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The Effect of Flocculants and Surfactants on the Filtration Dewatering of Iron Ore Fines

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ABSTRACT

The filtration and separation characteristics of iron ore fines have been investigated. The experimental work included: characterization, evaluation of suitable flocculants to enhance settling and dewatering rate, determination of optimum dosages of flocculants, and influence of surfactants on filtration dewatering of both flocculated and unflocculated iron ore fines by vacuum filtration. The results showed that 1) the settling rate can be enhanced many-fold (from 2.52 to 90 m/h) by a suitable flocculant, 2) the residual filter cake moisture content can be reduced from 18.2% without reagents to 12.6% with suitable surfactant dewatering aids while the filtration rate can be enhanced from 4.8 to 97.2 L/h with suitable dewatering filter aids, and 3) the specific cake resistance to filtration can be brought down from 8.6×10^9 to 1.0×10^9 m/kg by using a surfactant sodium petroleum sulfonate at a concentration of 1.47×10^{-3} kg/t. The economics of using flocculant filter aids and surfactant dewatering aids before thermal drying is described and discussed.

Key Words. Dewatering; Surfactant; Flocculant; Filtration; Separation

INTRODUCTION

Virtually all process industries in which fine particulate slurries are handled employ some form of filtration as a unit operation. In general, the aim is either the recovery of extensively dewatered solids prior to thermal drying or the disposal, shipment, or maximum recovery of filtrate to allow further processing of valuable solubles. In the minerals industry in particular, dewatering is employed in order to save on the energy required in thermal drying, to improve the handling properties of the concentrates, and to reduce transportation costs if concentrates are to be shipped. Although some particulate slurries filter perfectly well without chemical conditioning, the great majority of pulps of a widely varying nature can benefit from pretreatment. Flocculants and surfactants are commonly used for enhancing filtration separation and reducing filter cake moisture content.

The use of synthetic polymers for solid–liquid separation is becoming more and more common, and therefore this area has gained considerable attention. The basic function of a flocculant filter aid is to totally adsorb at the solid–liquid interface and aggregate particulate matter. Flocculants increase the pickup or filter yield of solids and produce a permeable filter cake amenable to rapid dewatering. Basically, long-chain polymers bridge between individual particles to give a multiparticle aggregate. Aggregation of fines by filter aids prevents filter cloth blinding which otherwise can seriously retard the kinetics of cake formation (1). The nature of flocs are of paramount importance in filtration separation. The characteristics of the flocs formed by the use of a polymer depends on several factors such as surface chemistry, size and size distribution, shape, density etc. of the solid component, viscosity, dielectric constant etc. of the suspension, chemical nature of the backbone and side chains, molecular weight, molecular weight distribution, and charge and charge density of the flocculant used. The adsorption of high molecular weight polymers on particle surfaces may be due to electrostatic, hydrogen, van der Waals, and covalent bonding (2). Polymers with a high charge density can adsorb onto any surface of an opposite charge, irrespective of their chemical nature, due to electrostatic forces (3). Polymers also adsorb on surfaces of like charge when the repulsion due to electrostatic forces is not strong enough to prevent polymer adsorption by the following mechanisms of interaction (4): 1) van der Waals interaction, 2) hydrogen bonding, 3) local electrostatic attraction, and 4) development of linkage between the similarly charged polymer and particle surface by a divalent or trivalent ion with the opposite sign. Some of the mechanisms considered to be responsible for floc formation are: 1) charge neutralization, 2) bridging, 3) depletion flocculation, and 4) polymer complex formation or network flocculation. The bridging mechanism is related to the adsorption of polymers with very high molecular weights which form links between the particles, where electrostatic charge is overcome (5). Flocs formed by bridging mechanism are usually stronger than those formed by charge neutralization or the electrostatic patch mechanism.

The addition of surfactant to either the feed slurry or the washwater of a filtration unit has been shown to assist in product dewatering. Significant reductions in filter cake moisture content following the addition of various surfactants have been reported in the dewatering of iron ore (6), alumina trihydrate (7), and coal (8). Moisture contents reached by vacuum filtration are usually not sufficient due to the limitation of pressure difference by the ambient pressure. As an alternative to expensive thermal drying, the applied pressure difference may be raised, the capillary pressure of the porous filter cake may be lowered by surface-active agents, or both measures may be taken. Due to a worldwide tendency toward fine-grained products, removal of moisture from fines becomes more difficult because of the increase in surface area which in turn retains larger amounts of moisture. The mechanism by which a surfactant brings about improved dewatering is complex and involves a reduction of the filtrate surface tension and an increase in the solid-to-liquid contact angle (1, 8, 9). The lack of a fundamental understanding of the effect of surfactants is evidenced by the vast array of reagents which are reported to have been used for improving dewatering in filtration. Both classes of chemical additive—the flocculant as a filter aid and the surfactant as a dewatering aid—can be used separately or in combination depending on the particular requirements of the filtration application. The present study investigated the synergistic effects of optimum dosages of flocculant followed by varying dosages of dewatering aids at different experimental conditions.

