conditioners with other sludge dewatering equipment such as the belt filter press and the screw press.

Organic polymers have some advantages over inorganic chemicals: (Novak, 1980; WPCF, 1988; Groves, 1995)

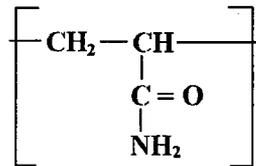
- They are effective at much lower dosages than inorganic chemicals so contribute little to additional weight of the sludge. The use of inorganic chemicals such as ferric chloride and lime can add as much as 20 to 40% more dry weight to the final dewatered sludge.(WPCF, 1988)

- For incineration, a benefit of adding organic polymers is that they are easy to incinerate, while the inorganic chemicals tend to cause corrosion and surface deposit build-up problems in incinerators and as well they lower the heating value of the sludge.
- Their use eliminates lime handling and storage problems, corrosion from iron salts, scale buildup from lime conditioning, and general wear-and-tear associated with inorganic conditioners.
- They are effective over a wider pH range. The use of inorganic chemicals such as ferric chloride and aluminum salts lower the pH value due to the hydration of ferric and aluminum salts. Lime has to be used to adjust the pH value to the optimum for ferric and aluminum salts working. Ferric chloride coagulation works best above pH 6.
- Organic polymers can vary in structure, charge, charge intensity, and molecular weight and therefore should be capable of being tailored to the specific sludge characteristics.

The disadvantage of organic polymers use is their high cost. The cost at individual sludge treatment facilities may be considerable, even exceeding the cost of power for aeration of the wastewater. Many pulp and paper mills use organic polymers. Less than 30% of pulp and paper mills use ferric chloride or ferric chloride + lime as coagulant as reported in a 1994 survey (Kenny, 1995).

### **2.3.3 Organic Polymers**

Organic polymers are synthetic giant molecules formed by a long string of monomer units, linked together in a linear or branched configuration, with functional groups located periodically along the chain. In general, the functional groups may possess a negative charge (anionic polymers), positive charge (cationic polymers), or an overall neutral charge (nonionic polymers). If they have charges, they are also called polyelectrolytes. Most of the polymers used in sludge conditioning are cationic polymers, because the majority of sludge particles bear a negative charge. Most of these organic polymers are polyacrylamides. The structure of an acrylamide monomer is shown in Figure 2-7.



**Figure 2-7. Acrylamide monomer**

### **2.3.3.1 Characteristics of Polymers**

The characteristics of a polymer include: charge type, chemical “back-bone” or species, molecular weight, charge density and active solids. Charge type refers to whether the polymer is cationic, anionic or nonionic. The chemical “back-bone” refers to the basic polymer type that is used e.g. acrylamide. The molecular weight is a measure of the back-bone length or chain length. It is associated with the ability of the conditioner to effect bridging between sludge flocs. Bridging is usually associated with medium- and high-molecular-weight polymers. Electric charge density is a measure of the amount of sites on a polymer chain that has a charged group associated with that polymer. Thus on a cationic polymer, all positively charged sites would be measured. The remaining sites would be neutral or possibly have a negative charge. Charge density is expressed as percentage of available sites or amount of positive or negative charge (in meq) contained in one gram of polymer, as determined by titration (WERF, 1993). Charge density is employed in neutralization of electric charge built on the double layer surrounding the solids particle’s surfaces. The mechanism will be described in section 2-4. The final variable is active polymer content, the actual polymer content of a particular conditioning product. Such products may also contain water, mineral oils, surfactants, and impurities, and therefore contain less than 100% actual polymer (WERF, 1993).

### **2.3.3.2 Polymer Forms (WPCF, 1987)**

Polymers are manufactured in five different physical forms:

**DRY POLYMER.** Dry polymers are produced as either powders, granules, beads, or flakes. They are very high in active polymer content. Varying with the manufacturer, the active solids can be as high as 90 to 95 %. The remaining portion is either water or inert solids.

**EMULSION POLYMERS.** Emulsions are dispersions of polymer particles in a hydrocarbon oil. The active solids level can achieve 25 % to 50 %. The apparent viscosity of the neat emulsions ranges from 300 to 5000 cp (centipoise).

**MANNICHS.** Mannichs are clear-to-milky liquids with high molecular weights and very high viscosities. They range in active solids from 2 % to 10 % and may have pH as high as 12.

**LIQUIDS.** These are low-to-medium in molecular weight. Active solids can range from less 10 % to 50 %.

**GELS.** Gels are solutions of high molecular weight polyacrylamide polymers or co-polymers. They require special make-up equipment.

### **2.3.4 Inorganic Polymer- PAC**

Polyaluminum chloride (PAC) is actually the intermediate product of aluminum ion hydrolysis. The general formula for is  $Al_n(OH)_mCl_{3n-m}$ , where  $0 < m < 3n$ . It has better coagulation/flocculation properties than traditional inorganic coagulants such as aluminum chloride, aluminum sulfate and ferric chloride. Although the effect on flocculation is less than organic polymer flocculants due to its lower degree of polymerization, the price is much lower than organic polymer flocculants (Tang, 1999). The polymeric aluminum species have higher ionic charge than either alum or aluminum chloride. PAC can maintain its cationic charge into the alkaline pH range. Addition of PAC does not change the system pH as much as an equivalent amount of aluminum as aluminum chloride or sulfate (Gess, 1998).

## **2.4 Mechanisms of Chemical Conditioning**

There are two major mechanisms for sludge conditioning by organic polymers: charge neutralization and bridging.

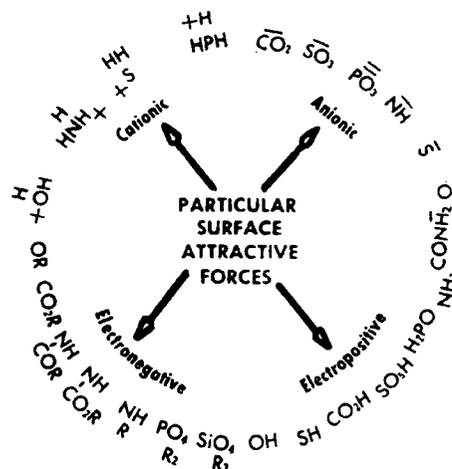
### **2.4.1. Charge Neutralization**

#### **2.4.1.1. Origin of Surface Charge**

When solids are in contact with an aqueous electrolyte solution, there are several possible mechanisms for the development of a surface charge (Elimelech, et al, 1995, Weber, 1996):

- (a) a difference in affinity of ions for the two phases. In the case of ionic solids, one or other of the constituent ions may have a greater tendency to dissolve in the aqueous phase.
- (b) ionization of surface sites on slightly soluble compounds of weak acid anions, such as sulfides, carbonates, phosphates.
- (c) acidic and basic functional groups on organic colloids such as proteins, polysaccharides and other substituted polymer structures.
- (d) physical restriction of certain ions to one phase. This includes materials with an inherent excess charge as a result of isomorphous substitution. For instance substitution of Si (IV) by Al(III) in many clay minerals gives a net negative lattice charge.

Various surface charge groups are illustrated in Figure 2-8.



**Figure 2- 8. Particle surface charges (Preising, 1962)**

### 2.4.1.2 Double Layer and Zeta Potential

Sludge particles from wastewater treatment systems are typically negatively charged due to any one of the causes listed above. This surface charge must be exactly balanced by an equal and opposite charge in the surrounding liquid. This balancing of charges is accounted for by an excess number of oppositely charged ions (counterions) in the solution adjacent to the charged

surface and a deficit of similarly charged ions (co-ions). Thus an electric double layer is formed. Two competing processes (diffusion and electrostatic attraction) spread the charge in the wastewater over a diffuse layer. The highest concentration of positive counterions exists near the particle surface in the *compact double layer* (Stern layer), and the excess ions less strongly attracted are arranged more loosely around the particle creating the *diffuse double layer* (Gouy-Chapman layer). The potential gradient over this region is called the zeta potential. See Figure 2-9.

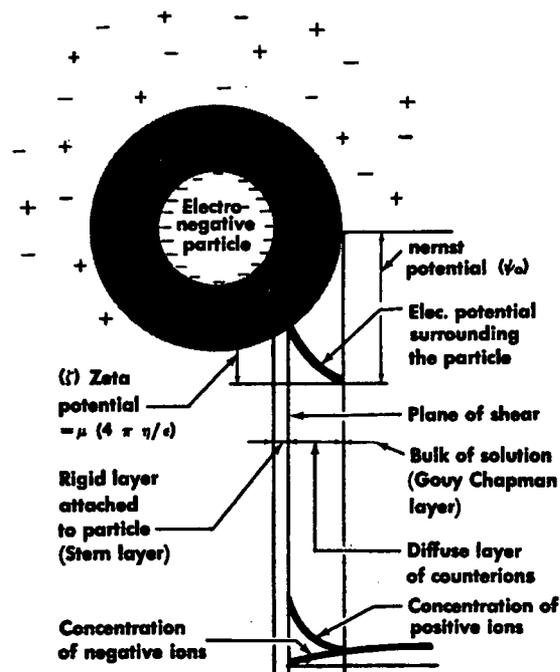


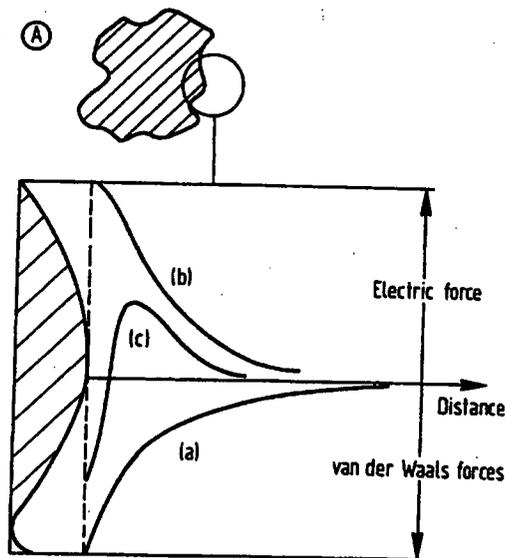
Figure 2- 9. Double layer and zeta potential

#### 2.4.1.3 Colloid Stability (Lyklema, 1988; Elimelech, 1996)

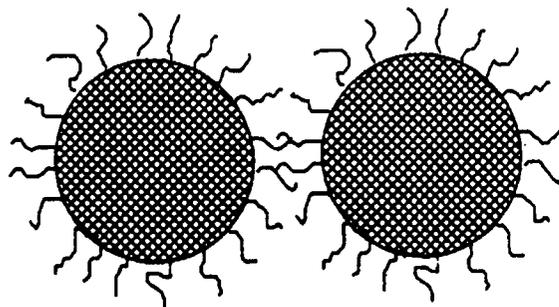
The theory of Derjaguin, Landau and Verwey, Overbeek (DLVO theory) describes two forces between colloidal particles: the van der Waals' attraction and the electric double layer (EDL) repulsion. These two forces are additive. The total interaction, usually expressed as potential energy  $V_T$ , is the sum of the van der Waals attraction  $V_A$  (Figure 2-10 curve a,) and the EDL interaction  $V_R$  (curve b):  $V_T = V_A + V_R$  (curve c). When the repulsion outweighs the attraction,

there is a potential energy barrier, which tends to prevent contact of particles and so prevents aggregation (Figure 2-10). The colloid particles must overcome this energy barrier in order to come into close enough contact to adhere to one another.

There is another force between particles, which is not considered in DLVO theory, i.e. steric interaction. It is caused by adsorbed polymers. If the particles have adsorbed a large amount of polymer, they are stabilized by an effect of steric stabilization as illustrated in Figure 2-11.



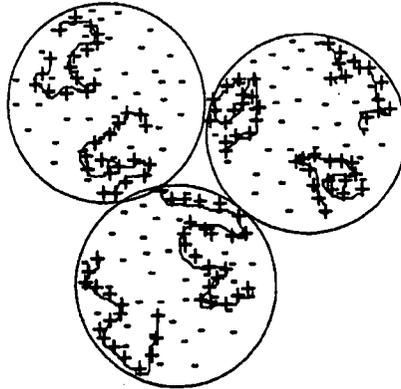
**Figure 2-10. Potential energy curve for stabilized particles**



**Figure 2-11 Steric stabilization (Elimelech, 1996)**

#### 2.4.1.4 Destabilization by Charge Neutralization

Chemicals can be added so as to alter the surface charge thus allowing particle agglomeration to take place. Polymers with positive charges acting as coagulants are adsorbed onto the negatively charged particle surfaces due to the attractive electrostatic forces. Gregory (1993) explained this in terms of an “*electrostatic patch*” model (Figure 2-12); in it adsorbed polymers give rise to ‘patches’ of charge, surrounded by areas of opposite charge. Multivalent metal ions such as  $Al^{3+}$  and  $Fe^{3+}$  can directly neutralize the negative charges on a particle surface. Polymers and ions, then, both destabilize colloidal particles by neutralization of surface charge, thus reducing EDL repulsion. The resultant total energy potential curve  $c$  lies below the zero value (Figure 2-13) thus enabling the particles to coagulate if they should become close enough.



**Figure 2-12. Electrostatic patch model**

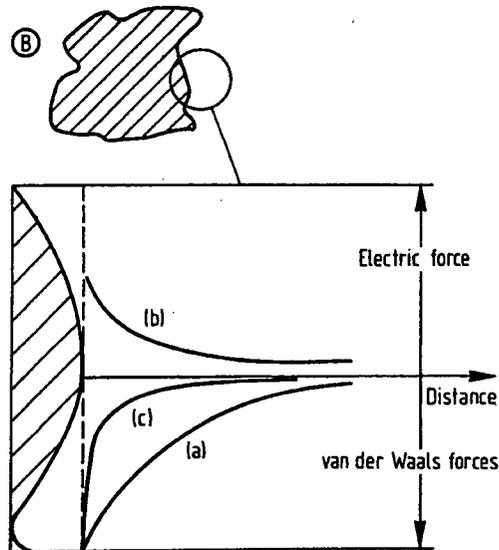


Figure 2-13. Potential energy curve for coagulated particles

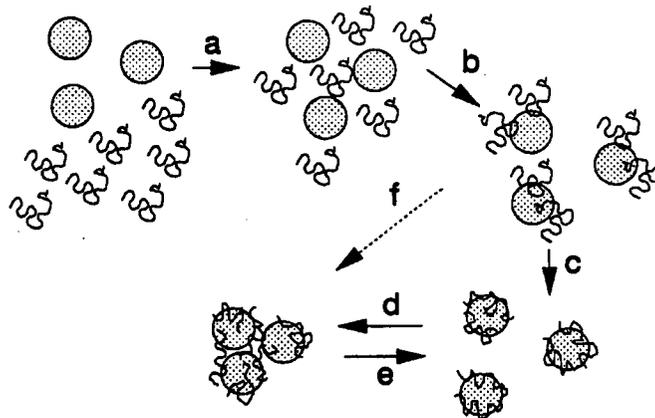
#### 2.4.2 Bridging Mechanism

In many cases high molecular weight polymers are effective by virtue of a bridging mechanism, in which segments of a single polymer chain are adsorbed on two or more particles, thus bridging them together. In this way particles can form aggregates or attach themselves to macroscopic surfaces even though they might be oppositely charged and repel each other. With polyelectrolytes and oppositely charged particles, like a cationic polymer and sludge, there is also the possibility of charge effects; either simple charge neutralization or some form of 'electrostatic patch' interaction.

At the moment of addition of a polymer solution to a suspension of particles, usually under conditions of agitation, several processes take place, see Figure 2-14 (Elimelech, 1995).

- (a) mixing of the polymer solution throughout the suspension
- (b) adsorption of polymer molecules on the particles, to give some degree of destabilization (either by charge neutralization or by providing an opportunity for 'bridging' between particles)
- (c) rearrangement of adsorbed polymer molecules from initially extended state to a flatter, equilibrium conformation

- (d) collisions between destabilized particles to give aggregates (flocs)
- (e) break-up of a floc under the influence of applied shear



**Figure 2-14. Flocculation by polymer**

These processes occur at different rates and simultaneously to form a complicated system, which is difficult to analyze.

Classic colloid chemistry models do not adequately explain the mechanism of polymer sludge conditioning (WPCF, 1988). Colloids tend to respond to chemical conditioning predictably. Waste treatment sludges, on the other hand, tend to be more variable in characteristics and more likely to experience short-term changes in conditioning requirements. Such factors as biological solids content, solids concentration, aging properties, pH and alkalinity, all play a role in sludge conditioning processes.

## **2.5. Factors Affecting Conditioning and Dewatering**

### **2.5.1. Sludge Properties**

#### **2.5.1.1. Particle Size Distribution**

Karr and Keinath (1978) found that sludge particle size distribution was the single most important factor determining the dewaterability. Table 2-1 displays their particle size classification scheme. The effects of factors such as mixing, storing, pH and conditioning could be explained by changes in particle size distribution. Of the solid fractions investigated the supra-

colloidal fraction (1-100  $\mu\text{m}$ ) had the greatest adverse influence on dewaterability due to its ability to blind both the cake and the filter media used in Buchner funnel filtration tests. The true colloidal particles had relatively little effect on sludge dewatering characteristics.

Table 2-1. Particle size classification scheme.

Category	Screen particle dimensions	Subcategories used
Settlable	less than 100 $\mu\text{m}$	Rigid settlable
		Fragile settlable
Supracolloidal	1 to 100 $\mu\text{m}$	None
True colloidal	0.001 to 1 $\mu\text{m}$	None
Dissolved	Less than 0.001 $\mu\text{m}$	None

### **2.5.1.2 Biochemical Composition**

Biological sludges are comprised of viable and non-viable microorganisms, cellular debris, biopolymers and abiotic substances (Bowen, 1982). Each type of material has a different surface for polymer attachment. Surface characteristics of sludge particles are significantly altered by the presence of biopolymers and exocellular material. Proteins, carbohydrates and lipids are components of cellular surfaces, as well as constituents of biopolymers. Both the exocellular biopolymers and the cell wall structure determine the amount of anionic surface charge (Tyagi, 1989).

Roberts and Olsson (1975) and Novak and Haugan (1978) emphasized the importance of anionic biocolloids and biopolymers in the conditioning of dilute, waste-activated sludges with cationic polyelectrolytes. They described organic conditioning as charge neutralization of these biocolloids and biopolymers with cationic polyelectrolytes. The more anionic biopolymers that exist in suspension the greater the amount of conditioning chemicals needed.

Bowen and Keinath (1984) found that the presence of biopolymers and the resulting bioflocculation aid the conditioning process. The production of these polymers depends strongly

on the loading of the treatment process, the composition of the wastewater as well as on the design and the operation of the treatment plant (Eriksson and Alm, 1993). Biopolymers, which are composed mainly of carbohydrates, may surround the cells forming a highly reactive surface. As carbohydrate content increases, the Polymer Index (a measure of minimum polymer dose required to achieve maximum dewaterability) was found to decrease indicating better polymer performance.

Biochemical composition of the sludge definitely influences conditioning and dewatering. How the conditioners work with these components still remains unclear.

### **2.5.2. Mixing Conditions**

In many cases, effective polymeric flocculants are of quite high molecular weight and are added as rather viscous solutions. It is important to disperse them immediately, or local over dosing will occur. So a high mixing intensity is required. On the other hand, if the mixing intensity is too high or the mixing time is too long, the flocs formed could be destroyed, thus decreasing the ease of sludge dewaterability. Therefore mixing conditions including mixing intensity and mixing time have been identified as the most important parameters, beside the dosage of conditioning chemicals (Christensen, 1995; Plaisier, 1997), in sludge dewatering.

Mixing time is easily to characterize, while mixing intensity is considered more difficult to adequately characterize. Mixing intensity is usually characterized by the mean velocity gradient  $G$ , which is expressed as:

$$G = \sqrt{\frac{P}{\mu \cdot V}} \quad \text{or} \quad G = \sqrt{\frac{2\pi \cdot N \cdot T}{\mu \cdot V}} \quad (2-12)$$

where  $P$  = the power (w);  $\mu$  = the sludge absolute viscosity (Pa s);  $V$  = the sample volume ( $\text{m}^3$ );  $N$  = the impeller rotational speed ( $\text{s}^{-1}$ ); and  $T$  = the torque on the impeller shaft (Nm).

But people have questioned whether  $G$  is a valid measure for mixing intensity.  $G$  is only a valid parameter for the flocculation of particles smaller than the Kolmogoroff microscale of turbulence, which is not typical of common water and wastewater flocculation practice. Power

input per unit mass to the two third power is a more appropriate flocculation parameter than  $G$  for common water and wastewater flocculation practice (Cleasby, 1984). Different types of impeller even with the same  $G$  value still had an impact on flocculation. (Christensen, 1995)

In the literature, a controversy exists as to whether low or high intensity mixing is most effective. Novak and Haugen (1980) suggested sludge conditioned at low mixing speeds required the least amount of polymer. Langer and Klute (1993) found that during a period of one second after polymer addition, mixing with maximum intensity (rotational speeds up to 1700 rpm) at the proper polymer dose does not lead to deterioration in dewaterability. The "Guidance Manual for Polymer Selection in Wastewater Treatment Plants" (1993) recommends a mixing condition for sludge and polymer of 200 rpm for 2 minutes.

Plaisier et al. (1997) argued that sludge flocculation differs from the flocculation of dilute suspensions, for which a lot of flocculation theory has been developed. In the latter case slow flocculation for an extended period of time is necessary to maximize particle contact, without breaking apart existing flocculated particles. However, for flocculating sludge, the challenge is no longer the polymer/particle contact, but efficient polymer dispersion soon after polymer addition. They found that high-speed mixing (500-1000 stirrer rpm) and short duration (2-5 seconds) were the best mixing conditions for the combined sludges from a bleached kraft mill and a thermomechanical pulp mill. They also did a full-scale trial, and found that replacing a static mixer with a high-speed, rotary, in-line mixer at the point of flocculant addition could reduce the required polymer dosage, increase the final sludge cake consistency and reduce the pressate suspended solids.

Different sludges have different energy requirements for mixing in order to obtain optimal dewaterability. The mixing requirement strongly dependent on the polymer dose. High polymer doses are less sensitive to mixing, while at low doses, intense mixing becomes important (Langer, 1993).

### **2.5.3 Temperature**

Temperature has direct effect on liquid viscosity. As temperature increases, liquid viscosity decreases. Thus water at a high temperature should flow through the sludge cake more readily than cooler water (refer to equation 2-10). High temperature (heat treatment or thermal conditioning) also can break down or lyse the cell walls of microorganisms contained in biological sludges, releasing the bound water perhaps allowing the remaining solids to dewater more easily. The other advantages of heat treatment are sterilization of the sludge and enhancement of sludge digestibility (WPCF, 1987). In general, increased temperature will be beneficial to sludge dewatering. That is the reason why some screw presses are designed to allow the addition of heat by injecting steam into the hollow screw shaft of the press.

In previous work, Guild (1998) found there was a strong positive correlation between the temperature and both sludge cake solids and the filtrate flowrate for a combined kraft mill sludge.

#### **2.5.4 Pressure**

Sorensen (1993,1997) observed that when the pressure is above a critical value, the filtrate flow rate is nearly independent of filtration pressure for a highly compressible cake structure. This is contrary to the power law which described the compressibility (equation 2-11). According to 2-11, when the pressure drop increases, the specific resistance also increases, thus the filtrate flowrate ( $V/t$ ) should have decreased (inversely related to specific resistance, see equation 2-10). The reason for this is that nearly all the liquid pressure drop happened in the lowest few per cent of cake, just above the medium where a highly impermeable filter skin formed. They found that the critical filtration pressure for waste activated sludge was 2-20 kPa, a value well below that used in dewatering processes, which is usually between 100-1000 kPa (Novak et al.,1999).