The aim of the present work is to examine the usefulness of such reagents with particular regard to the interactive effective of such additions on the overall dewatering process, i.e., their mutual influence on the filtration characteristics and the residual moisture content of the resultant filter cakes. The overall objective of this study was to obtain baseline laboratory data for eventual use in the evaluation of a continuous pilot-plant scale vacuum filter for fine iron ore dewatering and separation of bulk solid from liquid phase at a preparation plant site.

Basic Theory

To clarify the method of approach, it is necessary to discuss the generally accepted basic theory and present model for filtration dewatering on the basis of calculations that have been made (10). Filtration as a phenomenon of fluid flow through a porous medium is considered as in the original Poiseuille capillary tube model postulate.

$$\frac{dv}{dt} = \frac{\pi r^4}{8\eta}\frac{dp}{dL} \tag{1}$$

where dv/dt is the fluid flow rate, r is the radius of the capillary tube, η is the viscosity of the fluid, and dp/dL is the pressure gradient across a tube length of L.

This simple model is unable to explain the fluid flow phenomenon for the porous system of a filter cake. A later model put forward by Darcy (1856) considered a porous bed to be a network of capillary systems and was postulated as follows:

$$\frac{dv}{dt} = \frac{KAP}{\eta L} \tag{2}$$

where A is the cross-sectional area of the porous medium, K is a constant which depends on the properties of the solid and the fluid (commonly known as the "permeability constant"), P is the pressure drop across the porous medium of thickness L, and η is the viscosity of the fluid flowing through the medium. The Darcy model was further modified by Tiller (1974) to simplify the determination of K:

$$\frac{dv}{dt} = \frac{P}{\eta(R_c W + R_m)} \tag{3}$$

where P is the applied pressure, R_c and R_m are the flow resistance of the filter cake and the filter medium, respectively, W is the mass of dry solid deposited per unit area of the filter, and η is the viscosity of the filtrate. Assuming that at a given time in a vacuum or pressure filtration system the pressure drop across the filter cake (P_t), R_c, and R_m are constant, then

$$P_t = \frac{\eta R_c C V^2}{2} + \eta R_m V \tag{4}$$

where $W = CV$ when C = solid content of the filtered slurry and V = volume of filtrate per unit filter area. Further, assuming the resistance due to filter medium R_m is considerably smaller than resistance due to filter cake R_c,

$$P_t = \frac{\eta R_c C V^2}{2} \tag{5}$$

Equation (5) can be rewritten as

$$P_{\text{atm}} = \frac{\eta R_c W^2}{2C}$$

or

$$W = \frac{(2CP_{\text{atm}})^{0.5}}{\eta R_c} \tag{6}$$

where P_{atm} = vacuum + average hydrostatic head.

Equation (6) can be further simplified to

$$\frac{W}{t} = \frac{(2CP_{\text{atm}})^{0.5}}{\eta R_c} t^{-1} \tag{7}$$

or

$$W/t = Yt^{-1} \tag{8}$$

or

$$\log(w/t) = -\log t + \log Y \tag{9}$$

where

$$Y = \frac{(2CP_{\text{atm}})^{0.5}}{\eta R_c}$$

which can be assumed to be constant for a steady-state filtration system. Equation (9) can describe a filtration system in which the rate of cake formation diminishes with time. In terms of the moisture content of a filter cake, a general relationship for the residual saturation (Sr) of the cake is

$$\text{Sr} = \frac{1}{H} \frac{(KAP)^{\eta}}{(Lg\gamma \cos \phi)} \tag{10}$$

where H is a constant (about 40 for a vacuum filter producing a cake thickness up to about 5 cm), η is about 0.26, γ is the surface tension, and ϕ is the contact angle.

EXPERIMENTAL

Materials

Iron Ore Fines

The sample of iron ore fine used in this work was supplied by the washing plant of the Noamundi area, Bihar, India. The particle size and chemical analysis of the supplied sample are given in Table 1. As shown, the passing size for 60% of the particles was about 45 μm and D_{50} (i.e., 50% passing) was 30 μm. The specific surface area as determined by a Malvern Particle Size Analyser (UK) was 1.4 m^2/cm^3.

TABLE 1
Sieve and Chemical Analysis of Solids in the Slurry

Size (μm)	Weight (%)	Fe (%)	SiO_2 (%)	Al_2O_3 (%)
+150	12.16	55.0	3.5	6.9
−150 + 104	15.02	56.2	3.2	6.5
−104 + 75	7.72	57.0	3.4	6.8
−75 + 45	5.00	62.1	2.9	5.7
−45 + 30	22.32	62.9	2.9	5.6
−30	37.79	58.9	4.6	9.1
Head	100.00	58.9	3.6	7.3

Flocculant Filter Aids

Flocculants filter aids were commercially available materials kindly supplied by Allied Colloids, UK (Magnafloc 1011, Magnafloc 592, Magnafloc 351) and Rishfloc, Bombay (Rishfloc 258, Rishfloc 440MV, Rishfloc 1226). The characteristics of the flocculants used are given in Table 2. All other reagents were water soluble and were prepared by dissolving 0.1 g of the reagent in 100 mL of double-distilled water.

Surfactants Dewatering Aids

Surfactant additives were supplied by Merck, Germany and Lab Chemical, India. Two anionic [sodium petroleum sulfonate (SPS) and sodium lauryl sulfate (SLS)], one cationic [cetyl trimethyl ammonium bromide (CTAB)], and two nonionic chemical additives were used. Selected prop-

TABLE 2
Characteristics of Flocculants Used

Flocculant name	Trademark	Type	Nature	Molecular weight (Million)	Solubility
Magnafloc	592	Powder	Cationic	2.5–5.0	Water
Magnafloc	1011	Powder	Anionic	5.0	Water
Magnafloc	351	Powder	Nonionic	2.5–5.0	Water
Rishfloc	440MV	Powder	Strongly anionic	—	Water
Rishfloc	258	Powder	Weakly anionic	—	Water
Rishfloc	1226	Powder	Nonionic	—	Water

erties of these aids are shown in Table 3. They were used as received. All other reagents were AR grade from B.D.H.

Methods

The settling rate of the sample was determined by a standard jar test (11). The iron ore fine slurry and water were placed in a 500 mL graduated cylinder, the flocculant was added, and the cylinder was stoppered, inverted five times for proper mixing, and then allowed to settle. The time was recorded when the interface between the suspension and the supernatant reached a predetermined mark, and after that the average settling rate was calculated graphically. The optimum flocculant dosage was chosen to be the lowest dose at which the filtrate was clear and colorless.

Filtration and dewatering experiments were carried out in laboratory bench-scale vacuum filtration equipment. A schematic diagram of the filtration arrangement is shown as Fig. 1. A known quantity of slurry was mixed with the reagents solution and then conditioned using a constant speed stirrer at 100 rpm for 5 minutes (slurry pH 5.2). The conditioned slurry was then poured into a Buchner funnel and subjected to vacuum filtration using Whatman No. 541 filter paper at a constant pressure difference of approximately 92 kPa. The filtrate flow rate was calculated by monitoring the filtrate volume collected at intervals of 5 seconds. The drying period was kept at 30 seconds after each experiment. Finally, the cake was carefully removed from the Buchner funnel, weighed, and dried in an electric oven at 105°C overnight. The dried filter cake was allowed to cool and was then weighed. The final moisture content was calculated

TABLE 3
Selected Properties of the Surfactants

Surfactant	Tradename	Nature	Molecular weight	Types of hydrophobic group	Types of hydrophilic group
Sodium petroleum sulfonate	SPS	Anionic	273.3	Alkyl	Sulfonate
Sodium lauryl sulfate	SLS	Anionic	288.3	Lauryl	Sulfate
Cetyl trimethyl ammonium bromide	CTAB	Cationic	364.4	Cetyl	Ammonium
Triton X-100	TX100	Nonionic	628.0	Octylphenol	PEO
Polyethylene oxide	PEO	Nonionic	3×10^4	Polyethylene	Oxide