Guild (1998) found that in pressing combined kraft mill and paper mill sludge cake solids consistencies increased with increased applied press pressure. The increase leveled off as the applied air pressure reached 482.65 kPa (70 psi).

## **2.6. Streaming Current as a Control Indicator of Polymer Addition in Sludge Dewatering**

Streaming current is one of the electrokinetic phenomena. All colloidal particles in water carry electric charge (see section 2.4.1). This leads to a concentration of oppositely charged ions, the so-called counter-ions, on the colloid surfaces. If these counter-ions are separated from or sheared off the dissociated macromolecule or particle, a streaming current produced. It can be measured in mV. A streaming current of zero mV denotes the point of zero charge; i.e. all existing charges in the sample are neutralized.

Dentel and Abu-Orf (1993) and WERF (1995) have reported that a streaming current detector (SCD) can be utilized as a tool for polymer selection and dose control to improve sludge conditioning prior to centrifugation. The polymer dose range for optimum dewaterability, as indicated by CST and solids recovery, corresponded to zero streaming current in the conditioned sludge and slightly negative SC in the centrate. The findings in support of this conclusion were based upon laboratory centrifugation of a conditioned cellulose suspension, laboratory centrifugation of anaerobically digested sludges from two different facilities, and belt filter pressing of both digested and undigested sludges. They reported the SCD was successful in controlling polymer dependably and appropriately, including periods with both intentional and unexpected perturbations in conditions. SCD is feasible in principle as an automatic sludge conditioning control technology.

## Chapter 3

### MATERIALS AND METHODS

#### 3.1. Recycle Mill Sludge

Primary deinking sludge appears grayish in color. The solids of secondary deinking sludge are also gray, but the supernatant is yellow and has an odour. At the recycle pulp mill, from whence came the sludge used in this study, the primary sludge is approximately 86 % (dry wt) of the total sludge produced.

Table 3-1 Quantity and consistency of the waste deinking sludge stream

Sludge Stream	Flowrate (L/min)	Consistency (% solids)
Primary sludge to gravity table #1	1200	2
Primary sludge to gravity table #2	1600	2
Secondary sludge to gravity table #2	800	1.4

The recycle mill sludge samples were obtained from Newstech Recycling Partnership. In that mill there are two gravity tables for thickening the sludge. One is for primary sludge, the other is for combined sludge. The thickened sludges then go to two separate screw presses. The sludge flowrates and consistencies are given in Table 3-1.

Two batches of samples were collected. One was combined sludge, the primary to secondary ratio was around 2.9:1, the other consisted of separate samples of primary and secondary sludges. Some characteristics of these samples are in Table 3-2.

Table 3-2 Characteristics of deinking sludge samples

Type of sludge	pH	TSS (wt %)	TS(wt %)
Combined	6.64	1.8	2.2
Primary	6.87	2.6	3.1
Secondary	6.67	1.6	2.0

### **3.2 Kraft Mill Sludge**

Kraft mill sludge samples were obtained from Western Pulp Partnership's mill at Squamish B.C.. The primary sludges were very dilute, with consistencies ranging from 0.5 % to 1 %. They contained an appreciable amount of long fibers. Since only 50g of sample were used and there was a tendency for the sample to not be homogeneous, an accurate measurement of consistency was difficult. The secondary sludges were very concentrated with consistencies ranging from 3.5 % to 4.1 %. This resulted in low primary to secondary ratios, or, conversely in high secondary to primary ratios. The primary and secondary sludge samples were obtained separately. To simulate the mill's sludge dewatering practice, the combined sludge was mixed using a primary to secondary ratio of 3:2 (volume ratio). Western Pulp's Squamish mill sludge dewatering system is described in Appendix A-3. The data for the different batches of samples are in Table 3-3.

It was found that the sludge samples from the kraft mill changed a lot from time to time. The primary sludge consistency varied from 0.5 % to 1 %, secondary sludge consistency from 3.5% to 4.1%. So at the same primary to secondary sludge volume ratio of 3:2 that simulated the mill, the primary to secondary sludge ratios by weight were different.

Table 3-3 Sampling data summary

Sampling date	Consistency (%)			Primary to secondary ratio (mass based)	pH of combined sludge
	primary	secondary	combined		
Feb-20/2000	0.6	3.7	2.4	1:4.1	6.71
Mar-21/2000	0.5	4.1	1.65	1:5.5	6.73
May-4/2000	1.0	3.6	1.95	1:2.4	6.68

### 3.3 Conditioning Chemicals

#### 3.3.1. Organic Polymers

##### 3.3.1.1 PERCOL Series

PERCOL 721, 734 and 765 were supplied by Ciba Specialty Chemicals. They were synthetic, high molecular weight, water soluble polyelectrolytes.

PERCOL 734 is a very high molecular weight, cationic acrylamide copolymer, with a medium cationic charge density. It is supplied in a free flowing granular form.

PERCOL 721 is an ultra high molecular weight, low charge density, cationic copolymer of acrylamide and acryloxyethyltrimethylammonium chloride. It is supplied as free flowing granules. Its structure is shown in Figure 3-1.

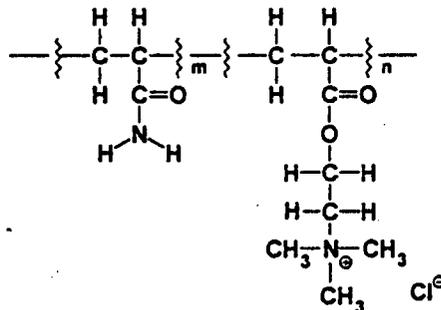


Figure 3-1. Structure of PERCOL 721

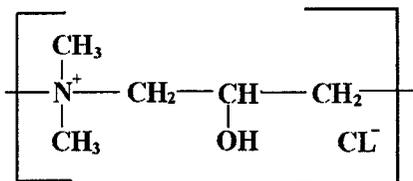
PERCOL 765 is a high molecular weight, medium charge density, cationic acrylamide copolymer, supplied in a unique micro-bead form, which has a very rapid dissolving rate.

### 3.3.1.2 Perform pc8704

Perform pc8704 was the flocculant used by Western Pulp Squamish at the time the samples were obtained. It was supplied by Hercules Chemical Co. It is polyacrylamide based with a charge density of 5 mol percent cationic and a molecular weight of around 6.5 million.

### 3.3.1.3. C4030

C4030 is a polyamine coagulant. It was supplied by Callaway Chemical Company. It is used in dual polymer systems followed by a flocculant. Its charge density is 100 % charge at pH = 6. Its molecular weight is about 150,000. Its structure is as in Figure 3-2.



**Figure 3-2. Structure of C4030**

### 3.3.1.4. F4848

F4848 is a flocculant based on polyacrylamide. It too was supplied by Callaway Chemical Company. Its charge density was 48 weight % cationic charge. Its molecular weight was 12-15,000,000. There are two kinds of functional groups in F4848, they are shown in Figure 3-3.

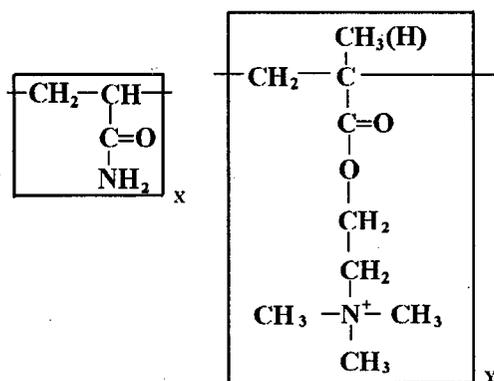


Figure 3-3. Structure of F4848

### 3.3.2. Inorganic Chemicals

#### 3.3.2.1 Ferric Chloride

The ferric chloride utilized in the experiments was a purified grade chemical reagent from Fisher Scientific.

#### 3.3.2.2 Poly Aluminum Chloride (PAC)

Poly aluminum chloride is a new generation inorganic polymer. It was obtained from Raisio Chemicals of Vancouver, BC. The content of  $\text{Al}_3\text{O}_2$  is 17%.

### 3.4. Chemicals for Use in the Particle Charge Detector

#### 3.4.1. Poly(diallyl dimethyl ammonium chloride) (PDADMAC)

A standard, 0.001N, cationic, polyelectrolyte solution, was supplied by Mutek Analytic GmbH of Hamilton, Ontario.

#### 3.4.2. Potassium polyvinyl sulfate (KPVS)

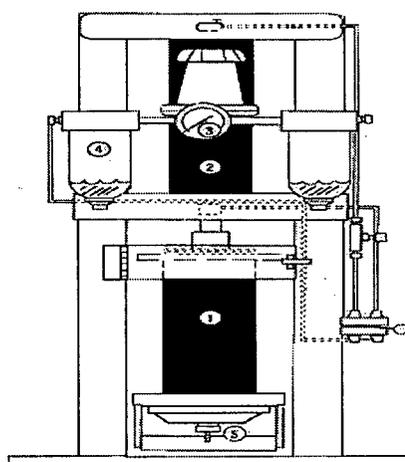
A standard, 0.001N, anionic, polyelectrolyte solution, was supplied by Mutek Analytic GmbH of Hamilton, Ontario.

### 3.5. Apparatus

#### 3.5.1. Laboratory Sludge Press

The laboratory sludge press, which was adapted from a pulp press design of PAPRICAN, is illustrated in Figure 3-4, consists of a 75 mm diameter, 1 liter total volume plexiglas cylinder that can be rotated out for sample loading, an 85 mm diameter, 25 mesh screen that is used at the bottom of the cylinder to separate the dewatered sludge from the pressate or filtrate and an air-operated piston in cylinder to drive a stainless steel plunger (74 mm diameter) down into the sludge slurry. The air pressure in the cylinder is controlled by a valve and pressure gauge on the inlet side, and is capable of applying pressure up to 260 kPa (80 psi). At the base of the unit is a small nozzle for the pressate to exit the cylinder. (Groves, 1995; Rehmat, et al 1997)

The pressure displayed on the pressure gauge is the pressure exerted on the piston in the air cylinder. The pressure applied to the sludge cake is different from the pressure read on the gauge due to the area difference between the piston in the air cylinder and the piston in the sludge cylinder. The pressure actually applied to the sludge can be calculated based on this piston area difference, thus the pressure which is actually applied to the sludge is the air pressure multiplied by 0.46.

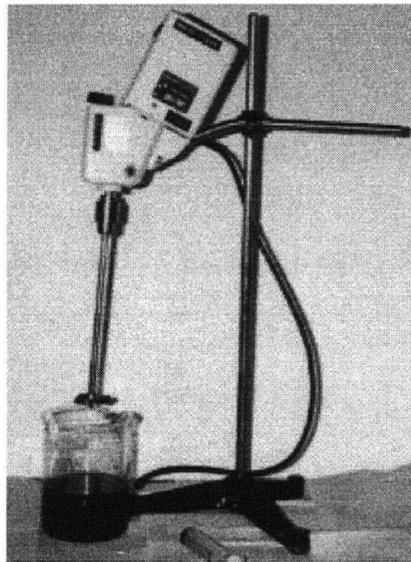


**Figure 3-4 Schematic of the laboratory sludge press.**

(1) plexiglas cylinder, (2) hydraulic cylinder, (3) pressure gauge, (4) coalescing filters, (5) pressate exit nozzle.<sup>2</sup>

### 3.5.2. Mixing System

For sludge and polymer mixing, a stirrer with a .05 m diameter, three-bladed, marine type impeller was utilized. The stirrer was a Caframo RZR1 which can vary the speed from 400 rpm to 2000 rpm. The sludge containers used were 1000 ml beakers. See Picture 3-1.



**Picture 3-1. Mixing system**

### 3.5.3 Data Recording System

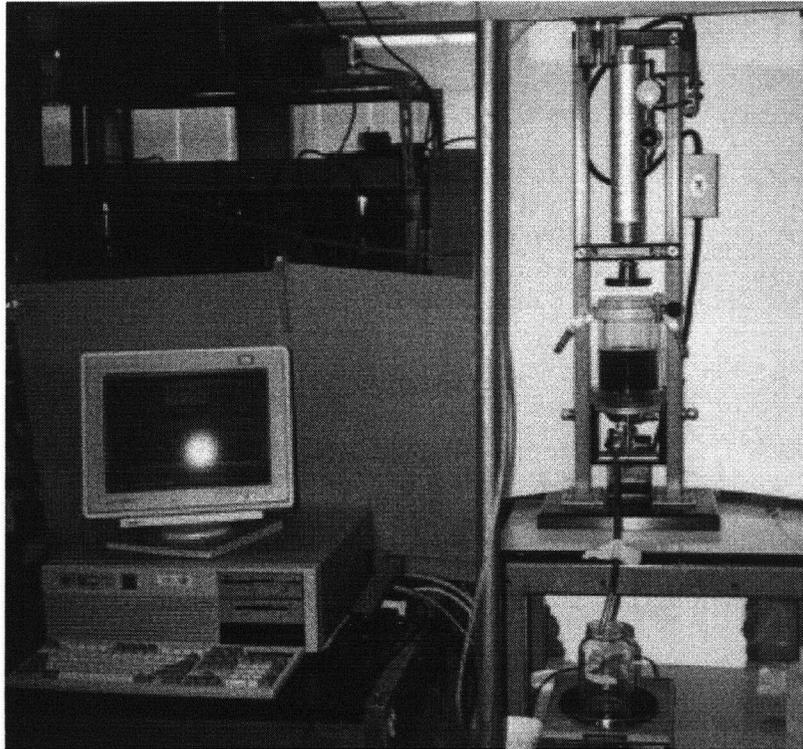
A digital balance connected to a computer was used to measure the weight of the pressate as a function of time. Since it was not possible with this system to record the volume of filtrate automatically and continuously, divide the two sides of equation (2-10) by  $\rho$  to give:

$$\frac{t}{\rho V} = \frac{\mu c \alpha}{2\rho^2 A^2 \Delta P} \rho V + \frac{\mu R_m}{\rho A \Delta P} \quad (3-1)$$

or

$$\frac{t}{W} = \frac{\mu c \alpha}{2\rho^2 A^2 \Delta P} W + \frac{\mu R_m}{\rho A \Delta P} \quad (3-2)$$

$W$  in equation (3-2) is weight of the filtrate,  $\rho$  is the density of the filtrate. So a plot of  $t/W$  vs.  $W$  should give a straight line from the slope of which the specific resistance to filtration could be estimated. This was accomplished by the balance connected to a computer. See Picture 3-2.



**Picture 3-2. Experimental System**

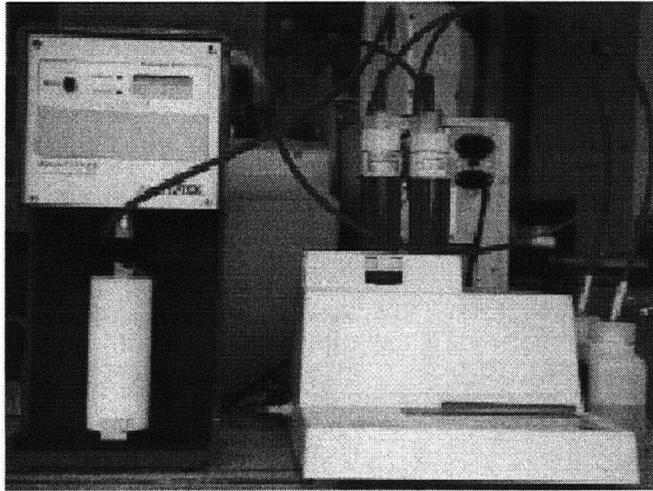
#### **3.5.4. Particle Charge Detector (PCD) and PCD-Titrator**

##### **3.5.4.1. Principle of Operation of the PCD**

The particle charge detector measures the streaming current generated by a flowing colloid system. If an aqueous sample is loaded into the measuring cells of the PCD, colloiddally suspended molecules will adsorb at the plastic surface of the piston and on the cell walls under the action of van der Waal forces. The counter-ions remain comparatively free. A defined narrow gap is provided between cell wall and piston. Driven by a motor, the piston oscillates in the measuring cell and creates an intensive liquid flow which entrains the free counter-ions, the counter-ions then induce a current which is measured by the built-in electrodes (Mutek Manual).

### **3.5.4.2. PCD and PCD-Titrator**

The PCD monitors the streaming potential in mV. The PCD-titrator has a titration end point of zero charge (0 mV). Titration is discontinued as soon as the point of zero charge is reached. The volume of titrant used to bring the streaming potential to 0 is a measure of the initial charge on the particles. The titrant is a standard solution of a polymer having an opposite charge to the charge on the particles. The result can be reported by specific charge quantity (eq/g) if the solids content of the sample is known. In the case of similar sample solids and same titrant concentrations, the measured volume of consumed titrant in ml may be directly used, which is called anionic or cationic demand (Mutex, 1998). Picture 3-3 shows the PCD and PCD-titrator.



**Picture 3-3. PCD and PCD-titrator**

## **3.6 Procedures**

### **3.6.1 Chemicals Preparation**

#### **3.6.1.1 Solid Polymers**

The desired amount of solid polymer was accurately weighed out and dissolved in a weighed amount of distilled water which was stirred by a stir-bar. The mixture was stirred for 2-3 hours to form a homogenous solution, making sure no 'fish eyes' were generated. Since the solution storage time before addition to the sludge sample was typically 12 hr, enough polymer solution for a full day's experimental plan had to be made just before the experiment. The concentration of polymer to be applied was as recommended by the suppliers.

Table 3-4 Polymer concentration for sludge samples

Sludge sample	Polymer concentration
Deinking sludge	0.05%
Kraft sludge	0.2%

### **3.6.1.2. Liquid Polymers**

Both C4030 and F4848 were supplied as liquid emulsions by Callaway Chemical Co. Prior to each day's experiments, a fresh batch of 0.5% (volumetric basis) solution was made for both the coagulant and the flocculant. For the coagulant, 2.5 ml of neat emulsion were added to 500 ml of room temperature, distilled water. The solution was then mixed for 2 minutes using a magnetic stirrer. For the more viscous flocculant 2.5 ml of neat emulsion were added to 500 ml of room temperature, distilled water. The solution was then mixed as recommended by the supplier at 500 rpm for 10-15 seconds, forming a homogeneous solution. The mixture was left standing at room temperature for 45 minutes for aging.

### **3.6.1.3 Inorganic Chemicals**

In the experiments, in which PAC was utilized a 10 % (weight basis) solution was used.

In the experiments utilizing ferric chloride a 1% (weight basis) solution was applied.

Both PAC and ferric chloride were used as coagulants in sludge conditioning, followed by a flocculation stage using organic polymer flocculants.

### **3.6.2 Sludge Conditioning**

435 ± 1 g of a well-mixed sludge sample were weighed in a 1000 ml beaker. The coagulant or flocculant was extracted using a pre-weighed syringe which then was weighed to get the desired amount of conditioner. The mixer was put into the beaker with the sludge. A specific mixing speed was chosen, the sample was homogenized for 10 seconds, then without stopping the mixer,

the coagulant or flocculant was added during 1-2 seconds, and mixed for an additional 2-5 seconds (Plaisier, 1997). If using dual polymers, after adding coagulant 5 minutes were allowed for charge neutralization, then flocculant added following the steps noted above.

### **3.6.3 Sludge Dewatering**

The Plexiglas cylinder was rotated out, at the beginning, the conditioned sludge was carefully poured into the cylinder, some liquid in the sludge went through the screen to drive the air in the bottom out, then the rest of the sludge was poured in. After loading the sludge, the cylinder was rotated back and locked in the vertical position. Then the pressure was adjusted to the desired value. The press was turned on while simultaneously opening the valve on the pressate tube, starting a stopwatch to record the time and running the computer program to record the pressate weight. After 120 seconds, the computer program was stopped, the press was turned off, at the same time, forcing the piston back to its original position, with the vacuum created pulling the sludge cake up. Then the cylinder was rotated out again to allow the water remaining in the bottom to drain as much as possible to the jar on the balance. The final weight of pressate was recorded. As well, the sludge cake was removed from the vessel, its wet weight recorded, and then was transferred to the oven at 105 °C overnight for the dry weight determination.

### **3.6.4. Charge Demand Measurement**

The pressate from the sludge press was filtered through a 200 mesh metal screen. A 20 ml sample of this was centrifuged at 4000 rpm for 10 minutes. Then an accurately weighed, 10 g, centrifuged sample was put into the measuring cell of the PCD, the piston put on and the motor switch opened. If the potential display was negative, PDADMAC was used as the titrant. If the display was positive, KPVS was used. Since the samples were treated by the same procedure, the solids content of the samples were similar. The charge demand is the volume of polymer consumed.

### **3.7 Evaluation Criteria**

Four indicators were used to determine the dewaterability of a conditioned sludge sample. They were sludge cake solids consistency, the pressate total suspended solids (TSS), the flowrate of the pressate ( defined as 250 g of pressate/ the time it took to collect it) and specific resistance to filtration (SRF).

Flowrate of the pressate indicates how fast the water comes out from the sludge matrix. Another similar index is the TTF (time to filter).

Pressate TSS is also an important index for sludge dewatering. Since pressate is recycled to the primary clarifier, if the TSS in the recycle stream is too high, the additional solids load could increase significantly and decrease the overall process efficiency (WPCF, 1983).

### **3.8 Sample Analyses**

The sludge cake solids content and pressate TSS were determined using the American Public Health Association's Standard Methods (APHA, 1992).

### **3.9 Dewatering Profile and SRF Determining**

A C language program written in Matlab was used to draw three figures:

- 1: a plot of weight vs. time, this is the dewatering profile.
- 2: a plot of weight vs. time/weight.
- 3: the linear part of the second figure, using linear regression. Inserting the slope into the equation (3-2), allows the SRF value to be calculated.

## Chapter 4

### RESULTS AND DISCUSSION

#### 4.1 Sludge Press Test Reproducibility

5 replicate tests were done on two batches of combined kraft sludge. The results are shown in Table 4-1.

Table 4-1. Results of sludge press reproducibility test

Ratio of primary to secondary		1:5.5	1:2.4
Air Pressure (psi)		80	60
Mixing Method		Pour 12 times	Mixing + pour
Temperature ( $^{\circ}$ C)		21	27
Dosage (kg/odt)	C4030	10.4	5.6
	F4848	10.4	8.4
Cake Solids Consistency (%)	Average	21.50	24.28
	STDV	0.73	0.23
Filtrate Flowrate (g/s)	Average		60.12
	STDV		1.33
Filtrate TSS (mg/kg)	Average	44.3	11.6
	STDV	4.3	8.9

For the sample with the higher content of secondary sludge (ratio of primary to secondary was 1:5.5), the average cake solids consistency was 21.50% and the standard deviation was 0.73%. The average filtrate TSS was 44.3 mg/kg and the standard deviation was 4.3 mg/kg. For the sample with lower content of secondary sludge (ratio of primary to secondary is 1:2.4), the average cake solids consistency was 24.28%, standard deviation was 0.23%. The average filtrate flowrate was 60.12 g/s, standard deviation was 1.33 g/s. The average filtrate TSS was 11.6 mg/kg, the standard deviation was 8.9 mg/kg. Generally speaking it can be concluded that the reproducibility of the laboratory sludge press process is good, particularly so in the case of filter cake solids consistency.