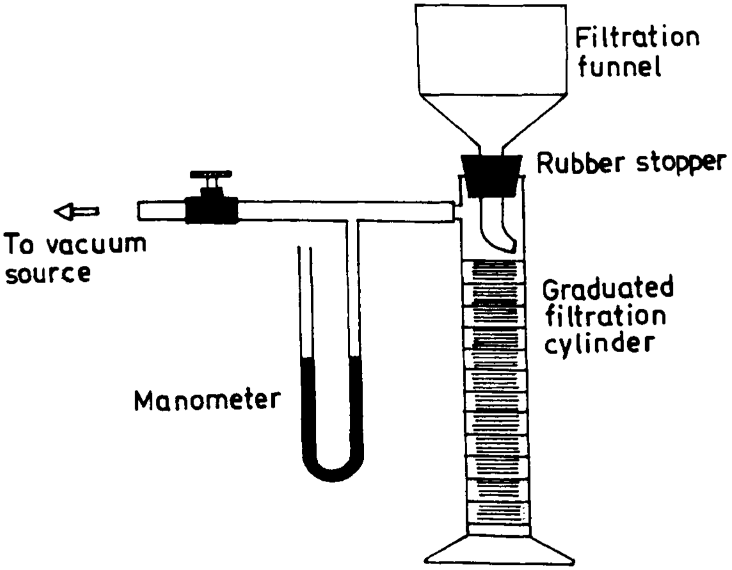

FIG. 1 Bench-scale dewatering equipment setup.

from the difference in the weight of the filter cake. Specific cake resistance and cake permeability were also calculated by using the integrated Darcy Eq. (4) (12).

Electrophoretic Mobility

Electrophoretic mobilities of the mineral particles were measured with a Rank Brothers Mark-II microelectrophoresis apparatus utilizing the two-electrode mode and a flat quartz cell. A detailed description of this apparatus and the procedure for preparing test suspensions have appeared elsewhere (13). The value of the effective interelectrode distance (L) determined at the start of the study was used throughout (13).

RESULTS AND DISCUSSION

Settling Characteristics

Our settling study shows that the settling rate of iron ore fines containing 6% solids by weight was very slow (2.52 m/h). It is well established that the settling rate and the flocs strength are of paramount importance in the filtration separation and dewatering process. The flocs should have sufficient strength to withstand pressure during filtration. The settling rate can be increased by flocculation using high molecular weight anionic polymers. Six different high molecular weight polymeric flocculants were employed to improve the settling rate of fines samples. Detailed results are shown in Table 4. Magnafloc 1011 and Rishfloc 258 (weakly anionic in nature) had a significant effect on the settling rate (90 m/h) at doses of 0.025 and 0.018 kg/t, respectively, but the flocs strengths were quite poor and it was therefore thought prudent to see if a combination of two polymers would improve upon the flocs strength. It was observed that apart from flocs strength, the settling rate was also increased significantly (180 m/h) at a dosage of 0.043 kg/t (1:1). The rate increased by a factor of 75 compared to an untreated sample. In general, it is seen that at higher dosages of flocculants, the settling rate decreases. This may be due to either restabilization caused by a higher concentration of anionic charges on a particle's surface or because the floc size is very small due to a higher number of polymer molecules interacting on the same number of particles. Figure 2 illustrates the response of the hindered settling rate of iron ore fine suspension to flocculant additions. The curves exhibit the characteristic optimum value of flocculant addition, beyond which further additions lead to an adverse effect on the settling rate. This type of behavior is quite typical and has been observed by a number of authors. Apart from an increased settling rate, flocculant filter aids produce higher throughout of the thickener and lower cake moisture, and they therefore allow for better performance of existing equipment or allow for a reduction of the equip-

TABLE 4
Summary of Flocculation Tests Results

Flocculant name	Time taken for complete settling (seconds)	Optimum dosages (kg/t)	Maximum settling rate (m/h)	Clarity of supernatant liquid
No flocculant	375	0.000	2.52	Very turbid
Magnafloc 592	18	0.037	50.04	Clear
Magnafloc 1011	10	0.025	90.00	Very clear
Magnafloc 351	22	0.037	40.68	Clear
Rishfloc 440MV	14	0.018	64.08	Turbid
Rishfloc 258	10	0.018	90.00	Very clear
Rishfloc 1226	13	0.025	69.12	Clear
Rishfloc 258 + Magnafloc 1011	05	0.025 + 0.018	180.00	Super clear