In many of the figures which follow error bars are presented with the average data; for example see Figure 4-17. These error bars represent one standard deviation (calculated by Excel and plotted by Sigmaplot).

#### **4.2 Polymer-Sludge Mixing**

Based on the paper of Plaisier et al. (1997), in which they found that high-speed mixing (500-1000 stirrer rpm) and short duration time (2-5 seconds) were the best mixing conditions for the combined sludges from a bleached kraft mill and a thermomechanical pulp mill, for our deinking sludge mixing condition tests, we chose various mixing stirrer speeds (for stirrer type refer to Section 3.5.2), and mixing duration times of 1-3 seconds. The pressure applied to sludge in the sludge press was 190.3 kPa (27.6 psi). The results are shown in Figure 4-1.

There are three plots in Figure 4-1; they are plots of cake solids consistency, filtrate flowrate and filtrate TSS vs. mixing speed. First consider the primary sludge results. An optimum mixing speed of around 1000 rpm is evident for the cake solids consistency and filtrate flowrate. Stirrer speed had little effect on the filtrate TSS. For the combined sludge clear optima were not observed. Above 1000 rpm the cake solids consistency, filtrate flow rate and filtrate TSS were not much affected by mixing speed. As a result of these tests 1000 rpm was chosen for subsequent use in the experiments of deinking sludge dewatering.

The shapes of the cake solids concentration and the filtrate flow-rate plots for primary sludge are different from those for the combined sludge. This is perhaps because at the lower polymer dosages (primary sludge dosage 0.84 kg/odt), the cake solids consistencies and filtrate flowrates were more sensitive to mixing speed. At the higher dosages used with combined sludge (1.26 kg/odt), when the mixing speed was over 1000 rpm, the cake solids consistencies and filtrate flow-rate didn't change too much. This may have occurred because a higher polymer dose could result in stronger flocs, which had higher resistance to shear. But even at a dosage of 1.26 kg/odt, a low mixing speed of 400 rpm resulted in lower cake solids consistency and high filtrate TSS. That indicated insufficient mixing.

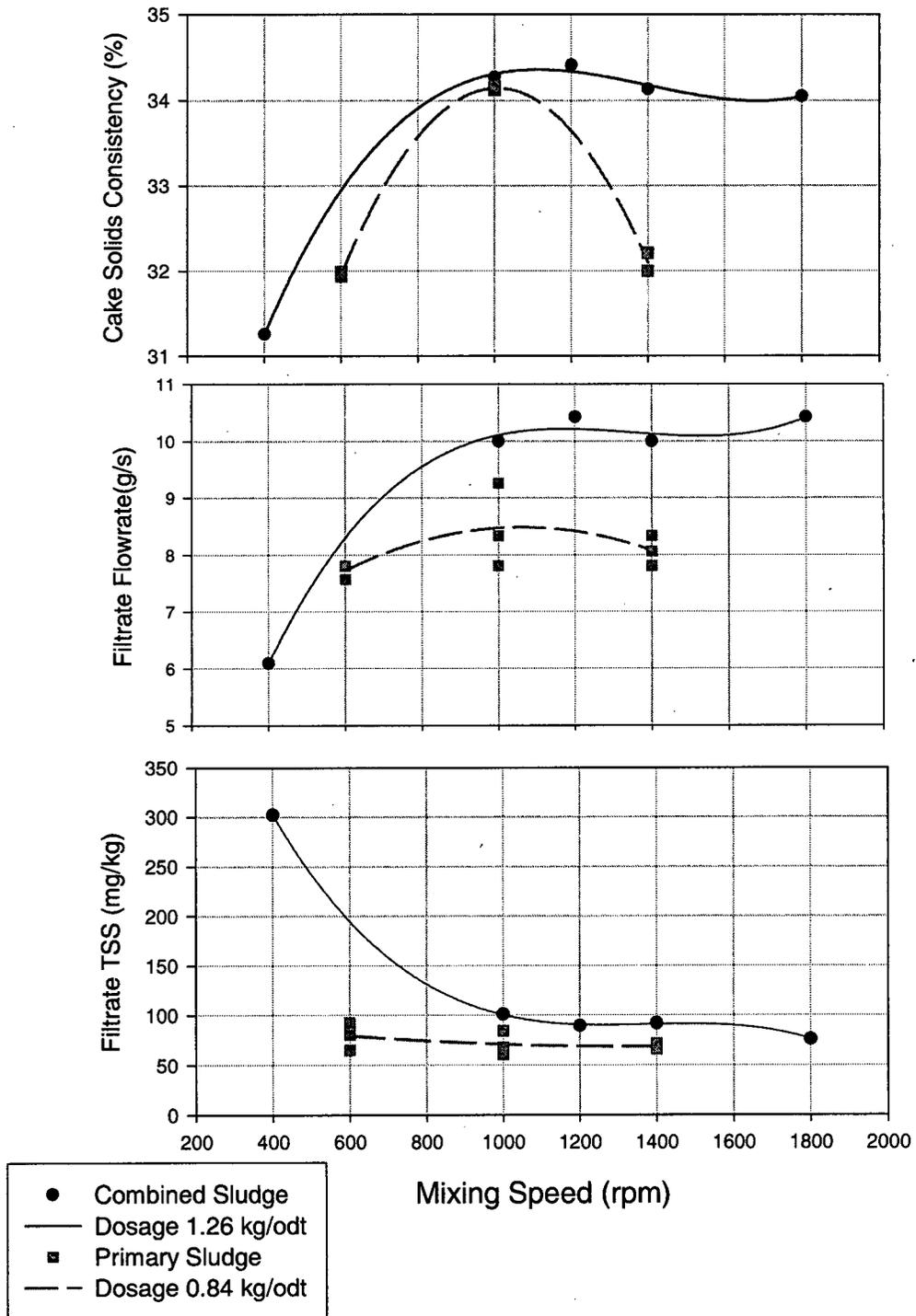


Figure 4-1 Effect of mixing speed on deinking sludge dewatering

In these mixing condition tests, although a lower polymer dosage was applied to the primary sludge, primary sludge still had a lower filtrate TSS than combined sludge. The reason for this could be that there were more small particles in the secondary sludge and hence in combined sludge compared to primary sludge alone and that they were more difficult to flocculate onto larger particles. Thus rather than being retained in the cake they tended to be removed in the filtrate.

In all of the experiments concerned with deinking sludge, unless otherwise noted, polymer-sludge mixing conditions were stirrer speed = 1000 rpm and mixing time = 1-3 seconds.

The first step in the tests on dewatering of kraft mill sludge was again to determine the optimal rpm of the stirrer in the mixing vessel. The results are shown in Figure 4-2. From these it was again concluded that 1000 rpm was the best speed for mixing at the same mixing time 1-3 seconds. Note that in contrast to the deinking sludge tests of Figure 4-1 with kraft mill sludge the filtrate TSS increased with increased mixing stirrer rpm.

In previous work (Groves 1995, Guild et al. 1998) mixing of polymer and sludge was accomplished by adding the polymer solution to the sludge and pouring the mixture back and forth 12 times between two mixing beakers. For a comparison of these two modes of mixing (stirring or pouring) see Figures 4-3 and 4-4. The conditions for the deinking sludge and kraft sludge tests are listed in Table 4-2 and Table 4-3.

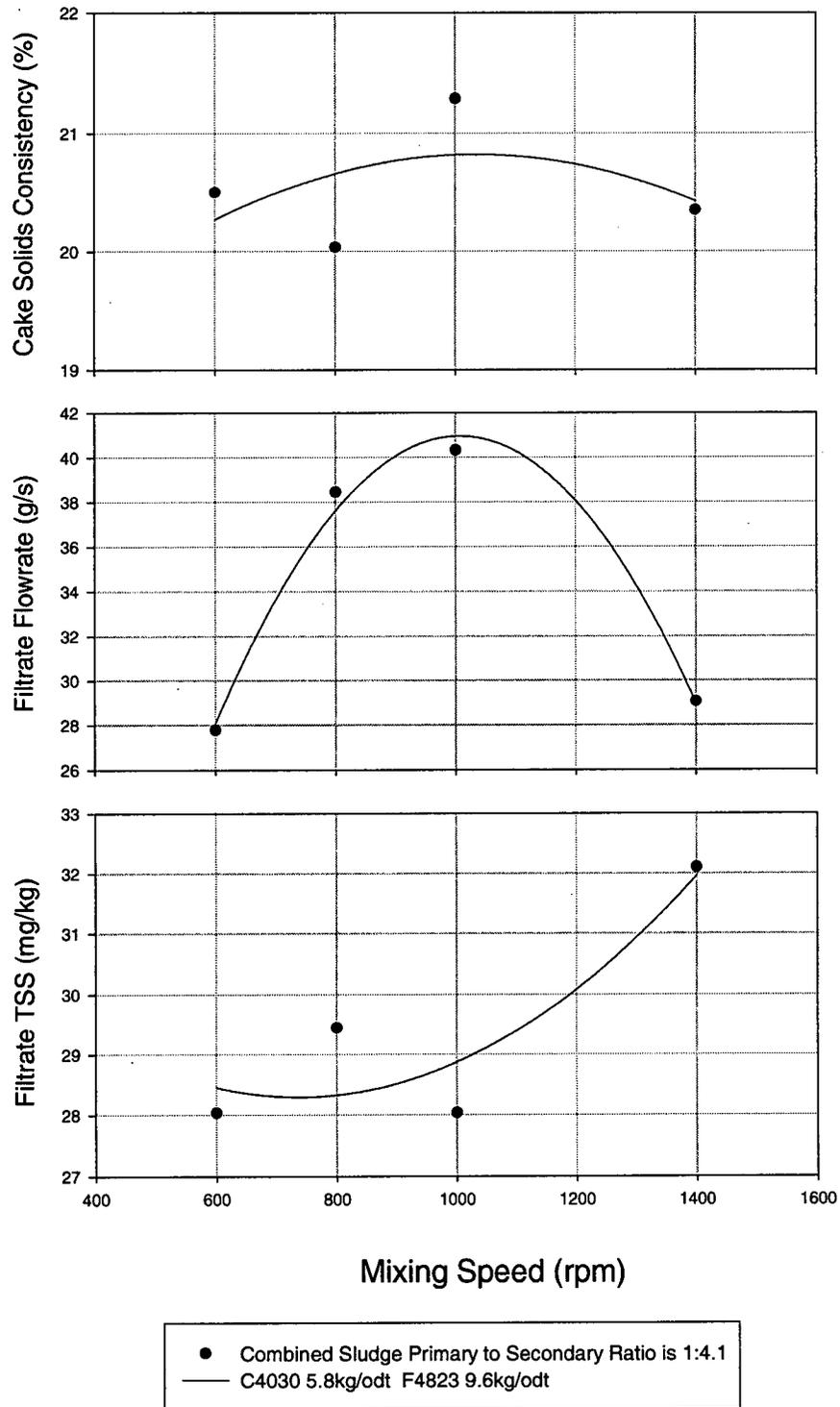


Figure 4-2. Effect of mixing speed on kraft mill sludge dewatering

Table 4-2 Conditions of deinking sludge mixing method tests

Test number	Sludge type	Polymer	Dose (kg/odt)	Air pressure (psi)
1	combined	Percol 734	1.27	60
2	combined	Percol 734	1.91	60
3	primary	Percol 734	0.84	75

Table 4-3. Conditions of kraft sludge mixing method tests

Tests	Primary to secondary ratio	Polymer	Dose (kg/odt)	Air pressure (psi)
1	1 : 4.1	C 4030	5.8	60
		F4823	9.6	
2	1 : 4.1	C4030	5.8	80
		F4848	9.6	
3	1 : 5.5	C 4030	8.5	55
		F 4848	11.3	
4	1 : 5.5	C 4030	11.3	55
		F 4848	11.3	

For the deinking sludges there appeared to be no significant difference between the two types of mixing in terms of the cake solids consistencies. For filtrate flowrate, stirring produced higher values than did pouring. Pouring resulted in higher filtrate TSS values in tests 1 and 3 and more or less the same value as stirring in test 2. Thus stirring at 1000 rpm was overall more effective than pouring.

For kraft mill sludges there was no significant difference between stirring and pouring in terms of cake solids consistencies. Pouring consistently resulted in lower filtrate TSS in all 4 tests. In the filtrate flowrate tests stirring produced higher flowrates than pouring in tests 1 and 2. In tests 3 and 4 the opposite was observed. Possibly this was because of the different flocculants used and/or the different primary to secondary ratio.

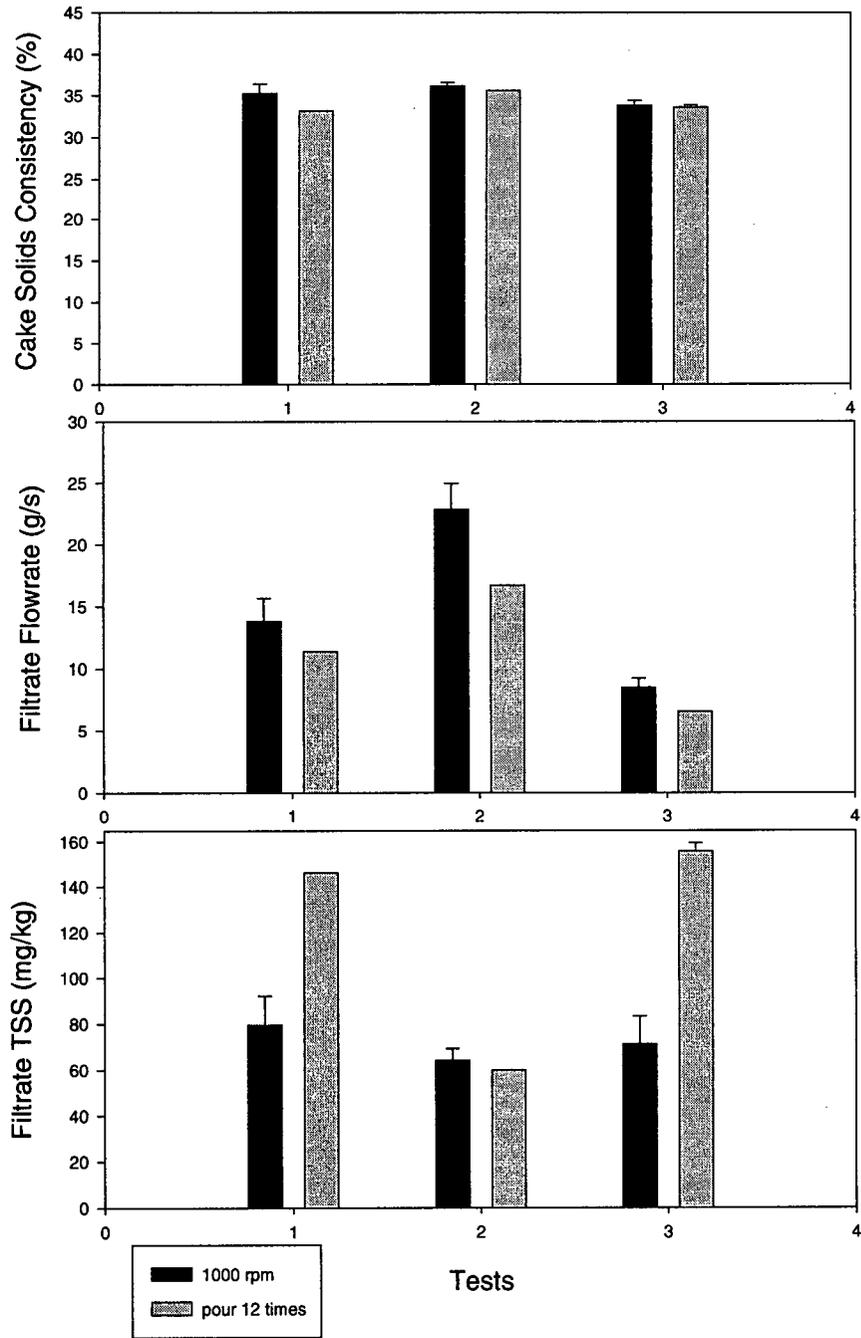


Figure 4-3. Results of deinking sludge mixing method tests

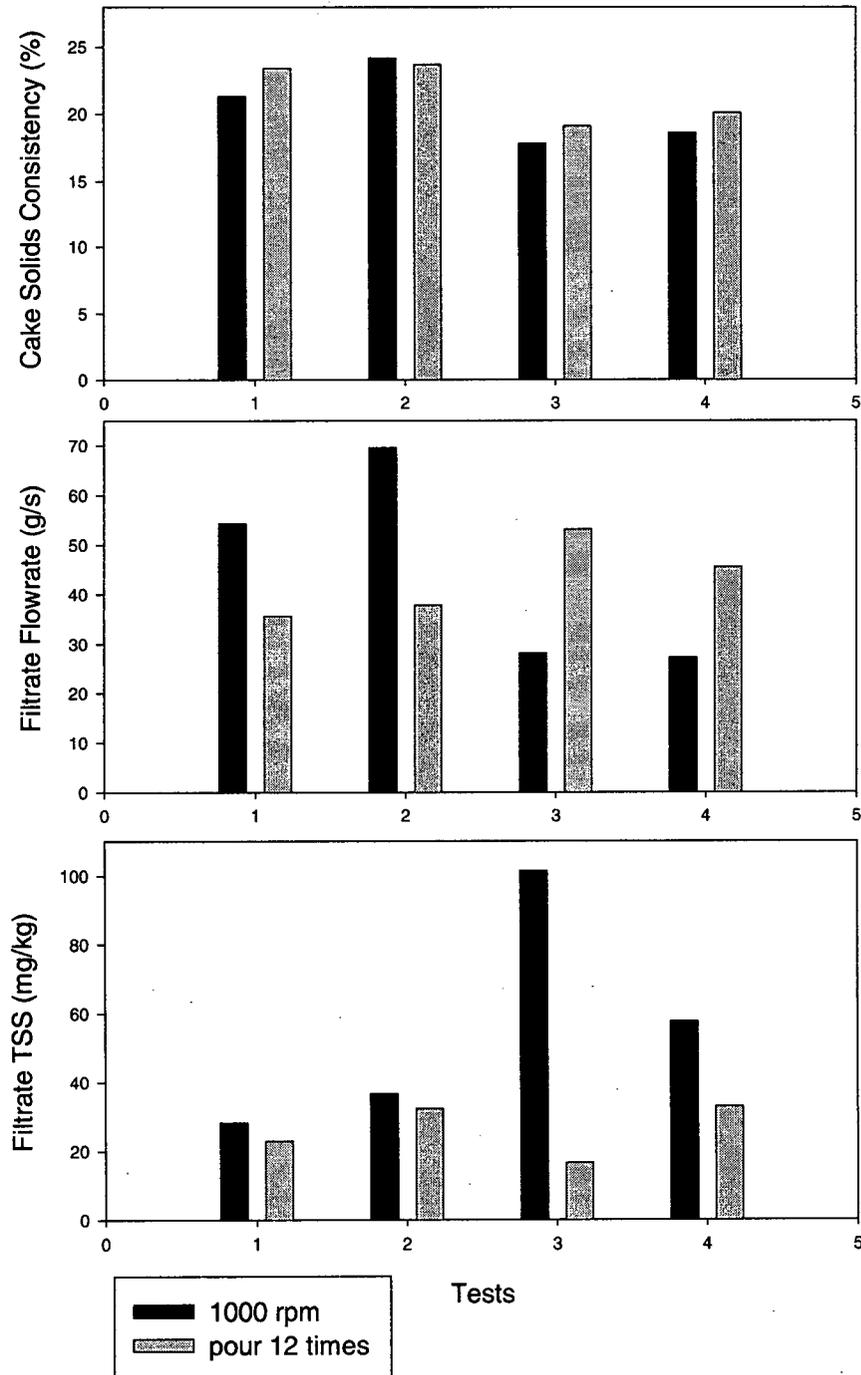


Figure 4-4. Results of kraft mill sludge mixing method tests

For kraft sludge, there was more secondary sludge than primary sludge in the combined sludge, thus there is more microbial cells in the sludge. When mixing by pouring, it was observed that the flocs deformed to a more compact structure and there was more water released from the flocs. Also the small flocs tend to join in the big flocs by pouring.

In the mills, some use pipe-mixing for polymer-sludge mixing, instead of that, some others add a stirrer in the line to improve the mixing efficiency. When sludge and polymer are mixed by a stirrer, the polymer can be dispersed more effectively. Pouring probably simulated pipe-mixing. Another trial on kraft mill combined sludge (primary to secondary ratio = 1:5.5) was done, the results of which appear in Table 4-4 to compare mixing by pouring, as described above, with stirring at 1000 rpm for 1-3 s followed by pouring back and forth six times. The latter treatment was to combine any beneficial effects attributable to each technique. The temperature was 21°C, polymer dosage was C4030 10.4 kg/odt, F4848 10.4kg/odt, and air pressure was 551.6 kPa (80 psi). Table 4-4 indicates that in terms of cake solids consistency no significant difference was found. However mixing by stirring followed by pouring did result in lower filtrate TSS.

Based on the test results above, the mixing method chosen for kraft mill sludge with primary to secondary ratio of 1:2.4 (last batch of sample) was stirring at 1000 rpm for 1-3 seconds, then pouring 6 times.

Table 4-4. Comparison of mixing of combined kraft mill sludge by pouring and mixer+pouring

Mixing Method	Series	Cake Solids Consistency (%)	Filtrate TSS (mg/kg)
Pour 12 times	1	20.88	45.58
	2	22.36	50.02
	3	22.09	38.41
	4	21.49	42.64
	5	20.68	45.10
	Average	21.50	44.35
Stirrer 1000rpm + Pour 6 times	1	21.77	30.21
	2	21.89	30.15
	Average	21.83	30.18

### 4.3 Pressure Effects

The effects of the pressure applied to the sludge during dewatering of deinking sludges (primary and combined) are shown in Figure 4-5. Both the primary and combined sludges exhibited similar trends. With an increase in pressure, the sludge cake solids consistency increased as more free and interstitial water was squeezed out. The filtrate total suspended solids (TSS) also increased a little. At higher pressures, the more fragile particles could have deformed and passed through the press screen. There was no obvious increase in filtrate flowrate.

Guild (1998) found that in pressing combined kraft mill and paper mill sludge cake solids consistencies increased with increased applied press pressure. The observed increase levelled off as the applied air pressure reached 482.65 kPa (70 psi), this is equalent to 222.02 kPa (32.2 psi) actual pressure applied to the sludge.

These results also agree with those found by Sorensen (1997) in that above a critical filtration pressure (2-20 kPa for activated sludge), filtrate flowrate is independent of pressure for highly compressible cake structures (for more on deinking combined sludge compressibility see Section

4.12 below). This critical pressure is well below the filtration pressure used in industrial processes, which is usually in the range of 100 –1000 kPa, and well below the pressures indicated on Figure 4-5.

For deinking sludge dewatering, even at the highest pressure which could be obtained in the lab sludge press the cake solids consistency for primary sludge (39 %) could not achieve the 50% cake solids measured by the mill from which the sludge samples were obtained, in their screw presses. Perhaps this is because the pressure applied in laboratory sludge press was not as high as the pressure (unknown and not measured) developed in the screw presses used in the mill. However Figure 4-5 indicates that further realistic increases in pressure would not substantially increase cake solids consistency. Also unknown are typical values for residence time used in a mill screw press. The laboratory press residence time used in these tests was 2 minutes. Figure 4-6 below indicates that substantial increase in cake solids consistency could be achieved if the pressing time was longer than 5 minutes.

Note that in Figure 4-4 the cake solids consistencies were higher in test 2 than test 1 (done at air cylinder pressures of 80 and 60 psi). Thus higher pressure tended to produce higher sludge cake consistencies for kraft mill sludge too. The sludge cake consistencies achieved for kraft mill combined sludge were in the range of 17-24%. In the mill typical sludge cake solids were 30% (with the addition of heat).

#### **4.4 Effects of Pressing Time**

Previous workers (Groves 1996, Guild et al 1998) using the same laboratory sludge press rather arbitrarily chose a pressing time of 2 minutes. To see if this time period was appropriate we did further testing to find the effect of pressing time on cake solids consistencies. The results are shown in Figure 4-6. The sample was combined kraft mill sludge; the primary to secondary ratio was 1:4.1. The air pressure was 551.6 kPa (80 psi). The effects of the different polymer combinations used in these tests will be considered in Section 4.9.2.1 below.

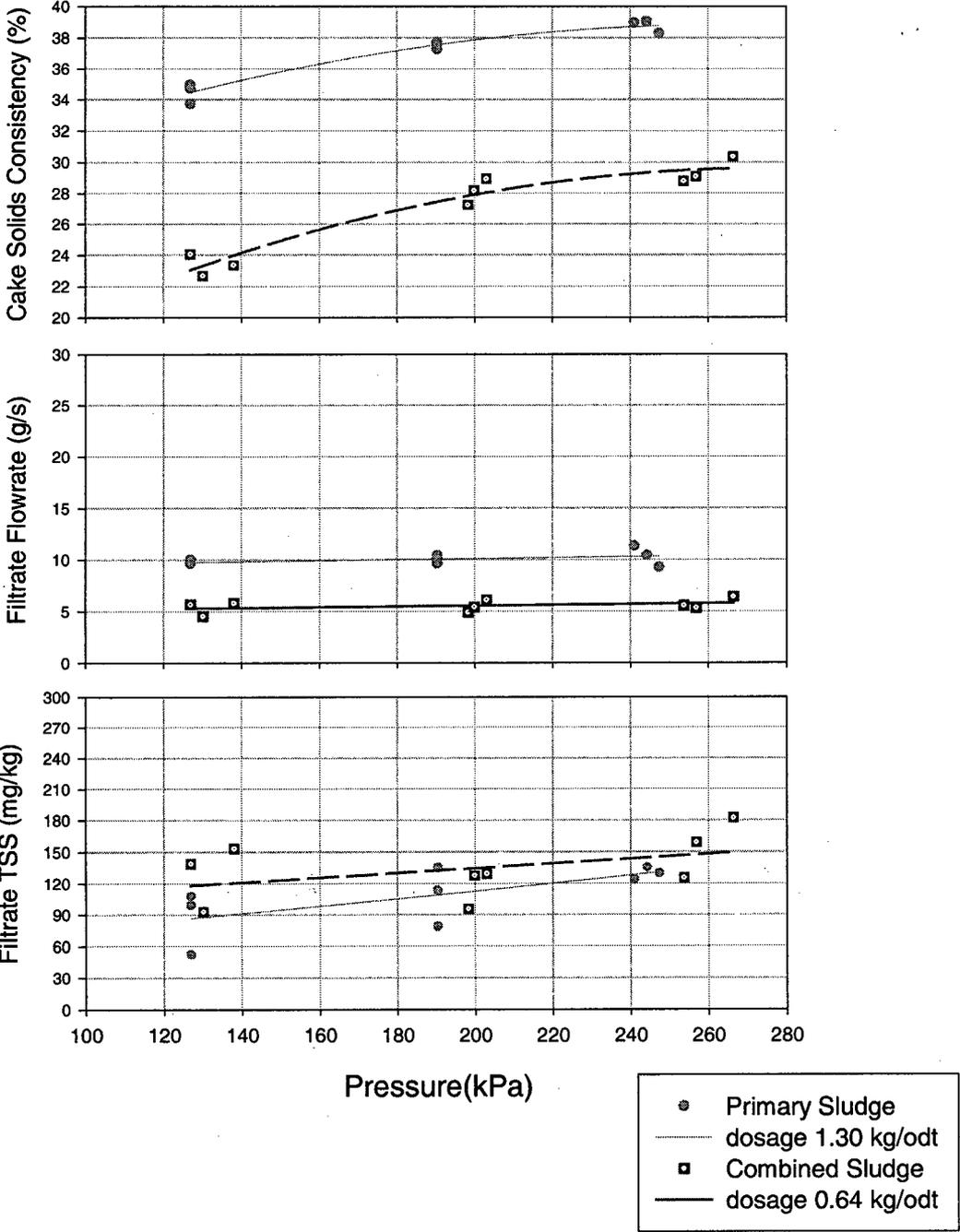
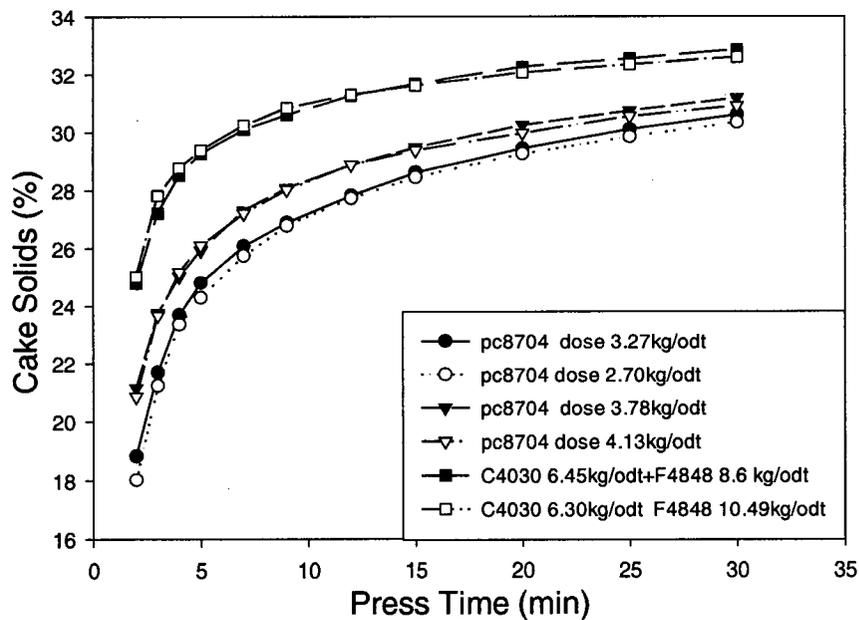


Figure 4-5 Effect of pressure on deinking sludge dewatering

With increased pressing time, cake solids consistencies increased rapidly up to about 10 minutes then increased more slowly. All the polymers used at various dosages had similarly shaped profiles of cake solids consistency vs. pressing time. The ranking order of cake solids consistencies at 2 minutes is almost the same as those observed at longer pressing times. So 2 minutes pressing time is suitable pressing time for laboratory sludge dewatering tests when the objective of the tests is to compare the performance of various coagulants and flocculants. However if the objective were to increase cake solids consistency a press time longer than 2 minutes would be better.



**Figure 4-6 Effect of press time on cake solids consistency**

#### **4.5 Temperature Effects**

Figure 4-7 shows the effect of temperature on the dewaterability of deinking primary sludge. With a temperature increase, the cake solids and filtrate flowrate increased linearly, but the filtrate TSS didn't show any obvious changes. As the temperature goes up, the viscosity of water would decrease, so at higher temperatures it would be easier for water to pass through the press cake.

At 21°C the viscosity of water is 1002 Pa s, at 50°C it's 547 Pa s. According to equation 3-2, the filtrate flowrate (w/t) is inversely proportional to filtrate viscosity. Thus if the observed increase in filtrate flowrate evident in Figure 4-7 was solely attributable to a change in viscosity (assuming that the filtrate viscosity = water viscosity or has the same temperature dependence).

$$\text{Flowrate @ 21}^\circ\text{C} \times \text{viscosity @ 21}^\circ\text{C} / \text{viscosity @ 50}^\circ\text{C} = 9 \text{ (g/s)} \times 1002/547 = 16.5 \text{ g/s}$$

The observed flowrate at 50°C was 13 g/s. Thus much of the difference between 21°C and 50°C could be attributed to filtrate viscosity differences. However the actual increase was less than the predicted increase.

Guild (1998) has reported that for a kraft mill combined sludge (primary to secondary ratio 2.9:1) conditioned with 4 kg/odt of coagulant (a cationic polyamine) and 3 kg/odt of flocculant (a cationic polyacrylamide) as the temperature rose from 20°C to 60 °C, the filtrate flowrate increased linearly

from 1.1 L/min to 1.47 L/min, the cake solids consistency increased non-linearly from 22.5% to 28.7%.

$$\text{Flowrate @ 20}^\circ\text{C} \times \text{viscosity @ 20}^\circ\text{C} / \text{viscosity @ 60}^\circ\text{C} = 1.1 \text{ (L/min)} \times 1002/466 = 2.36 \text{ L/min}$$

Thus in this case the predicted flowrate was much larger than the observed flowrate. Viscosity effects do not seem to be very important in the analysis of Guild's data with respect to temperature effects. His flowrate range (1.1-1.47 L/min) was greater than that shown in Figure 4-7 (0.54-0.78 L/min), so possibly inertial effects, not included in the filtration equation 2-10 were significant. It's also quite possible the compressibility effects were temperature dependent since the cake solids consistencies, a measure of cake compressibility, were higher for both types of sludge at higher temperatures.

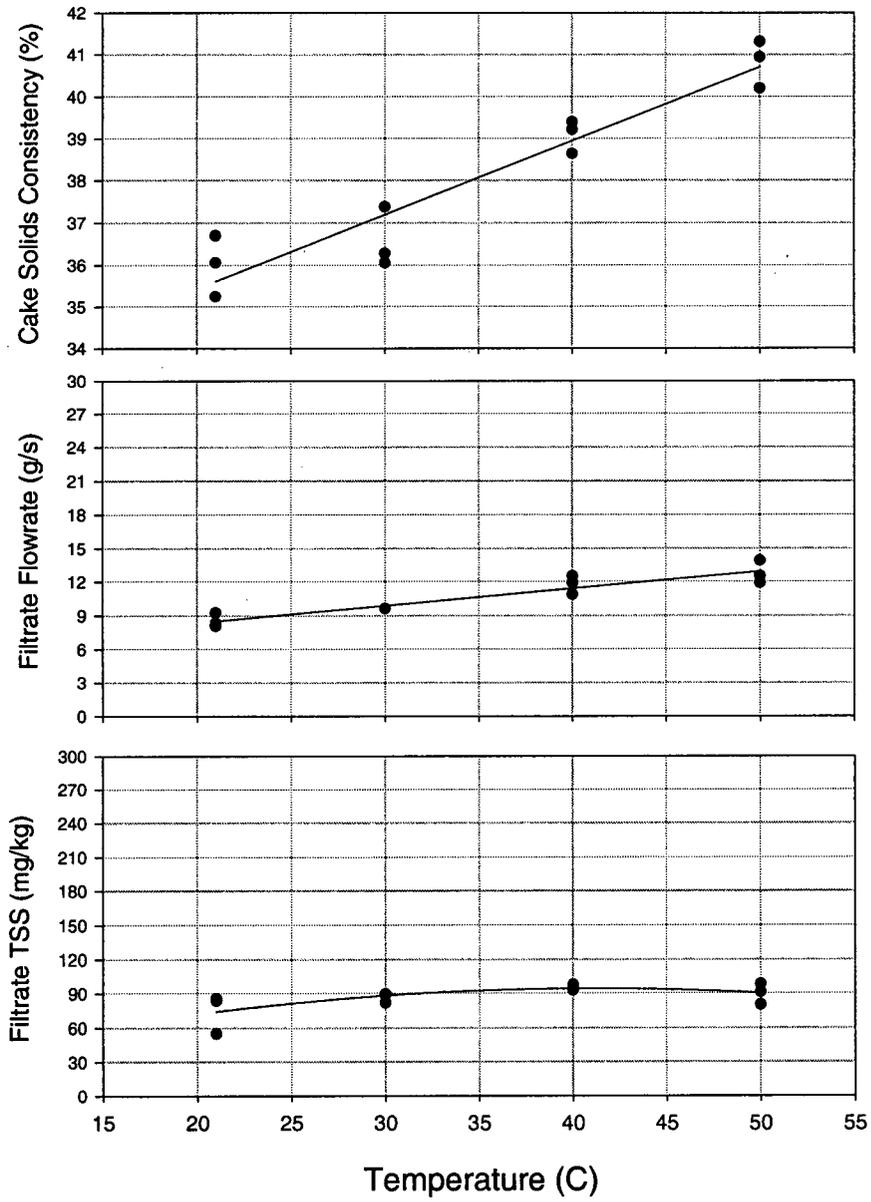


Figure 4-7 Temperature effect on deinking primary sludge dewatering

#### **4.6 Effects of Primary Sludge to Secondary Sludge Ratio**

Figure 4-8 shows the effects of varying the primary to secondary ratio (mass ratio based on dry solids) on deinking sludge dewatering. A single polymer conditioning process was employed using Percol 734. The dosage was from 1.8 kg/odt to 2.1 kg/odt. Air pressure was 551.6 kPa (80 Psi). With an increase in the primary sludge fraction, the cake solids consistencies went up linearly and the filtrate TSS decreased, asymptotically until a constant value of 50-60 mg/kg was reached. There is no obvious difference in filtrate flow-rate .

In the pulp and paper industry, secondary sludge is always combined with the primary sludge for dewatering. Due to the fibrous nature of the primary sludge, the combination for dewatering is very effective. The presence of fibre tends to increase the porosity and decrease the compressibility of the cake thereby providing an easier path for water to go through. The longer the fibre, the better results of sludge dewatering would be. But fiber is also a valuable resource, which can be recycled. So the tendency in mill operations is to reduce as much as possible the amount of primary sludge discharged which means there would be a lower ratio of primary to secondary sludge.

Can a secondary sludge be dewatered by itself or is there any other method to enhance the secondary sludge dewaterability? The following tests try to answer these questions.

For polymer conditioned, deinking secondary sludge, large flocs were observed to form in the mixing vessel but a suitable cake could not be formed in the lab sludge press no matter how much polymer was added. Secondary sludge has a very high compressibility. The laboratory sludge press is usually operated at an air pressure of 413.7 kPa (60 psi), which produces an actual pressure applied to sludge of 190.3 kPa (27.6 psi), this pressure is much higher than the critical pressure 2 –20 kPa for activated sludge (Sorensen,1997).

According to Sorensen and Hansen (1993), in a dewatering situation where the cake thickness typically is a few centimeters, the main part of the liquid pressure drop in the cake occurs in a few tenths of millimeter, just above the medium. In our tests, it appeared that the flocs that

formed were very soft and compressible. At 413.7 kPa (60 psi) constant air pressure when dewatering the deinking secondary sludge, after several seconds of filtration, a thin layer of cake formed just above the screen, this layer of cake was so compact that the water above the layer hardly had any way of getting through the formed cake, therefore, no final sludge cake could be formed.

#### **4.7 Sawdust and Hog Fuel as Filter Aids**

In order to retain permeability during sludge dewatering, physical conditioners (filter aids) can be used to deal with compressible sludge cake. They can provide void space and reduce compressibility thus making it easier for water to get through the cake. Smollen et al. (1997) used pyrolysed domestic refuse (char) to increase the structural strength and permeability of biological sludges. This gave us the idea that other wastes from the pulp and paper industry might be used to enhance secondary sludge dewaterability. We tried two kinds of such waste products: sawdust and hogfuel as filter aids. These materials are relatively incompressible and should be readily and cheaply, available at many mill sites. Both of them were effective in enhancing secondary deinking sludge dewaterability as indicated in Figure 4-10. The polymer conditioner was Percol 734 at a dosage of 1.74 kg/odt.

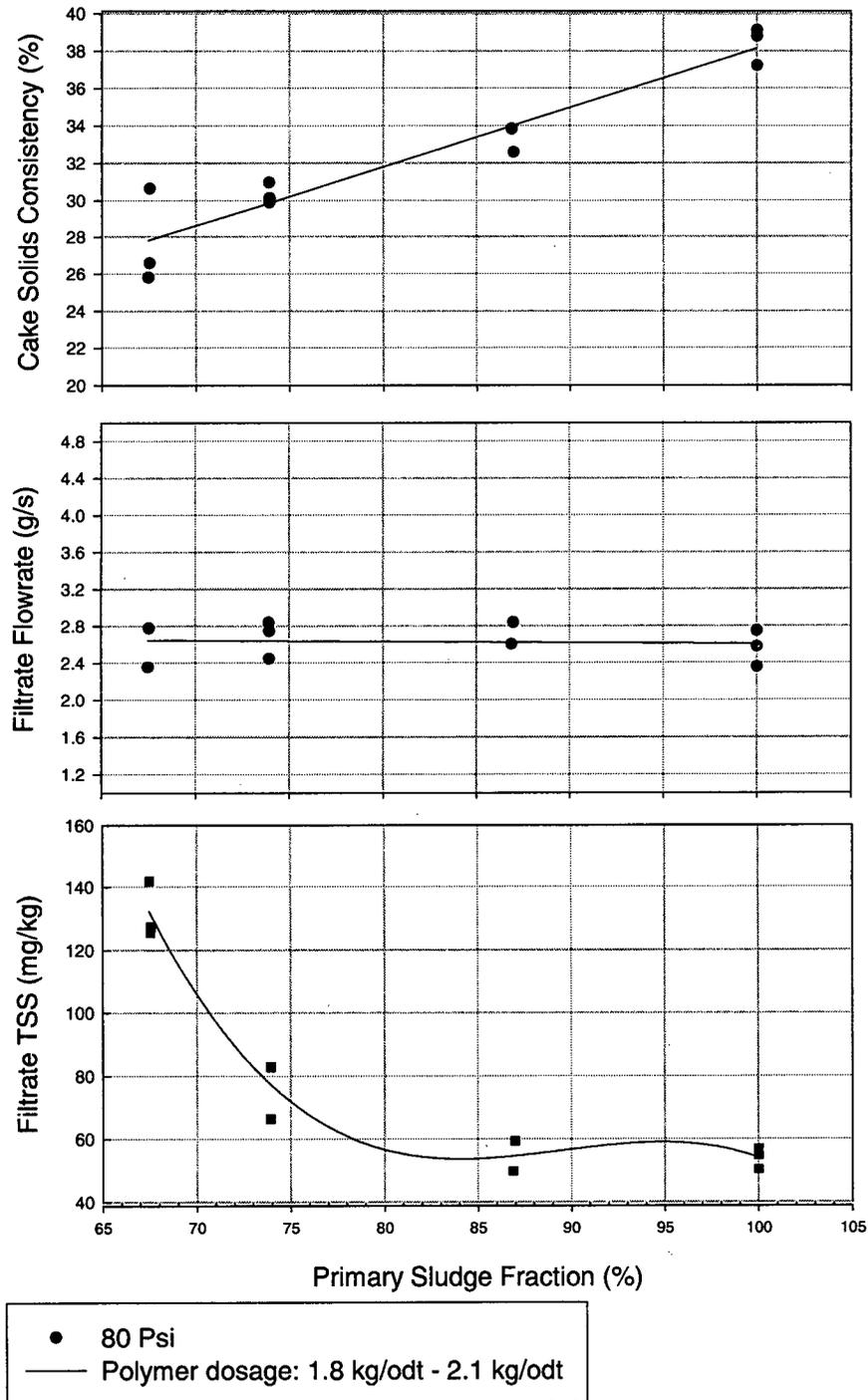


Figure 4-8. Effect of primary sludge fraction on deinking sludge dewatering

#### **4.7.1 Addition Sequence of Polymer and Sawdust**

There are two sorts of sequence to add polymer and filter aids, adding filter aids first then polymer or conversely. Figure 4-9 compares the deinking secondary sludge dewatering results of these two methods. The Percol 734 dosage was 2.70 kg/odt, the sawdust fraction was 35%. It is obvious that when sawdust was added after polymer addition higher cake solids and faster filtrate flowrate was achieved. This also indicated that filter aids could facilitate the dewatering process. Due to the large standard deviation associated with filtrate TSS, no significance can be attached to any differences. In the following tests, filter aids were added into already conditioned sludge.

#### **4.7.2. Sawdust and Hog Fuel Working as Filter Aids**

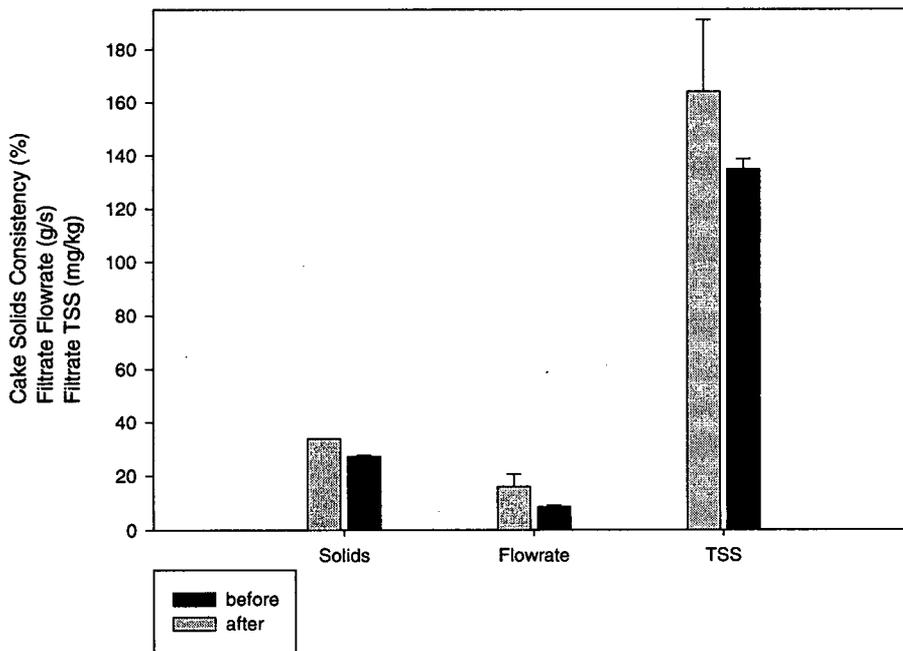
As the amount of filter aid supplied increased from 20% to 40%, cake solids increased from 28% to 35% for sawdust and from 30% to 33% for hogfuel. Filtrate flow rate increased and filtrate TSS decreased. Generally speaking sawdust performed better than hog fuel probably because the smaller particles of sawdust could be more homogeneously mixed with sludge. However these filter aids could not take the place of the polymer, nor even reduce the necessary polymer dosage (see Figure 4-11). Merely combining these filter aids with secondary sludge did not result in the sludge press being able to form a cake of solids.

Also note that the amounts of sawdust or hog fuel necessary to achieve a reasonable value for cake consistency were excessively high. If the final disposal of sludge is incineration on site, at the mill, and the final use of hog fuel is fuel for the incineration device this excessively high requirement probably doesn't matter since the sludge and hog fuel are ultimately all going to the same place for disposal. The case for sawdust is less clear since it has some value as raw material for pulping.

A later test was done to investigate the behaviour of hog fuel alone in the sludge press. An initial slurry of hog fuel in water was made at a solids concentration of 1.6%, which was similar to the solids content of secondary sludge slurries prior to dewatering. After treatment in the sludge press at an air pressure of 413.7 kPa (60 psi) the cake solids consistency was 21%, which is

considerably less than the values achieved in pressing various combination of sludge and hogfuel. Thus when hogfuel alone was dewatered, it didn't compress too much due to its rigid nature, thus a considerable amount of water remained in the void spaces in the filter cake.