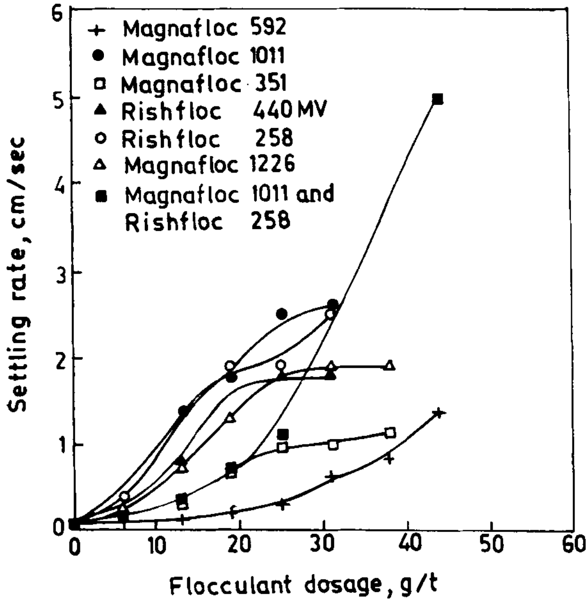

FIG. 2 Effect of flocculant dosage on settling rate.

ment scale necessary. Once we had selected the most effective flocculant and determined the required dosages, these were kept constant throughout our filtration separation and dewatering studies.

Baseline Dewatering Studies

Baseline dewatering data of the iron ore fines without using any reagents is shown in Fig. 3. After 7 minutes filtration time, the filter cake contained 18.20% moisture. Additional filtration time did not make any significant difference in final cake moisture. The electrophoretic mobility of the iron ore fines suspended in distilled water is shown in Fig. 4. The sample under investigation shows a point of zero charge (PZC) at pH 5.2. Normally the PZC of hematite in distilled water is 4.2–4.5 (14). In the case of hematite it is well established that H^+ and OH^- are the potential-determining ions in aqueous media (15) and that the charge density on the mineral surface is dependent on the net abstraction of these ions. The sample under investigation was from an iron ore washing plant, i.e., a field representative sample containing many other minerals and ions so that there was some shift in the PZC of the sample. The effect of PEO (the best surfactant in

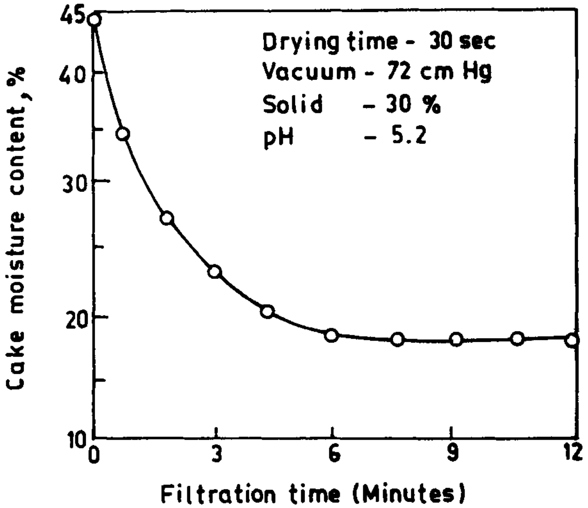

FIG. 3 Effect of filtration time on residual cake moisture of iron ore fines.

terms of moisture reduction) on electrophoretic mobility is also shown in Fig. 4. There was little shift in PZC from its original position toward the negative side. In general, the electrophoretic mobility becomes more negative with increasing surfactant concentration.

Effect of pH on Filtration Rate

Figure 5 shows the filtration rate with and without flocculant. The maximum flow rate was 5 L/h at pH 5.2 without flocculant; with flocculant it increased significantly to 49 L/h at the same pH. Thus, pH 5.2 was where the filtration rate was maximum.

Effect of Surfactants on Filtration Dewatering

The effects of different surfactants on the filtration and dewatering behaviors of the iron ore fines are shown in Figs. 6 and 7. Figure 6 shows the effect of SPS, SLS, and TX100 on the filtration rate and moisture content of the filtered cakes. It was observed that all surfactants enhance the filtration and dewatering rate and follow the same trend. The filtration rate first increased with increasing surfactant concentration. After reaching a maximum level, there was a decrease with increasing surfactant dosage. Similarly, the cake moisture first reduced with increasing surfactant concentration, but after reaching a minimum level, it remained more