The solids contents of sawdust and hog fuel were 95% and 73%. After dewatering with sludge, they were 28% -35%. So there is more water adsorbed into the filter aids.



**Figure 4-9 Deinking combined sludge dewatering results for sawdust addition before and after polymer added**

Some effects of the use of sawdust as filter aid with deinking sludge are shown below in Figure 4-11, where it can be seen that addition of sawdust makes it possible to press secondary sludge to reasonable consistencies. But in terms of cake consistency addition of 32% sawdust was not as effective as adding primary sludge (primary to secondary ratio is 2.9:1).

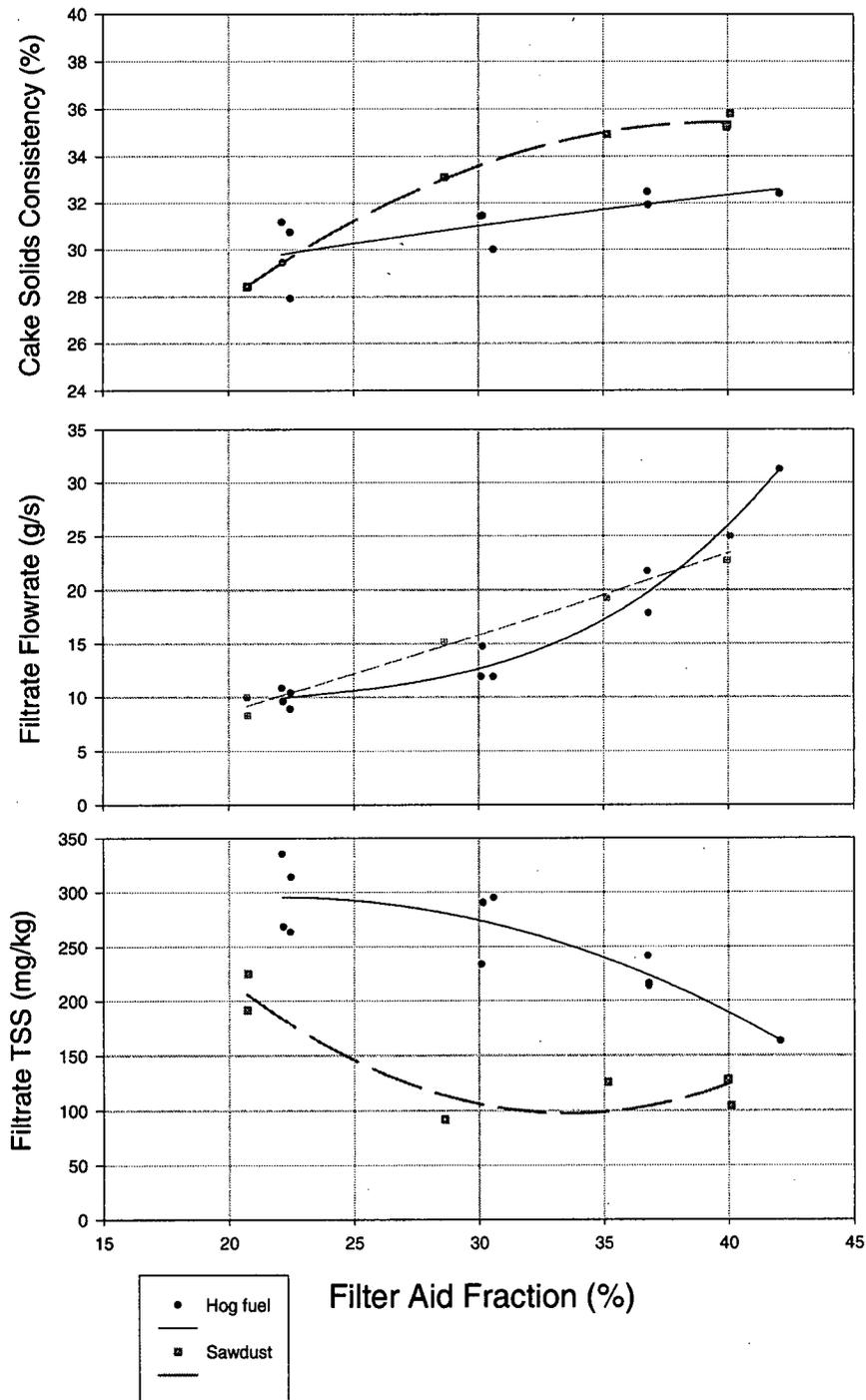


Figure 4-10. Effect of filter aids on deinking secondary sludge dewatering

## **4.8 Polymer Dosage Effects**

### **4.8.1 Deinking Sludge**

Polymer dosage effects on deinking sludge dewatering are shown in Figure 4-11. The polymer used was Percol 734, which was used by the mill from which the sludge was obtained. The optimal dosages, in terms of cake solids consistency, for the various sludge types can be determined from the topmost plot. There it can be seen that there is an optimum dosage for each type of sludge to achieve maximum cake solids. All of the optima are rather flat, thus near the optimum changes in polymer dosage result in only small changes in cake solids. There were few differences among the three types of sludge. The optimum polymer doses for cake solids consistency of the three sludges are in the range of 1.3 – 1.8 kg/odt. The uppermost plot of Figure 4-11 also tells us that the highest cake solids were achieved with the primary sludge, followed by the combined sludge. The cake solids of secondary sludge with sawdust as an additive filter aid was the lowest.

The middle plot of Figure 4-11 shows the polymer dosage effect on filtrate flowrate. With the dosage increasing, the filtrate flowrate increased much more so for the combined sludge and secondary sludge with sawdust.

The bottom plot of Figure 4-11 shows the polymer dosage effect on the filtrate TSS. From this figure we can see that with an increase in the polymer dosage, the filtrate TSS decreased. The lowest TSS occurred with primary sludge, then combined sludge and finally secondary sludge with sawdust as an additive. This is because the secondary sludge had smaller and more fragile particles thus it was easier for them to go through the press screen and so contribute to filtrate suspended solids.

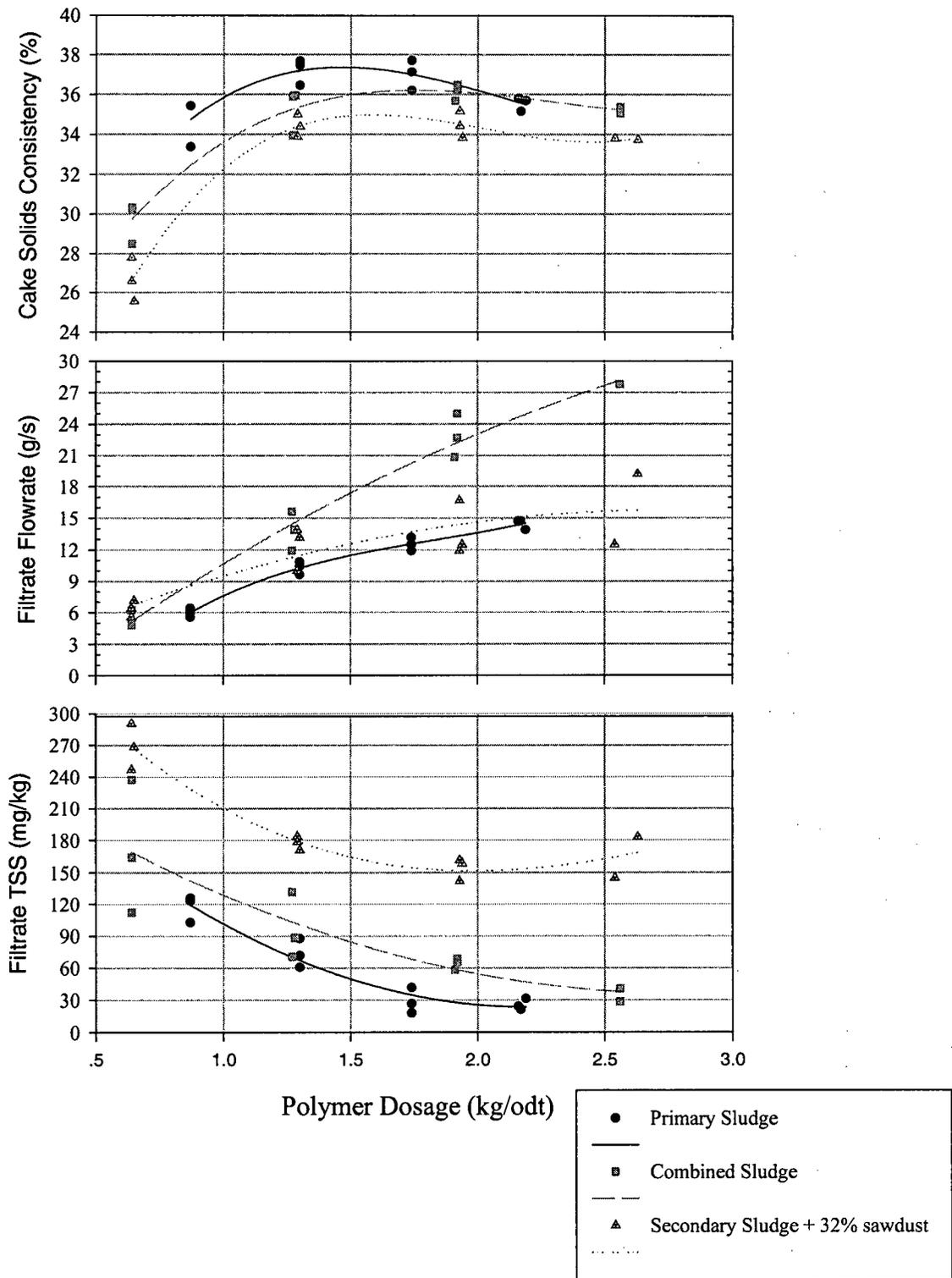
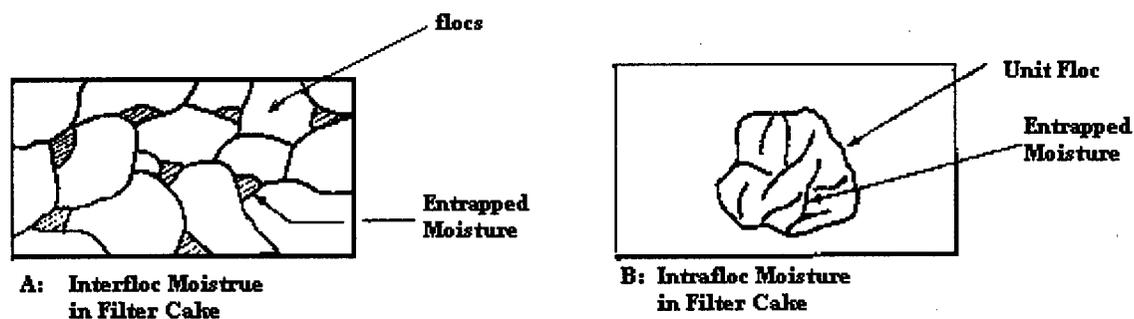


Figure 4-11. Dosage effect on deinking sludge dewatering

In Figure 4-11 it can be seen that the optimum polymer dose for cake solids consistency is in the range 1.3 – 1.8 kg/odt, but the filtrate flowrate continues to increase as the polymer dosage increases, without showing an obvious maximum. The filtrate TSS has a minimum at a polymer dose of around 2.0 kg/odt for the secondary sludge plus sawdust while for primary sludge and combined sludge the TSS tended to decrease with increases in polymer dose. That these opposing trends occur in case of filtrate flowrate and cake solids consistency may be attributable to different dewatering mechanisms (see Section 4-11): filtrate flowrate characterizes filtration stage, cake solids consistency is the result of both filtration and compression.

Mishra (1988) described two modes of liquid flow through a flocculated filter cake: interfloc flow and intrafloc flow (Figure 4-12).



**Figure 4-12. Inter and intra floc moisture content**

While increased flocculation of particulates may improve interfloc flow, which means a faster flowrate, due to stronger binding of the flocs there could be increased resistance to intrafloc drainage of the moisture. This could explain why the cake solids consistency decreased, while the filtrate flowrate was still increasing.

#### **4.8.2 Kraft Mill Sludge: Single Polymer Flocculant**

In combined primary and secondary kraft mill sludge conditioning sometimes a single flocculating polymer is used and sometimes a dual polymer system is utilized. In the latter a

polymeric coagulant is employed to reduce the charge density on the sludge particles followed by a flocculant polymer to cause flocculation. In this study the single polymer flocculant used was Perform PC8704, which the mill was using. The dosage effects of this single polymer are shown in Figure 4-13.

The first plot is cake solids consistency vs. flocculant dosage. When the flocculant dosage increased, the cake solids consistency increased at first. When the dosage was over 6 kg/odt, the cake solids didn't increase any more, but remained at an almost constant value of 21.8%. So the optimal dosage seems to be 6 kg/odt, which is the lowest dosage giving the highest cake solids.

The second plot is filtrate flowrate vs. flocculant dosage. When the flocculant dosage was increased, the filtrate flowrate increased almost linearly. Thus no optimal dose was evident from this plot other than to say, within the range tested, the higher the dosage the higher the filtrate flowrate. Again this is probably a result of the differences between filtration and compression dewatering mechanisms.

The third plot is filtrate TSS vs. flocculant dosage. As the flocculant dosage increased, the filtrate TSS decreased then leveled off at a low level around 50 mg/kg.

Considering the filtrate flow rate the more polymer that is added the greater the filtrate flowrate as seen in the middle plot of Figure 4-13. But considering the cake solids there is little to be gained by adding more than about 6 kg/odt of polymer. The same is true for the filtrate TSS. Thus in determining polymer dose it is important to decide which dependent variable it is most important to optimize.

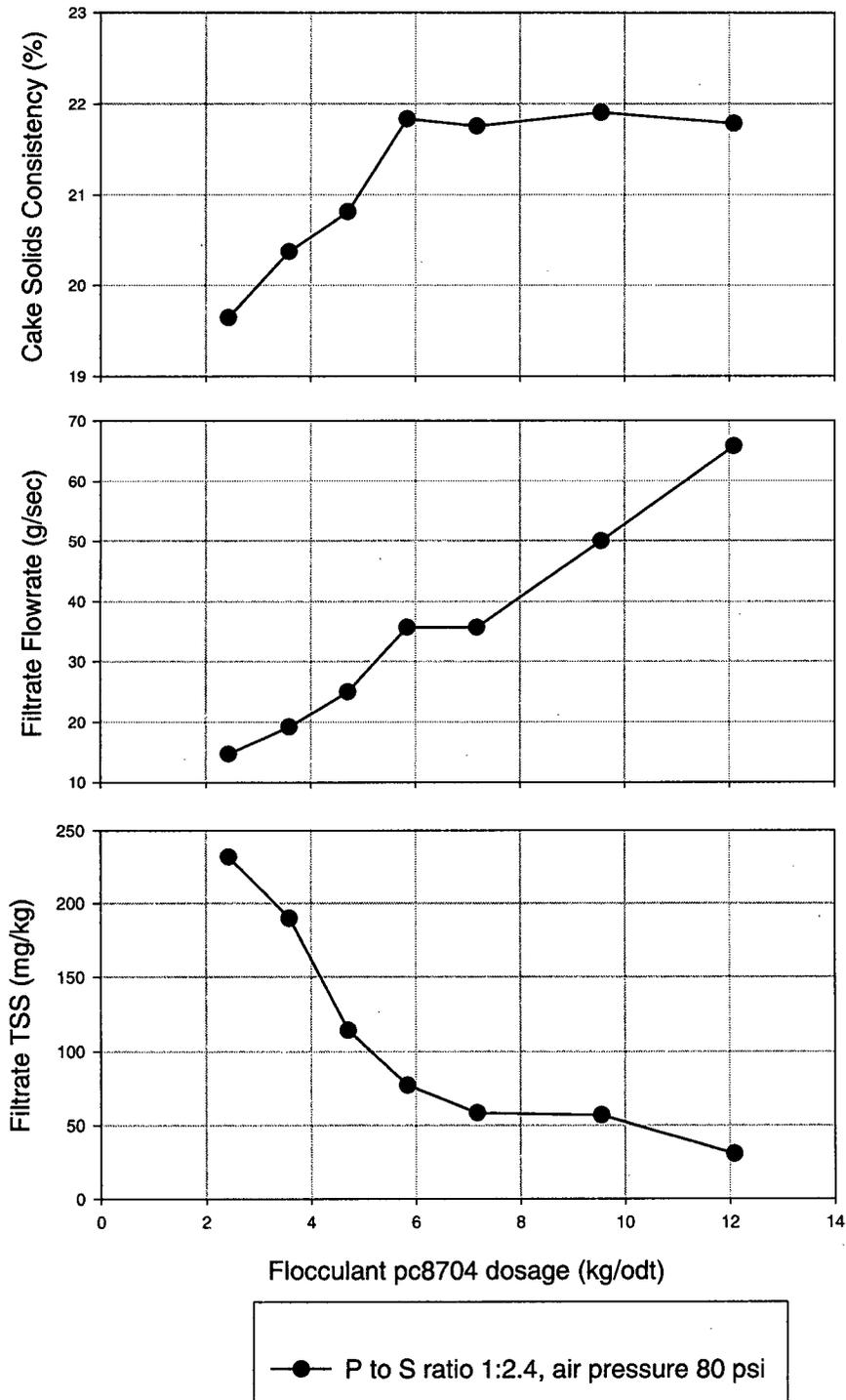


Figure 4-13. Single polymer dosage effect on kraft mill combined sludge dewatering

### **4.8.3 Kraft Mill Sludge: Dual Polymer System (Coagulant and Flocculant)**

Coagulant and flocculants were supplied by Callaway Chemical Co. The coagulant applied was C4030. There were three flocculants that were selected for use in these tests, F4823, F4848 and F4330. F4823 and F4848 are positively charged polymers; F4330 is negatively charged polymer. Since F4330 couldn't form big flocs due to its negative charge, only F4823 and F4848 were applied in the following tests.

Preliminary tests were done try to find a beginning coagulant dose and flocculant dose. It was easier to find a suitable flocculant dose since once the flocculant dosage went over a certain value, large flocs were observed to occur and the sludge could then be dewatered well.

Thus a fixed flocculant dose was chosen from the preliminary tests, then the coagulant dosage was varied to try to find the optimal coagulant dosage. The results for these coagulant dosage experiments are shown in Figure 4-14.

From the first and second plots of cake solids consistency and filtrate flowrate vs. coagulant dosage, it was easy to determine the optimal coagulant dosage for kraft combined sludge dewatering. But the optimal dosages for these two parameters were not completely consistent. From the three plots of Figure 4-14 3-6 kg/odt was chosen to be the optimal coagulant dose for the sludge with a primary to secondary ratio of 1:2.4, considering there was some variation in the measurements and the best values for cake solids consistency, filtrate flowrate and filtrate TSS. For the sludge with a primary to secondary ratio of 1:5.5 the best values was 8-11 kg/odt. Different air pressures in the range 55 to 80 psi didn't affect the determination of the optimal dosage (see Section 4.8.4). Further increases in the coagulant dosage above this optimal dosage, resulted in lower filtrate flowrates and cake solids consistency. At these higher dosages it would appear that overdosing was taking place. It seems that the sludge with lower secondary sludge fraction had a lower optimal coagulant dose, in our case, 3-6 kg/odt for sludge with P to S ratio of 1:2.4, 8-11 kg/odt for sludge with P to S ratio of 1:5.5. The third plot shows that as the coagulant dose increased, the filtrate TSS of the two kinds of combined sludge decreased then leveled off. The optimum ranges of polymer dose selected on the basis of sludge cake solids and filtrate flowrate which would also result in low level of filtrate TSS.

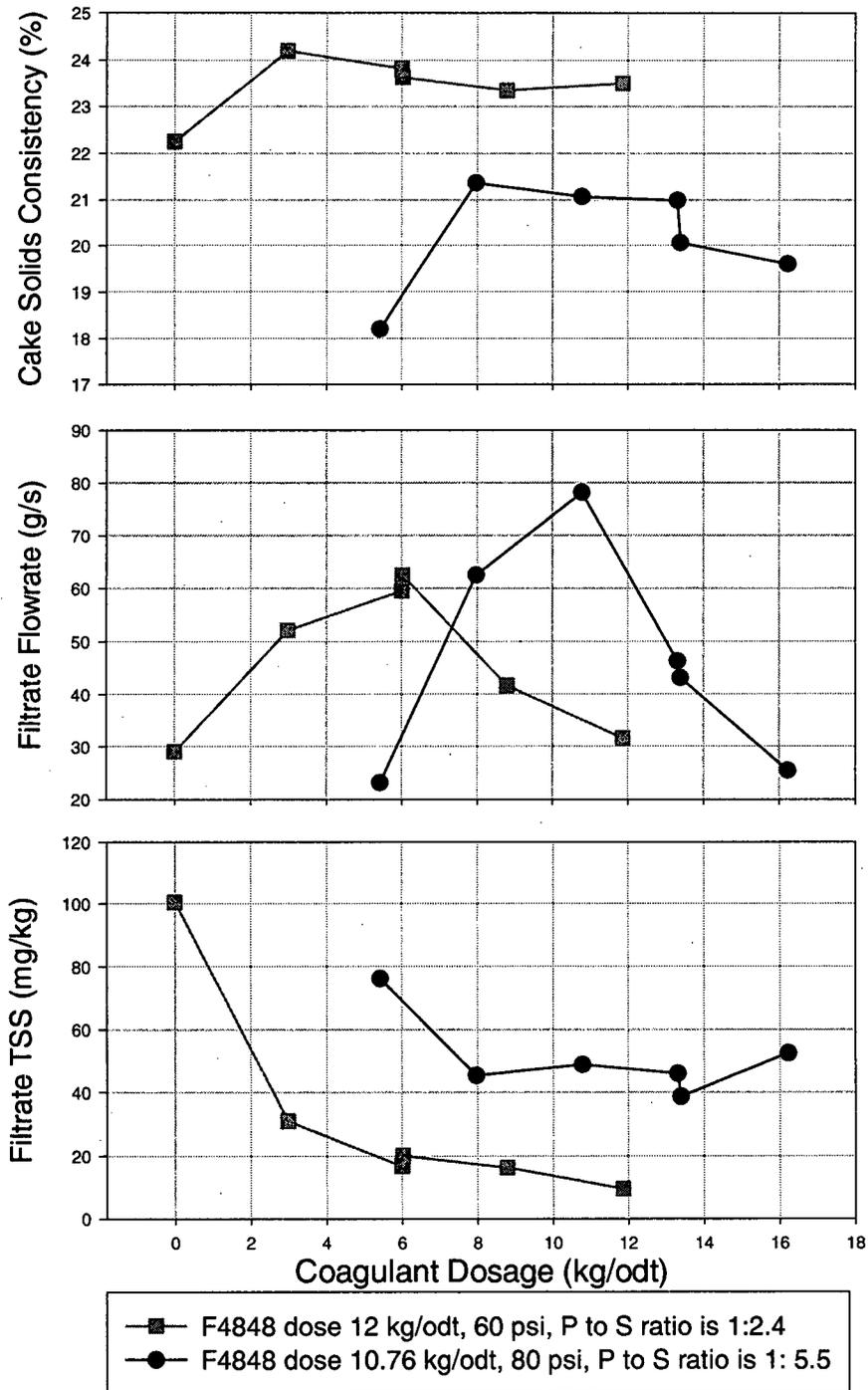


Figure 4-14. Organic coagulant dosage effect on kraft combined sludge dewatering

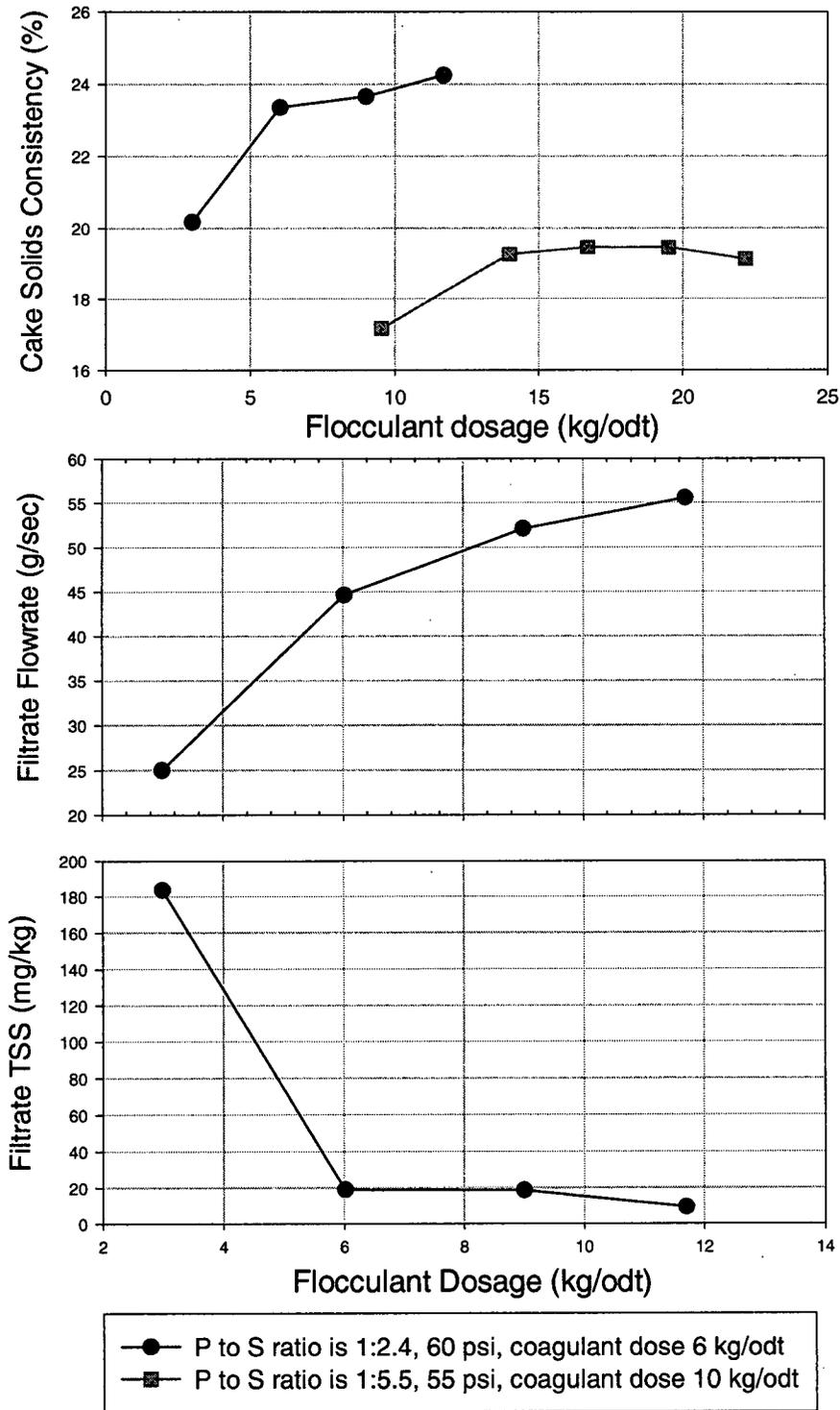


Figure 4-15. Flocculant dosage effect on kraft combined sludge dewatering

Next the coagulant dosage was set at 6 kg/odt for the combined sludge with P to S ratio of 1:2.4, and at 10 kg/odt for the combined sludge with P to S ratio of 1:5.5, the optimal doses obtained by the coagulant dosage test. Then the flocculant dosage was varied to try to find its optimal dosages. The results are shown in Figure 4-15.

The first plot is cake solids consistency vs. flocculant dosage. Both curves have similar trends. As the flocculant dosage rose, the cake solids consistency increased, rapidly at first, then the rate of increase slowed down or leveled off. With the lower secondary sludge fraction, the optimal flocculant dosage was lower, 6-12 kg/odt for sludge with P to S ratio of 1:2.4, compared to 14-19 kg/odt for sludge with P to S ratio of 1:5.5. Guild (1998) found similar results in his work that increasing the secondary sludge fraction would increase the coagulant and flocculant demand.

The second plot is filtrate flowrate vs. flocculant dosage. As the dosage increased, the filtrate flowrate increased. Above a dosage of 6 kg/odt, the rate of rise of flowrate with increasing dosage was reduced.

The third plot is filtrate TSS vs. flocculant dosage. As the flocculant dosage increased the filtrate TSS decreased rapidly, reaching an almost constant value for dosages higher than 6 kg/odt.

From Figure 4-14, we can see if the coagulant was overdosed, it was easy to detect due to the decreased flowrate, cake solids consistency, but flocculant overdosing was more difficult to detect, since the cake solids consistencies were almost the same and filtrate flowrate didn't decrease either. See Section 4.10 below for another approach to determine when overdosing is occurring.

#### **4.8.4 Pressure Effect on Optimal Coagulant Dosage**

In our experiments, we used various pressures, from 379.2 kPa (55 psi) to 551.6 kPa (80 psi). How would such pressure changes affect the optimal polymer dosage? We did some tests on kraft combined sludge (primary to secondary ratio 1:2.4), the results of which are shown in Figure 4-16. The results proved that different pressures didn't affect the optimal dosage of the coagulant in terms of filtrate flowrate.

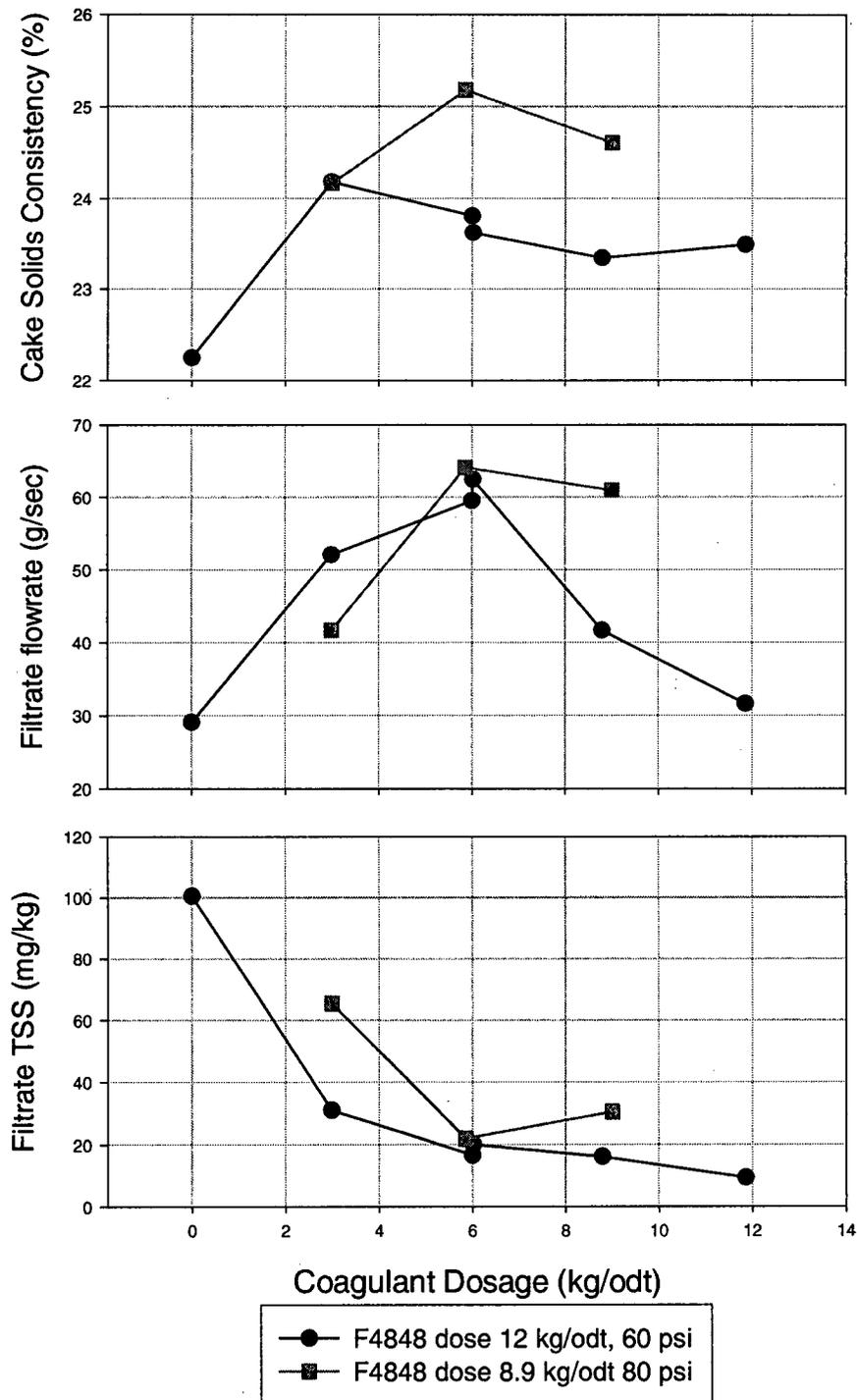


Figure 4-16. Pressure effect on optimal coagulant dosage

## **4.9 Polymer Selection**

### **4.9.1 Deinking Mill Sludge**

#### **4.9.1.1 Various Single Polymer Flocculants**

Four types of polymer at the same dose 1.3 kg/odt which was the optimal dose for Percol 734 (refer to Section 4.8.1) were tested to decide which worked best on deinking combined sludge. The polymers were Percol 721, 734, 765 from Ciba Chemical Co., and Perform pc8704 from Hercules Chemical Co. Table 4-5 outlines what is known about the Percol polymers. The results of their addition to this combined sludge are shown in Figure 4-17.

Table 4-5. Comparison of Percol polymers

Polymers	Charge density	Molecular weight
Percol 721	Lowest	Slightly lower than Percol 734
Percol 734	Higher than Percol 721	Highest
Percol 765	Highest	Lowest

The uppermost plot in Figure 4-17 is of cake solids consistency vs. the polymer used. All four polymers produced similar results although Percol 721 had slightly lower cake solids. Percol 734 gave the highest cake solids consistency but it was not significantly (statistically speaking) different from the others except Percol 721. The middle plot is of filtrate flow-rate vs. polymer used. Percol 721 exhibited the fastest filtration rate. Percol 765 was in second place, then Perform PC 8704. Percol 734 had the lowest filtration rate. The bottom plot in Figure 4-18 is of filtrate TSS. Here the results were quite different from those of cake solids. Percol 721 and 765 had lower filtrate TSS, then Perform PC 8704. Percol 734 had the highest filtrate TSS perhaps due to its lowest charge density. We observe that which polymer to select depends on which dependent variable we want to optimize.

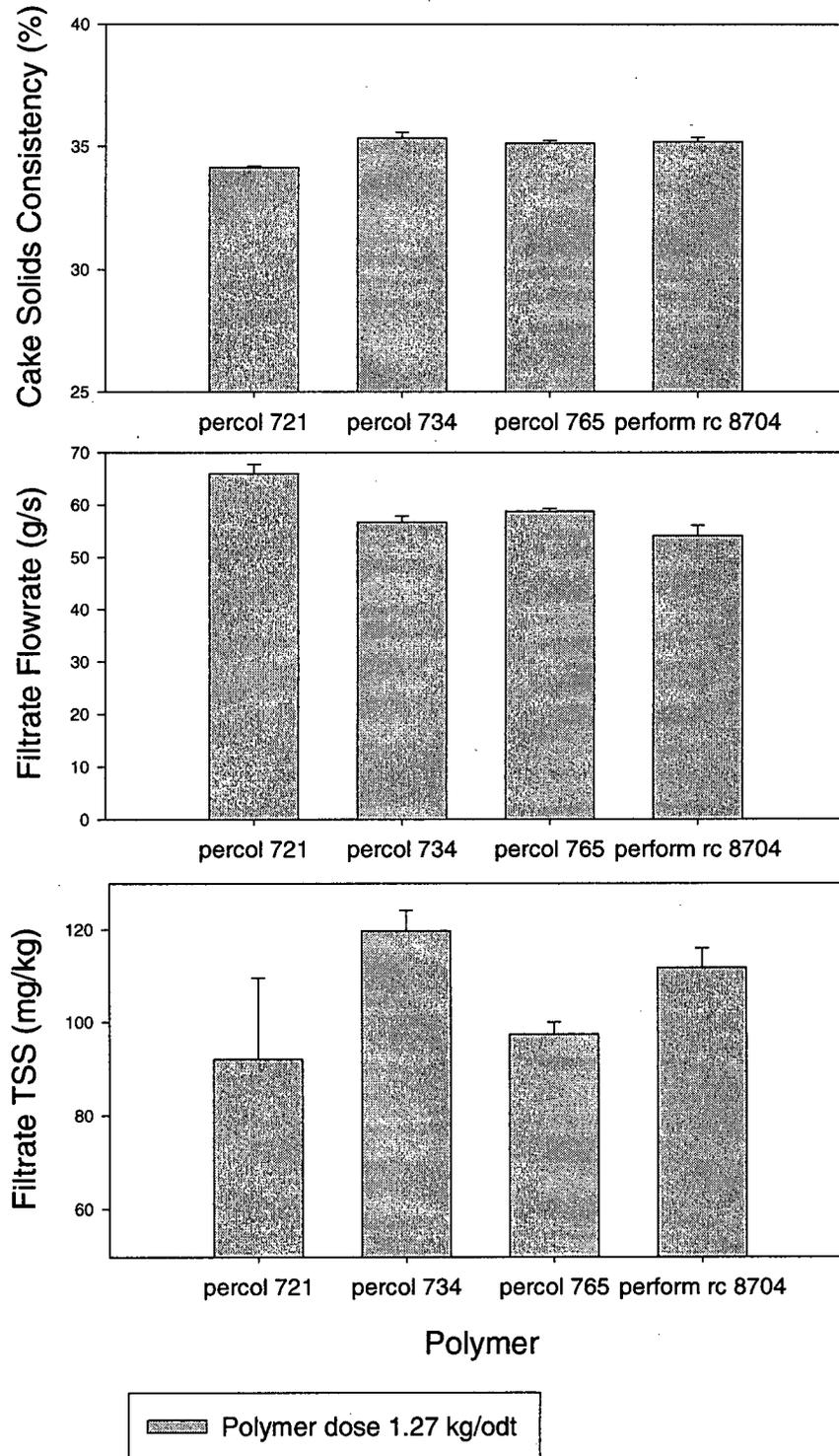


Figure 4-17. Polymer effect on combined deinking sludge dewatering

#### **4.9.1.2 Single Polymer vs. Dual Polymers**

Percol 734 is usually used as a single polymer flocculant for deinking sludge dewatering. In these experiments of deinking primary sludge dewatering it was used alone, where the coagulant dosage was 0. It was also used in a dual polymer system with coagulant polymer C4030. The results are compared in Figure 4-18 to those obtained using the same coagulant C4030 with flocculant F4848. The amounts of flocculant to be used were determined by preliminary tests, which showed the optimal dosages to be 2.36 kg/odt for Percol 734 and 7.8 kg/odt for F4848.

In terms of cake solids consistency the single polymer Percol 734 was less effective either alone or in combination with C4030 than flocculant F4848 alone or in combination with C4030. Addition of C4030 to percol 734 resulted in no significant improvement over using it alone. Too much coagulant resulted in lower cake solids consistencies. The use of coagulant C4030 with flocculant F4848 did not significantly affect the cake solids consistencies.

In terms of filtrate flowrate there was an increase as dosage of C4030 went from 0 to 2 kg/odt followed by a decrease for the dual polymer system C4030/F4848. Addition of C4030 to single polymer flocculant percol 734 resulted in a linear decrease in filtrate flowrate.

In terms of filtrate TSS there was a minimum C4030 dosage of 2 kg/odt for the C4030/F4848 system. A somewhat lower dosage (1 kg/odt) with the C4030/percol 734 system was observed to produce minimum filtrate TSS.

It seems dual polymers didn't improve cake solids consistency for deinking sludge dewatering, but they can reduce the filtrate TSS at optimal dose.

#### **4.9.2 Kraft Mill Sludge**

Western Pulp Squamish was applying the single polymer Perform pc 8704 for it's combined kraft mill sludge dewatering. In the literature (Kenny, 1997), for combined sludge dewatering, more mills were said to be using dual polymer systems (coagulant + flocculant). We compared single polymer use with dual polymer use on kraft combined sludge dewatering. The results are shown in Figure 4-19.

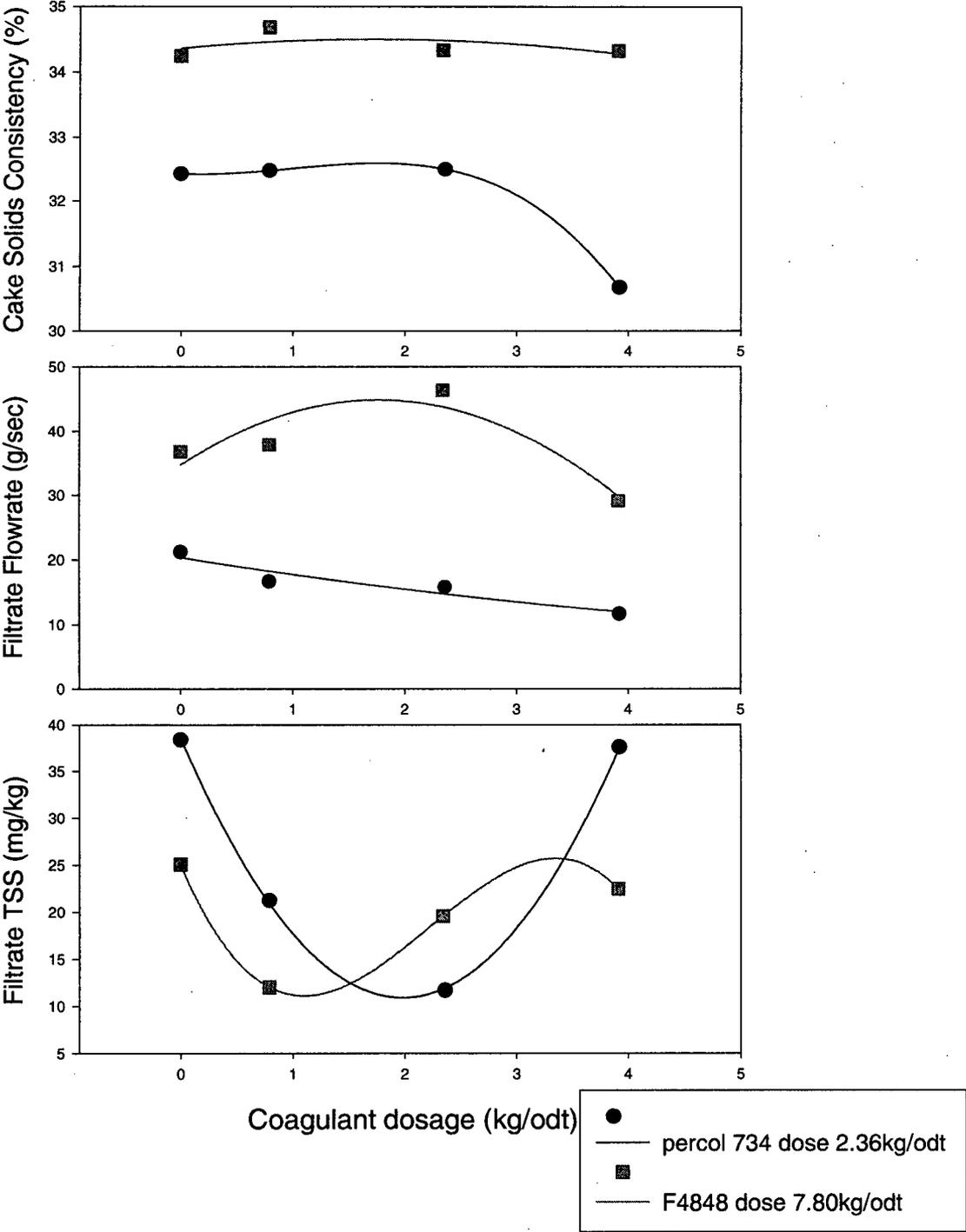
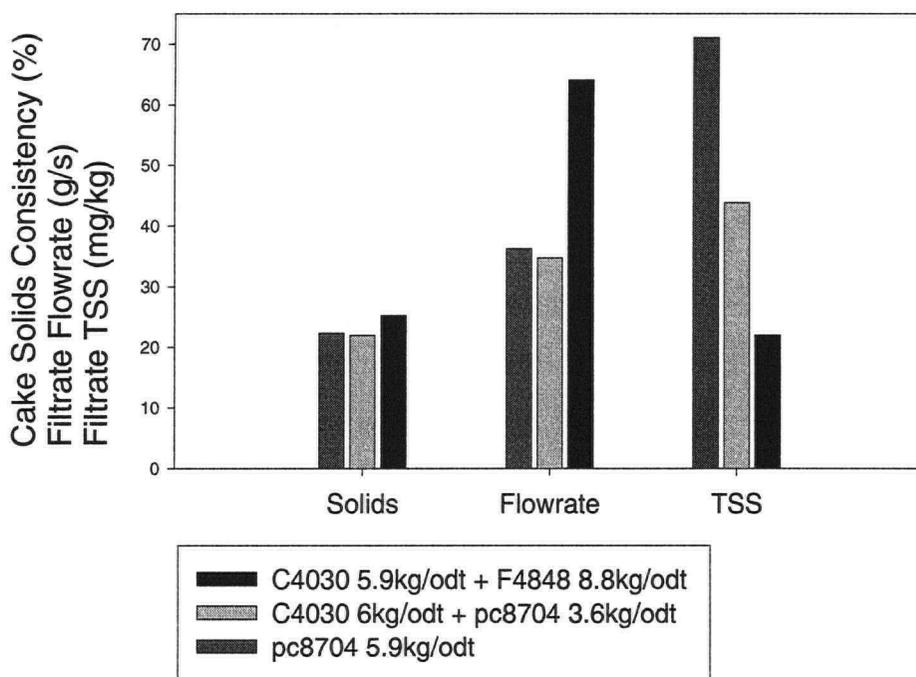


Figure 4-18. Comparison of single or dual polymer on deinking primary sludge dewatering

Comparing single polymer pc 8704 with dual polymers C4030/F4848 indicated that this dual system gave slightly higher cake solids consistencies, much higher filtrate flowrates and much lower filtrate TSS.

Using 6 kg/odt of C4030 with pc 8704 resulted in getting the same cake solids and filtrate flowrates with 3.6 kg/odt (optimal dose determined by a test) of pc 8704 that were obtained with 5.9 kg/odt (optimal dose see Figure 4-13) of pc 8704 alone. In addition the filtrate TSS were lowered.