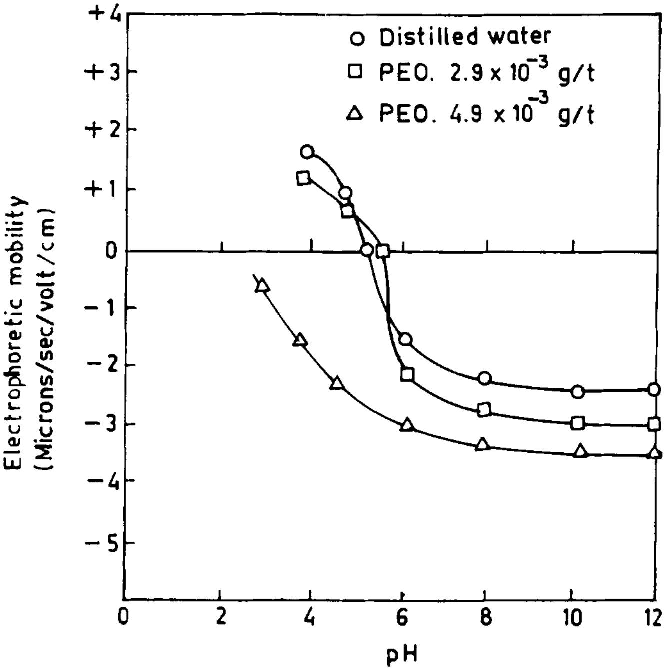

FIG. 4 The effect of PEO on the electrophoretic mobility of iron ore fines at 25°C.

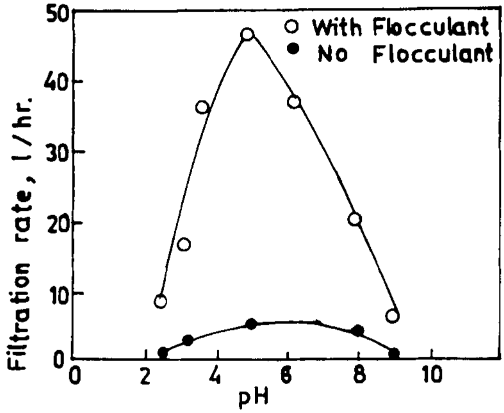

FIG. 5 Effect of pH on filtration rate.

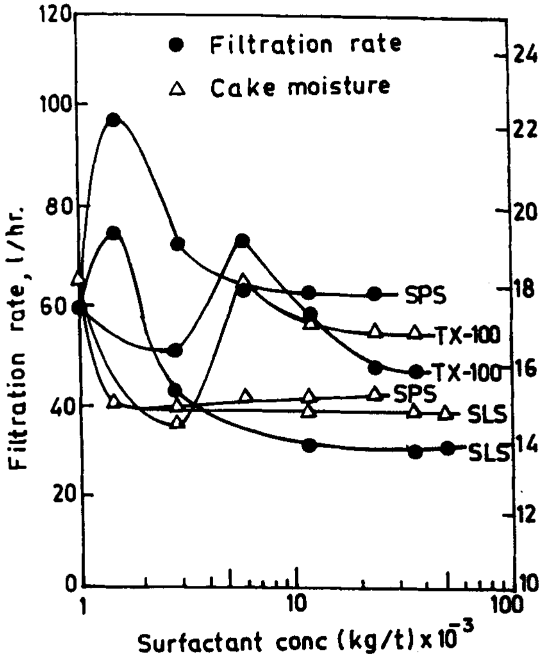

FIG. 6 The effect of surfactant concentration on the filtration rate and cake moisture contents.

or less constant. This is due to the fact that with increasing surfactant concentration there is a lowering in surface tension of the slurry before it reaches the critical micelle concentration (cmc). After reaching cmc, there is no appreciable change in surface tension and hence no appreciable change in cake moisture and filtration rate.

Figure 6 also shows the effect of TX100 on dewatering behavior. In this case the moisture content first decreases, then increases with increasing surfactant concentration, and then again decreases i.e., double minima and maxima. Many author have reported a double minima and maxima in the moisture content and filtration profile (15). The reason for a series of minima and maxima may be the interaction of the surfactant with the solids to be dewatered or with any organic material in solution (16). A similar behavior was also observed in the filtration rate. CTAB shows the same trend with respect to cake moisture (Fig. 7). The results obtained with PEO are also illustrated in Fig. 7. The minimum cake moisture content (12.6%) was obtained with PEO. In general, higher filtration rates

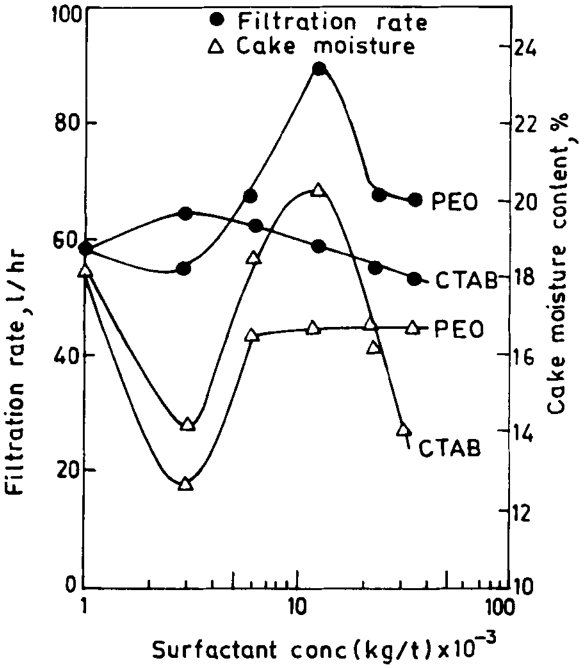

FIG. 7 The effect of surfactant concentration on the filtration rate and cake moisture contents.

were obtained in the presence of the surfactants together with a reduction in the final cake moisture content compared with untreated iron ore fines. The improvement in dewatering was observed only over a narrow concentration range, outside of which the surfactant did not show any appreciable difference. This observation is corroborated by the findings of other workers (17, 18). There was no definite pattern of change in the nature of dewaterability with respect to the surfactant concentration. Detailed comparative performance results of all our surfactant dewatering aids are shown in Table 5. The specific cake resistance and permeability were also calculated from the integrated form of Darcy's Eq. (4) (12). The maximum filtration rate and cake permeability were found to be 97.2 L/h and 1.2×10^{-12} m^2, respectively, with SPS at a dosage of 1.47×10^{-3} kg/t. The corresponding cake resistance was 1.0×10^{9} m/kg. However, the minimum cake moisture content of 12.6% was obtained with PEO at 2.94×10^{-3} kg/t.

TABLE 5
Comparison of Performance of Different Surfactants[a]

Surfactant tradename	Quantity (kg/t × 10^{-3})	Cake moisture (%)	Filtration rate (L/h)	Specific cake resistance (m/kg)	Cake permeability (m^2)
No (S + F)	—	18.20	4.8	8.6×10^9	1.3×10^{-13}
With F	50.00	15.30	8.2	1.3×10^{11}	9.2×10^{-15}
SPS	1.47	14.95	97.2	1.0×10^9	1.2×10^{-12}
TX100	2.94	14.45	51.6	5.8×10^9	2.2×10^{-13}
CTAB	2.94	14.00	64.8	2.4×10^9	5.2×10^{-13}
PEO	2.94	12.62	54.7	4.6×10^9	2.8×10^{-13}

[a] F = Flocculant. S = Surfactant.

Dewatering Kinetics

Figure 8 shows the dewatering kinetics in the presence and absence of chemical additives. It was observed that dewatering kinetics was very rapid and that steady-state saturation was achieved in less than 30 seconds with the most effective surfactant additive (PEO) whereas with the most effective combination of flocculants (Magnafloc 1011 and Rishfloc 258) and without any additive it was comparatively very slow. As a consequence of the lowered capillary pressure caused by the use of a surfactant, the driving potential for the dewatering kinetics is increased. In addition to an improved dewatering kinetics, the filtration rate is also accelerated. The increase of dewatering kinetics leads to a significantly lower moisture content in the product, even in the short times used for dewatering, thus making the process more efficient. Air throughputs increase with lowered moisture contents.

Economics of Using Flocculant Filter Aids and Surfactant Dewatering Aids before Thermal Drying

It is informative and illustrative to estimate fuel costs for running a thermal drier and to see how these can be offset by using filter aids and dewatering aids in filtration dewatering. This is estimated in the following stepwise systematic calculation.

1. Average fuel cost = Rs.9/L (July 1996 price)
2. Assumed calorific values = 4.19×10^7 J/kg
3. Energy required to evaporate 1 tonne of water at 100 C = 3.0×10^9 J

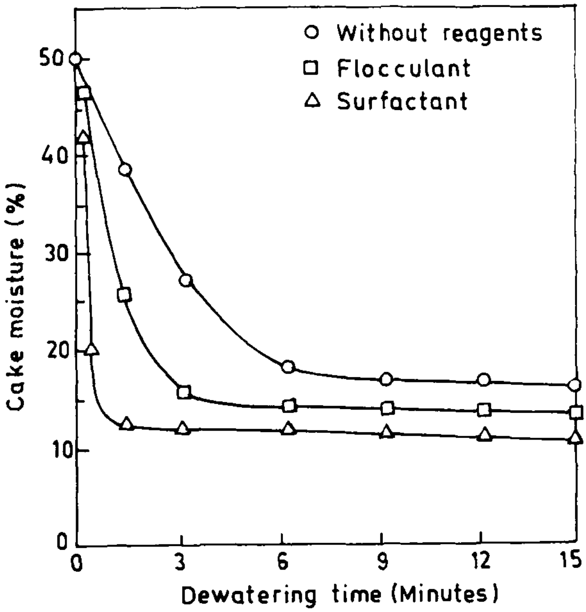

FIG. 8 Dewatering kinetics with PEO.