**Figure 4-19. Kraft combined sludge dewatering: Single polymer vs. dual polymers**

Obviously the combination of C4030/F4848 is the best for kraft mill combined sludge dewatering. For polymer pc8704, the combination with C4030 didn't improve cake solids consistency and filtrate flowrate, it lowered filtrate TSS.

### **4.9.3. Inorganic Coagulant vs. Polymer Coagulant**

Organic polymers are much more expensive than inorganic chemicals. Some inorganic chemicals such as ferric, ferrous and aluminum salts and lime are traditionally used as sludge conditioning agents for dewatering sludges on vacuum filters and filter presses. The disadvantage of inorganic chemicals is that they can add 20 to 50 % more weight to the amount of cake requiring disposal. If the sludge is incinerated, they lower its heat value, do not burn and thus increase the amount of ash (WPCF, 1988)

Watanabe, Kubo and Sato (1999) found that a combination of an inorganic coagulant and an amphoteric, polymeric flocculant could enhance organic sludge dewaterability. Plaisier and Dorica (1997) found that a ferric sulphate coagulant and a low-charge, cationic flocculant showed an acceptable dewaterability at low conditioning costs. The use of combinations of metal ion coagulants and organic flocculants could result in reduced polymer costs.

Based on the above discoveries we tried two kinds of inorganic chemicals as coagulants together with the flocculant F4848. They were ferric chloride and polyaluminum chloride (PAC). The dosage effects of the two inorganic coagulants are shown in Figure 4-20. The flocculant F4848 dosage was the optimal dosage determined in the dual polymers dosage tests (see Figure 4-15). Figure 4-21 compares the dewatering results by the three coagulants, C4030, PAC and  $\text{FeCl}_3$  at their optimum dosages with similar flocculant doses. In terms of cake solids consistency, there was no significant difference. C4030 had the fastest flowrate, next was PAC;  $\text{FeCl}_3$  had the lowest one. In terms of filtrate TSS,  $\text{FeCl}_3$  was lowest, C4030 was in the middle, and PAC was highest.

Table 4-6 indicates that the coagulant  $\text{FeCl}_3$  is much cheaper than the organic coagulant C4030. But coagulant PAC is more expensive than C4030. Since when ferric chloride is combined with organic polymers, the amount of inorganic added is less than when it is used by itself, the increased amount of ash may be tolerable during incineration.

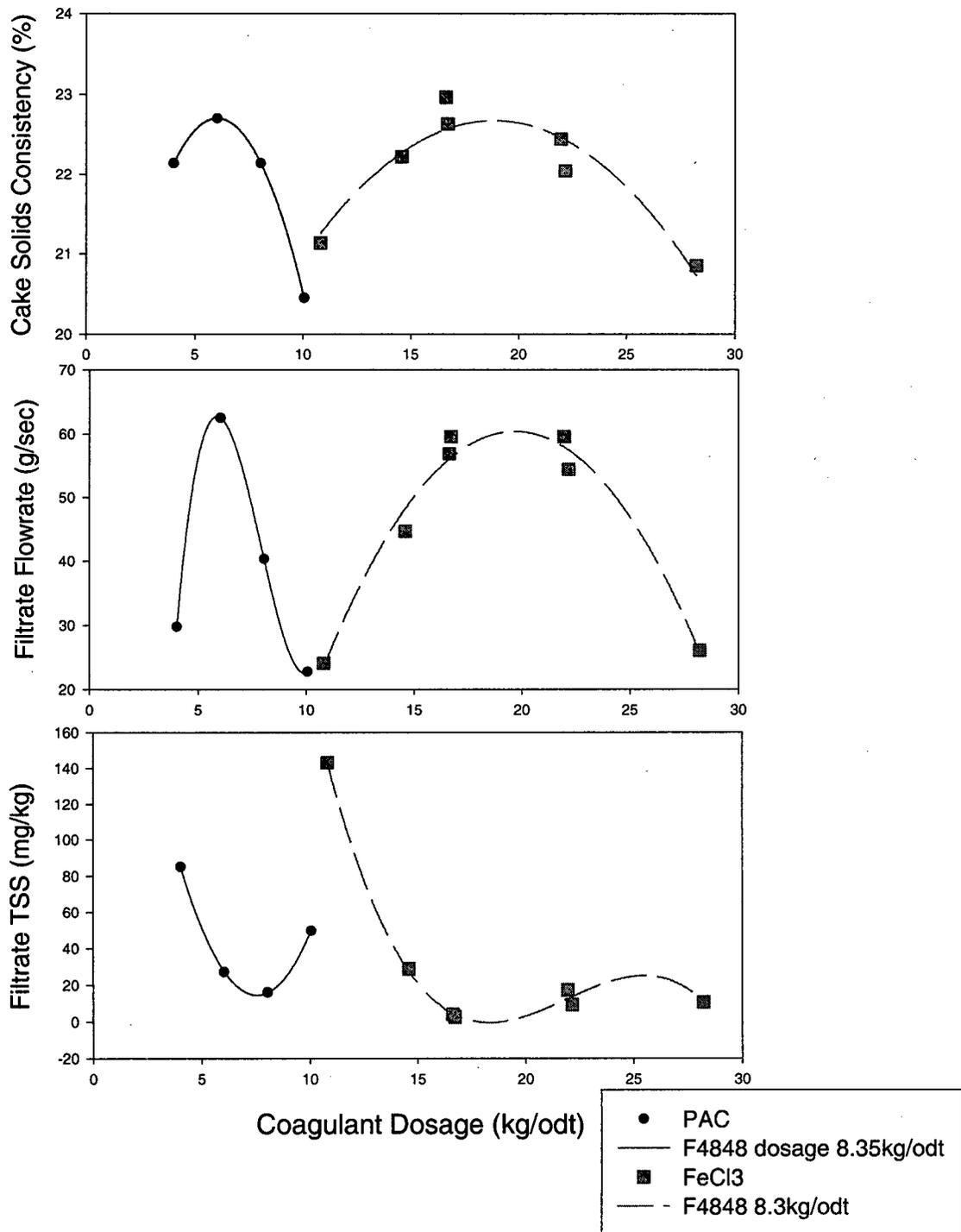


Figure 4-20. Inorganic coagulant dosage effect on kraft combined sludge dewatering

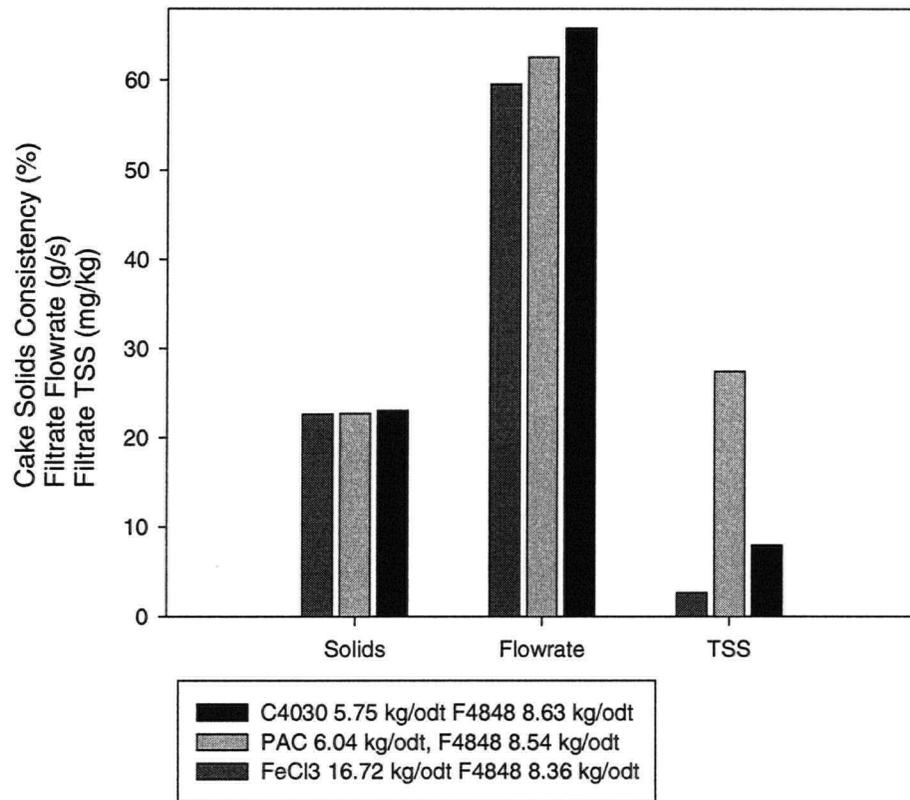


Figure 4-21. Comparison of organic and inorganic coagulants on kraft combined sludge dewatering

Table 4-6. Price comparison of various coagulants

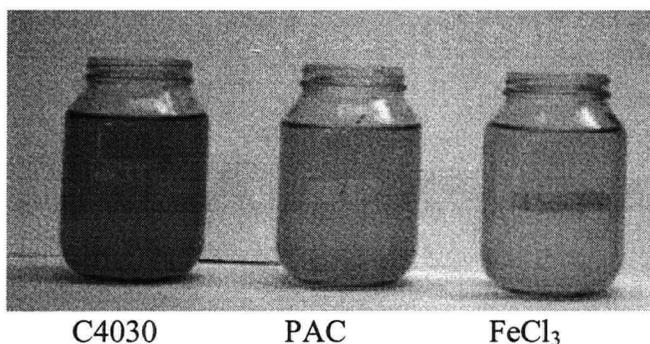
Coagulants	Optimum dose (kg/odt)	Price (CN\$/kg)	Cost (CN\$/odt)
C4030	6	3.45	20.70
FeCl <sub>3</sub>	17	0.30-0.42	5.10-7.14
PAC	6	7.06	42.35

These two inorganic coagulants have similar trends for final cake solids consistency, filtrate flowrate and filtrate TSS. But the optimal PAC dosage (around 6kg/odt) is much lower than that of FeCl<sub>3</sub>, which is between 17-22 kg/odt. The optimum dose of PAC is much more sensitive to

PAC concentration than is that of  $\text{FeCl}_3$ . Thus addition of PAC would have to be more carefully controlled.

#### **4.9.4 Comparison of Organic Coagulant and Inorganic Coagulants**

Organic coagulant C4030 was compared with the two inorganic coagulants, all at their optimal dosages, the dewatering results are shown in Figure 4-21. C4030 worked a little better on filtrate flowrate and cake solids consistency than PAC and  $\text{FeCl}_3$ . But  $\text{FeCl}_3$  gave the lowest filtrate TSS. We also found that the inorganic coagulants resulted in a clearer/lighter coloured filtrate. See Picture 4-1.



**Picture 4-1. Filtrates using various coagulants**

#### **4.10. Charge demand measurement**

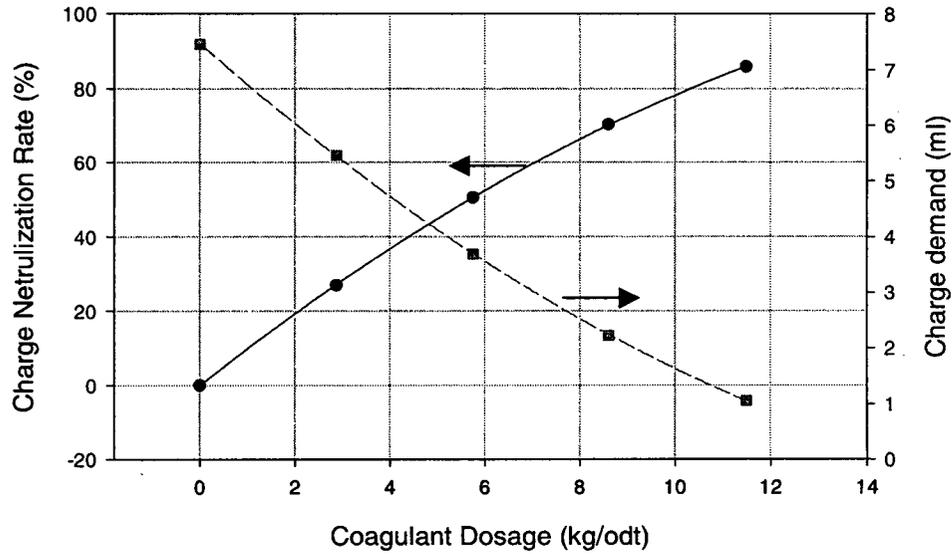
In our kraft mill sludge dewatering tests both coagulant and flocculant were cationic polymers. These were used to generate flocs from initially negatively charged sludge particles. To investigate the floc formation process the particle charge detector described in Section 3.5.4.2 was utilized.

Some results are presented in Figures 4-22 and 4-23 for kraft mill combined sludge using coagulant polymer C4030 and flocculant F4848. These results are presented in terms of charge demand which is given in ml of standard polymer solution necessary to bring the streaming potential to 0 and the charge neutralization which is the % of the particle surface charge that has been neutralized. Thus

$$\% \text{ charge neutralization} = (\text{CD}_0 - \text{CD}_1) / \text{CD}_0 * 100$$

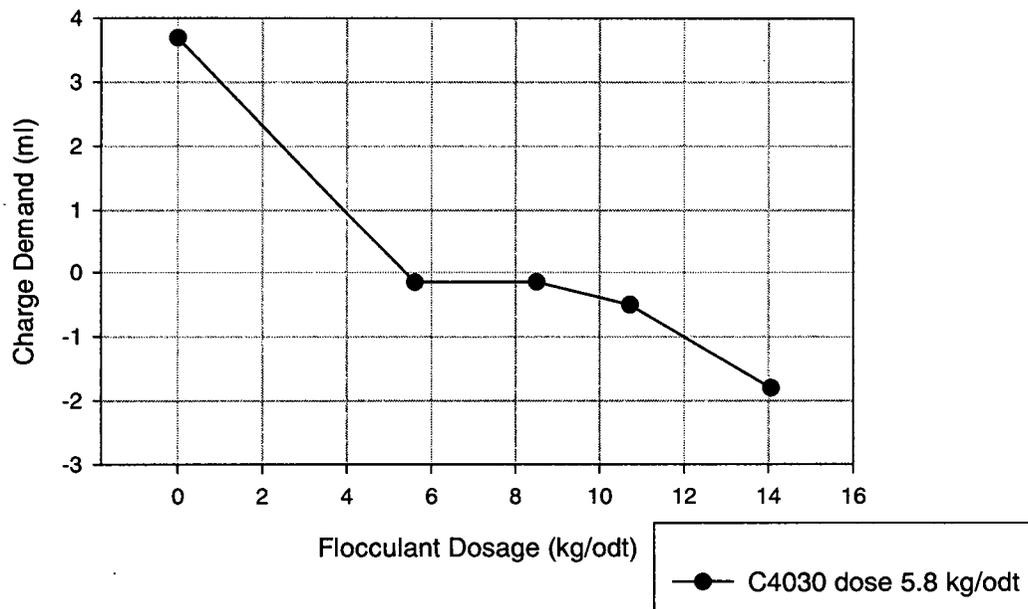
Where  $\text{CD}_0$  = charge demand of original sample

$CD_1$  = charge demand after a certain amount of neutralized polymer has been added



**Figure 4-22. Coagulant dosage effect on charge demand and charge neutralization rate**

In Figure 4-22 as the coagulant dose increased the charge demand decreased indicating that the negative charge on the sludge particle surface as measured by the streaming potential, was decreasing and the charge neutralization % increased. Since it had been shown (refer to Figure 4-14) that the optimum dose of C4030 was 6 kg/odt, Figure 4-22 shows that a charge neutralization of 50% was required to achieve this optimum.



**Figure 4-23. Flocculant effect on charge demand**

**Table 4-7. Charge demand and charge neutralization rate of coagulants**

Coagulants	Optimal dosage (kg/odt)	Charge demand (ml)	Charge neutralization (%)
C4030	5.8	3.5159	53
FeCl <sub>3</sub>	17.7	3.0047	60
PAC	6	2.4351	67

Figure 4-23 shows how charge demand changed with changes in the amount of flocculant F4848 added. Prior to the addition of flocculant 5.8 kg/odt of coagulant C4030 were added. As previously shown in Figure 4-16, the optimal dose of flocculant in terms of cake solids, filtrate flowrate and filtrate TSS was around 6 kg/odt. At this dosage, following coagulant addition at 5.8 kg/odt, Figure 4-23 shows that the charge demand was close to zero and the plot of charge neutralization indicates that at the optimal dose essentially all of the charge had been neutralized. Comparing Figures 4-16 and 4-23 tells us that the measurement of charge demand is more sensitive to flocculant dose than the measurements of cake solids, filtrate flowrate and filtrate

TSS. Note that under optimal conditions not all of the charge was neutralized by the coagulant; some remained to be neutralized by the flocculant.

We also measured charge neutralization for inorganic coagulants at their optimal dosages (see table 4-7). For PAC it was 67%, for  $\text{FeCl}_3$  60%. These confirmed that the coagulant didn't neutralize the all surface charges of particle, the positive flocculant also contributed to charge neutralization, but it seems there was a optimal charge neutralization for a certain coagulant, at which the flocculant worked best.

#### **4.11 Sludge dewatering profile**

Our laboratory sludge press aims at simulating the performance of filter presses and screw presses. Although before the filter press and screw press there is normally some thickening equipment such as a rotational screen thickener or a gravity thickener, the processes that occur in both filter presses and screw presses are filtration followed compression (expression). The laboratory sludge press also has these two stages, in which the filtration stage may be longer in duration since we didn't thicken the sludge first.

The dewatering profile given by the laboratory sludge press is the first plot in Figure 4-24. The figure also shows a plot of  $t/w$  vs.  $w$ , the slope of the linear part of which can be employed to calculate the parameter SRF. Figure 4-24 is just an example, which is deinking primary sludge conditioned by Percol 734.

The two stages of filtration and compression can be clearly distinguished in the first plot and the second plot. The main part of the filtration is linear in the second plot, which illustrates that the filtration part complies with filtration theory (refer to chapter 2). There is also a short section which connect the filtration and compression. The compression stage is that part of the profile which has a sleeper slope at the end of the test.

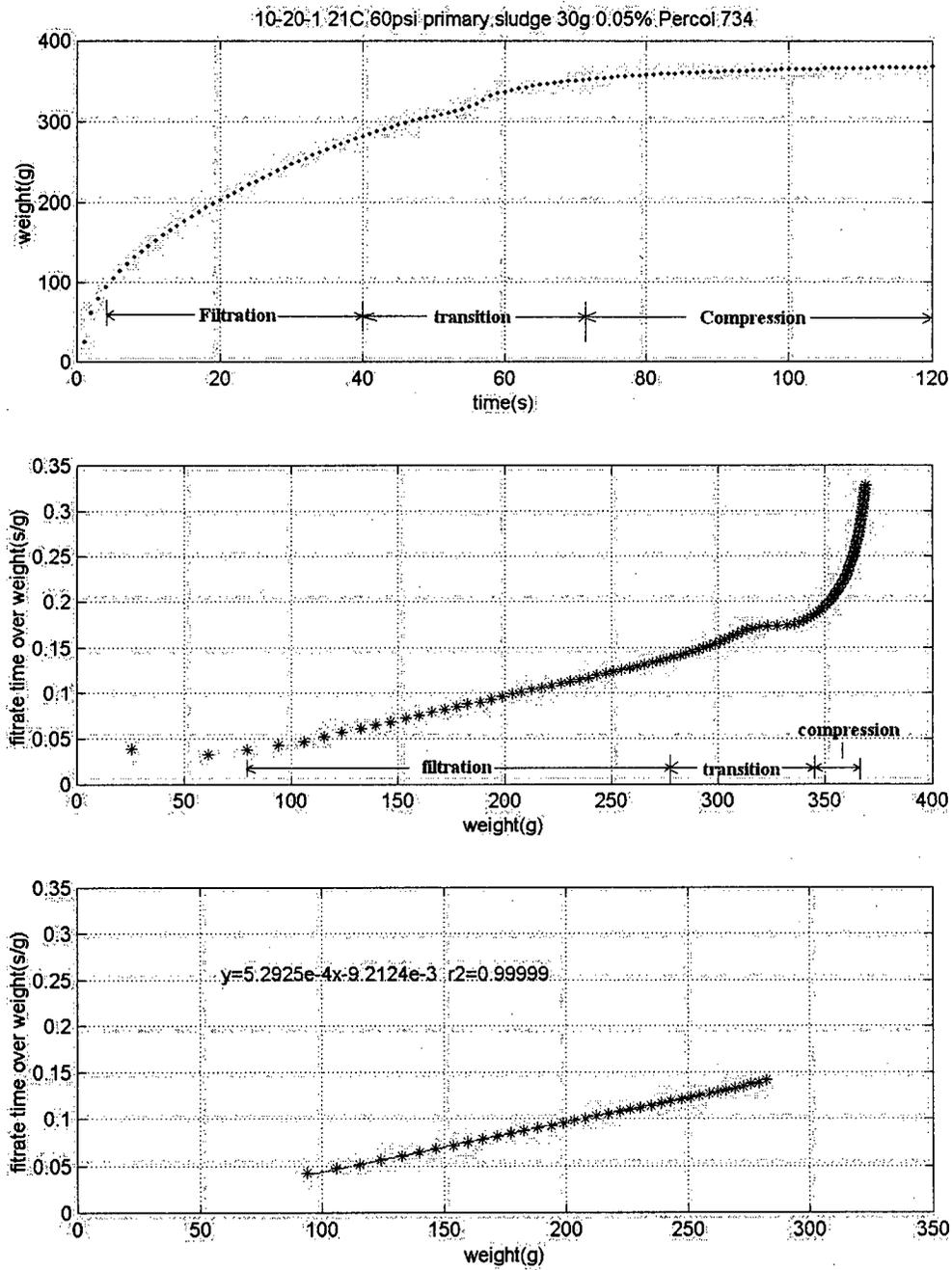


Figure 4-24. Deinking sludge dewatering profile and plot of t/w vs. w

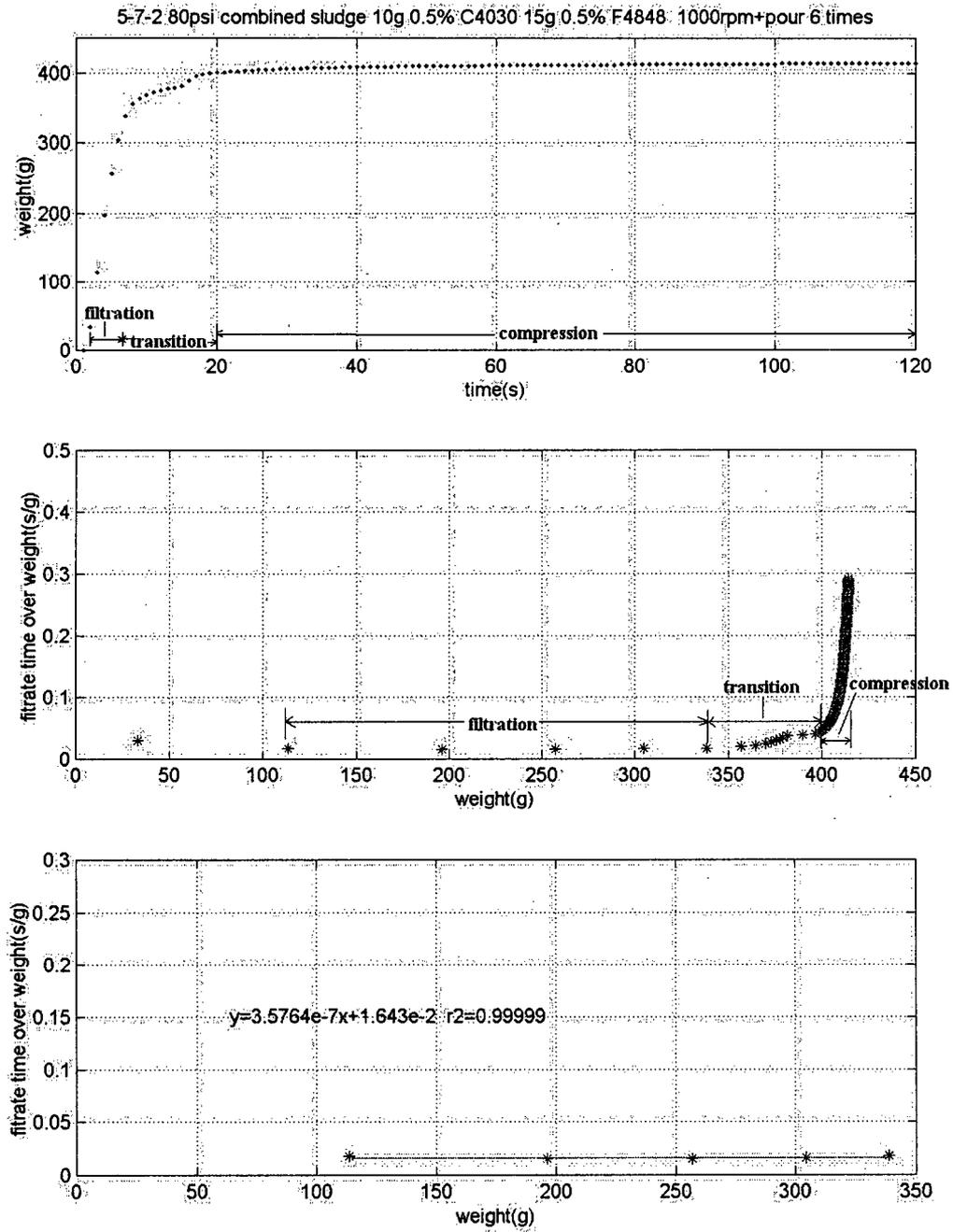


Figure 4-25. Kraft combined sludge dewatering profile and plot of  $t/w$  vs.  $w$

The third plot simply isolates part of the filtration stage and least squares fits it to a straight line, the equation of which is given.

Figure 4-25 displays a kraft sludge dewatering profile conditioned by a dual polymer system. From this figure, we can see it is hard to calculate the SRF since the filtration process was so fast that the slope of second plot was nearly zero. So for a well-conditioned sludge, it was difficult to calculate an SRF value. As a result of observations like these we decided not to pursue measurement of SRF as a criterion in our sludge dewatering studies.

#### 4.12 Compressibility of combined deinking sludge

According to equation  $\alpha = \alpha_0 \Delta P^n$  (2-11), a plot of  $\ln \Delta P$  vs.  $\ln \alpha$  ( $\ln$  SRF) should result in a straight line, from the slope of which the compressibility parameter  $n$  can be estimated. Such a plot is shown in Figure 4-26. The data used in preparing Figure 4-26 are presented in Table 4-8. The SRF values were calculated using sludge dewatering profile like those shown in the lowest plot of Figure 4-24. From Figure 4-26,  $n = 0.94$ . According to Leu (1981), the degree of compressibility is between moderate and high ( $n = 0$  indicates incompressibility).

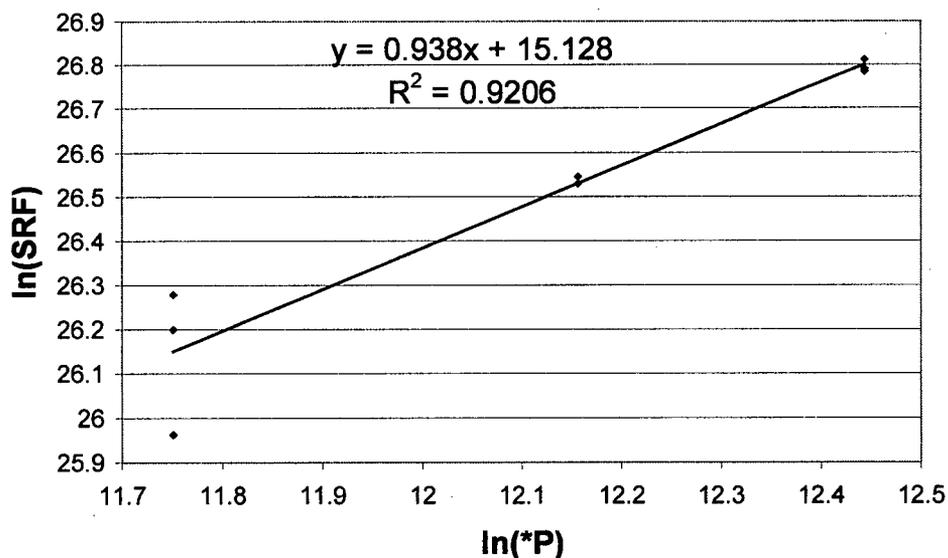


Figure 4-26. Plot for assessing compressibility

Table 4-8 SRF results for combined deinking sludge at various pressures

Samples	Air Pressure (psi)	Pressure drop (Pa)	Slope (S/g <sup>2</sup> )	SRF (m/kg)
1	40	126868	9.46E-04	2.59E+11
2	40	126868	6.89E-04	1.89E+11
3	40	126868	8.74E-04	2.39E+11
4	60	190302	8.24E-04	3.38E+11
5	60	190302	8.23E-04	3.38E+11
6	60	190302	8.11E-04	3.33E+11
7	80	253736	7.84E-04	4.29E+11
8	80	253736	8.06E-04	4.41E+11
9	80	253736	7.88E-04	4.31E+11

#### 4.13 Analysis of dewatering parameters

Frequently assessments of dewaterability are based solely on the filtration stage of overall process, for example measurement of by CST, SRF, TTF, Filtrate Flowrate, et al. Can these parameters effectively characterize the dewatering process? In other words is filterability synonymous with dewaterability? Consider the results shown in Figure 4-27. The filtrate flowrates are almost the same; these of course are characteristic of the filtration mechanism. However there is a significant difference in the cake solids contents. This difference is not accounted for by measuring filtrate flowrate in fact the slightly higher filtrate flowrate was observed with pc 8704 while the higher cake solids were noted with C4030 + F4848.

Consider also Figure 4-11. There it can be seen that as filtrate flowrate increases the cake solids go to a maximum and then decrease. Thus observations on filtrate flowrate cannot account for cake solids consistency effects. Cake solids consistency is probably the most important characteristic of dewatered sludge in terms of cost of its disposal.

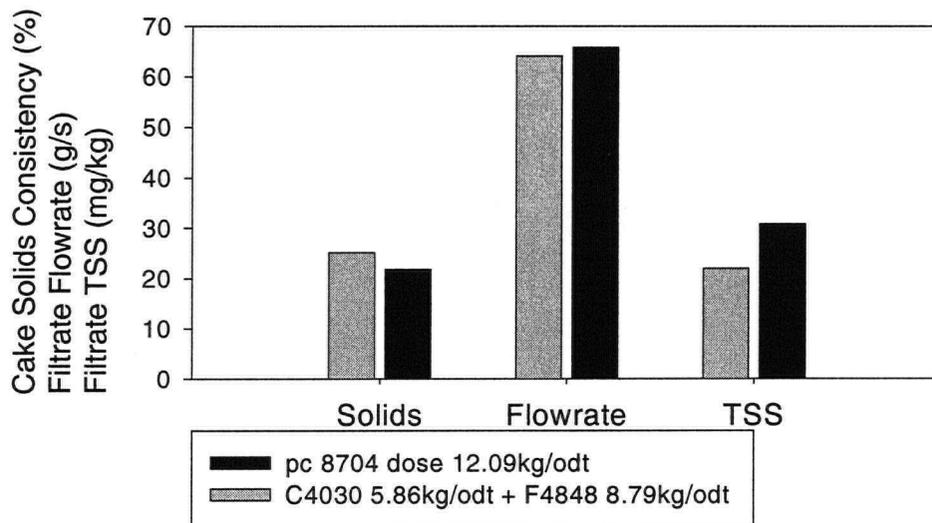


Figure 4-27. Comparison of three parameters of two runs

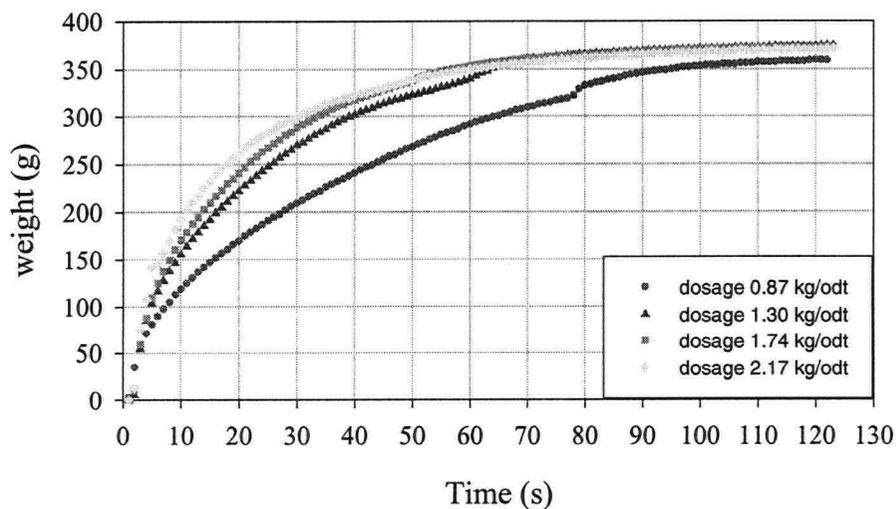


Figure 4-28. Dosage effect on deinking primary sludge dewatering profile

Figure 4-28 gives more insight. In it the weight of filtrate from the sludge press is plotted vs. time for primary deinking sludge with various doses of Percol 734. Note that as the polymer dose increases the filtrate flowrate, as indicated by the initial slope of the plot, increases but the final cake solids consistencies didn't increase, it even decreased at dosage 2.2 kg/odt (see Figure 4-11 dosage effect on primary deinking sludge).

Plots of filtrate flowrate and cake solids consistency vs. SRF using the data from primary deinking sludge dosage effect (Figure 4-11) are shown in Figure 4-29. The first plot shows there is strong linear relationship between SRF and filtrate flowrate as we expected. But the second plot doesn't show any relationship between SRF and cake solids consistency, which means that SRF didn't account for final cake solids consistency.

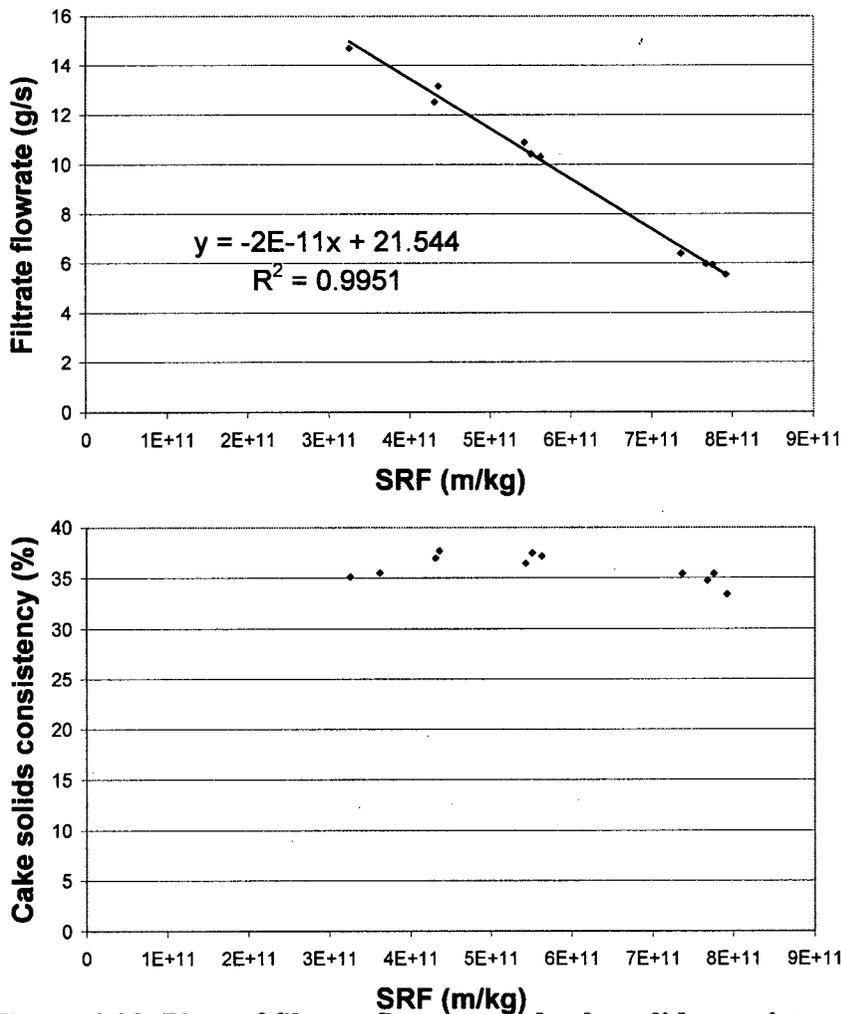


Figure 4-29. Plots of filtrate flowrate and cake solids consistency vs. SRF

## **Chapter 5**

### **CONCLUSIONS AND RECOMMENDATIONS**

#### **5.1 Conclusions**

The laboratory sludge press can be used to simulate the behaviour of sludge dewatering industrial equipment. It was shown that the sludge dewatering process consists of two stages, filtration for solids collection followed by compression of the collected cake to squeeze out water. Parameters such as filtrate flowrate (or time to filtration) and specific resistance to filtration (SRF) alone cannot completely measure sludge dewaterability with respect to final cake solids consistency. Under some conditions the laboratory sludge press can be used to determine SRF.

The laboratory sludge press can be used to study sludge/polymer mixing by measuring cake solids consistency, filtrate flowrate and filtrate total suspended solids as functions of mixing conditions. For deinking sludge the best sludge/polymer mixing conditions were 1000 rpm for 1-3 seconds using a 5-cm diameter marine type impeller in a 1000 ml beaker containing 435 g of sludge sample. The best mixing conditions observed with kraft mill sludge were 1000 rpm for 1-3 seconds followed by pouring the sludge polymer suspension back and forth between two beakers several times.

Using deinking sludge it was shown that increasing the pressure applied to the sludge resulted in an increase in sludge cake solids consistency, a small increase in filtrate TSS and no change in filtrate flowrate.

Increasing the time under pressure in the lab sludge press caused a rapid increase in sludge cake solids consistency during the first ten minutes followed by a more gradual increase. The usual

press time of 2 minutes was adequate to characterize the ranking order when comparing various polymer additions to the same sludge.

For deinking sludge an increase in temperature produced a linear increase in cake solids consistency and filtrate flowrate. Some of this increase could be attributed to decreased filtrate viscosity at higher temperatures but other factors also play a role.

In deinking sludge dewatering an increase in primary/secondary sludge ratio produced a linear increase in cake solids consistency and a reduction in filtrate TSS which tapered off at higher primary sludge contents.

Secondary sludges from both deinking and kraft mills could not be effectively dewatered alone without addition of polymers or filter aids. Sawdust and hogfuel as filter aids were effective in improving the dewatering of deinking mill secondary sludge but the amounts required were excessive and the amount of conditioning polymer need was not reduced by their presence.

The laboratory sludge press can be used to rapidly determine the optimal polymer dose for the dewatering of a particular sludge. The optimum values for cake solids consistency, filtrate flowrate and filtrate TSS are not always consistent with one another.

In optimizing polymer additions one must determine which of the dependent variables (cake solids consistency, filtrate flowrate, filtrate TSS) is the most important.

In dewatering combined primary and secondary kraft mill sludge as the primary/secondary ratio decreased the optimal doses of coagulant and flocculant polymer increased.

The laboratory sludge press can be used to quickly screen a variety of polymers to see which one is most effective in dewatering.

For the deinking sludge and the polymers used in this research dual polymer conditioning did not improve dewatering compared to single polymer conditioning in term of cake solids consistency, but it lowered the filtrate TSS.

Inorganic coagulants such as  $\text{FeCl}_3$  and PAC combined with an organic polymer flocculant were also able to produce good dewatering results with kraft mill combined sludge.

Measurement of sludge particle charge may be more sensitive for determining optimal sludge conditioning chemical addition than the laboratory sludge press. Both cationic coagulant and flocculants were shown to contribute to charge neutralization.

## **5.2 Recommendation**

The laboratory sludge press can produce reliable results in most cases. But it was observed that when kraft mill sludge with very high secondary sludge was pressed, the cake was too sticky to be pulled up by the piston, thus it remained on the screen and some water was sucked back into the cake, resulting in an error in cake solids consistency. This could be overcome by enlarging the filtrate-collecting base so that the water was not in contact with the bottom of the cake after completion of the press cycle.

A systematic study, using the methods described in this thesis, should be done using a variety of filter media (screen) of known void fraction and weaving pattern.

Sludge-polymer mixing is important for sludge dewatering. It seems there are specific mixing criteria for each kind of sludge. Further research is needed in this area.

Streaming current measurement seems effective in polymer dosage control, more tests and research on it are needed.

Further analysis of the mechanism of filtration followed by compression as it occurs in this lab press should be done. This would involve a mathematical description of the sludge dewatering process.

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## APPENDIX

### A-1. Calculation

1. Cake solids consistency:

$$= \text{Weight of wet pressed pad} / \text{weight of oven dried pad} * 100\%$$

2. Filtrate flowrate

$$= 250\text{g filtrate} / \text{time of 250g filtrate (s)}$$

3. Solids mass balance

Kraft mill sludge

Jul. 28/00. Combined sludge consistency: 2.1%

$$\text{Test 1: In: } 435.81 * 2.1\% = 9.15 \text{ g}$$

$$\text{Out: dry cake + filtrate TSS} = 8.92 + (0.9010 - 0.8860) = 8.935 \text{ g}$$

$$\text{Error: } (9.15 - 8.935) / 9.15 = 2.35\%$$

$$\text{Test 2: In: } 435.33 * 2.1\% = 9.14 \text{ g}$$

$$\text{Out: } 9.05 + (0.8707 - 0.8624) = 9.06 \text{ g}$$

$$\text{Error: } (9.14 - 9.06) / 9.06 * 100\% = 0.88\%$$

Deinking sludge

Dec. 6/99. Primary TSS: 2.6%

$$\text{Test 1: In: } 434.34 * 2.6\% = 11.32 \text{ g}$$

$$\text{Out: } (14.31 - 3.07) + (0.7721 - 0.732) = 11.28 \text{ g}$$

$$\text{Error: } (11.32 - 11.28) / 11.32 * 100\% = 0.35\%$$

$$\text{Test 2: In: } 435.14 * 2.6\% = 11.31 \text{ g}$$

$$\text{Out: } (14.36 - 3.08) + (0.8047 - 0.7558) = 11.33 \text{ g}$$

$$\text{Error: } (11.33 - 11.31) / 11.31 * 100\% = 0.18\%$$

Errors (< 3%) were produced by losses in transferring the conditioned sludge from a beaker into the sludge press cylinder. Another reason is that it is difficult to keep the sludge sample consistency constant during sampling. It may vary a little from one sample to another. It is more

accurate to decide the mass of a sample by dry weight (oven-dry cake solids plus the TSS in the filtrate) if the mass loss during transfer could be eliminated.

4. Polymer dosage

= polymer weight/(oven-dry cake + TSS in the filtrate (kg/odt))

5. SRF

According to equation (3-2),  $\frac{t}{W} = \frac{\mu c \alpha}{2\rho^2 A^2 \Delta P} W + \frac{\mu R_m}{\rho A \Delta P}$ , the slope of plot of t/w vs. w is:

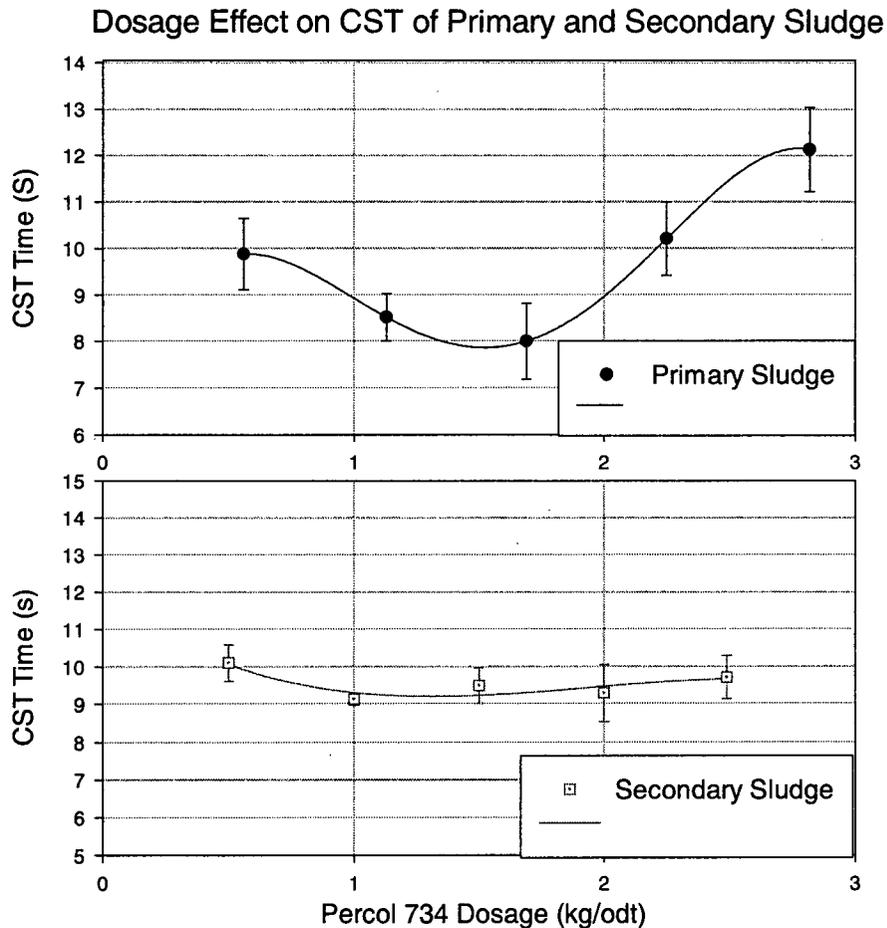
$$\frac{\mu c \alpha}{2\rho^2 A^2 \Delta P} \cdot SRF(\alpha) = \frac{slope \times 2\rho^2 A^2 \Delta P}{\mu c}, \text{ c is the mass of dry cake deposited per unit}$$

volume of filtrate.

$$c = \frac{s\rho}{1-sm}, \text{ Where s = mass fraction of sample, m = wet cake weight/dry cake weight}$$

### A-2. CST test on deinking sludge dewatering

Originally it was intended to compare data from the sludge press with data from capillary suction timer. However the capillary suction timer failed near the beginning of those tests and could not be replaced.



When this figure is compared to Figure 4-11, the polymer dosages for the lowest CST of primary and secondary sludge were similar to those for the highest cake solids consistencies of the same sludges. The CST values for secondary sludge were so close that it was difficult to determine the optimal dosage.

### A-3. Flow sheet of kraft mill sludge dewatering