4. The weight of oil required per tonne of water removal (where the drier efficiency *n* is assumed to be 62.5%) (19)

$$= \frac{3 \times 10^9}{4.19 \times 10^7 \times 0.625}$$

$$= 114.5 \text{ kg}$$

$$= 127.2 \text{ L}$$

5. The energy cost for removing 1 tonne water

$$= 127.2 \times 9$$

$$= \text{Rs.}1145$$

6. Cost of fuel for thermal drying 1 tonne of solids containing various amounts of moisture is shown in Table 6. It is evident from the data that as the moisture content decreases from 30 to 5%, the cost is decreased from Rs.490.6 to Rs.64.1

TABLE 6
Cost of Fuel for Thermal Drying One Tonne of Solids Containing Different Amounts of Moisture

Percent moisture (w/w)	Kilograms water/ tonne dry solids	Oil cost for drying to zero moisture, Rs.[a]
30	428.5	490.6
28	388.9	445.3
25	333.3	381.6
22	282.0	322.9
20	250.0	286.3
18	219.5	251.3
16	190.5	223.3
14	162.8	186.4
12	136.4	156.2
10	111.1	127.2
5	52.6	64.1

[a] 1 US$ = ~35 rupees (Rs.).

7. Cost comparison with and without filter and dewatering aids is shown in Table 7. It indicates that with an effective surfactant such as PEO, the net saving in oil cost is Rs.89.5 per tonne of dry solid, whereas with a flocculant the oil cost saving per tonne of dry solid is only Rs.48.

TABLE 7
Cost Comparison With and Without Filter and Dewatering Aids[a,b]

Surfactant tradename	Surfactant dosages, (kg/t) × 10^{-3}	Moisture (%)	Kilograms moisture per tonne of solid	Oil cost for drying zero to moisture	Oil cost saving/t or dry solid
No (S + F)	—	18.20	222.5	254.8	0.0
With F	—	15.30	180.6	206.8	48.0
SPS	1.47	14.95	175.8	201.3	53.5
SLS	1.47	14.93	175.5	200.9	53.9
TX100	2.94	14.45	168.9	193.4	61.4
CTAB	2.94	14.00	162.8	186.4	68.4
PEO	2.94	12.62	144.4	165.3	89.5

[a] F = Flocculant (5 × 10^{-4} kg/t). S = Surfactant.

[b] Conversion: 1 US$ equivalent to approximately 35 rupees.

CONCLUSIONS

Major findings of the present study are as follows.

1. The settling rate of iron ore fines can be greatly improved by a combination of two low charge weakly anionic polymer.
2. Flocs formed by a combination of two polymeric flocculants of high molecular weight was very suitable for vacuum filtration.
3. A combination of flocculants increased the floc strength significantly to withstand filtration dewatering pressure.
4. Flocculant filter aids produced higher cake yields and therefore allow for greater throughputs and better performance of existing equipment or provide an answer to reducing the equipment scale necessary.
5. Significant improvements can be achieved both in filtration rate and cake moisture content by using a suitable surfactant. There is a critical concentration of each surfactant (below cmc) at which maximum filtration rate and minimum cake moisture can be obtained.
6. Depending on the specific needs of a particular industry, the type of surfactant with respect to maximum filtration rate or minimum cake moisture content can be selected.
7. A cost comparison between the fuel needed to run a thermal drier and filtration additives has shown that chemical treatments are very viable economic propositions in solid–liquid separation technology.

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REFERENCES

1. M. J. Pearse and A. P. Ailen, *Filtr. Sep.*, *236*, 22 (1983).
2. S. K. Mishra, *Tetra. Technol.*, *21*(3), 102 (1989).
3. S. K. Kuz'Kim and V. P. Nebera, *Synthetic Flocculants in Dewatering Process*, Boston Spa, NLL, 1966.
4. R. W. Slater and J. A. Kitchner, *Discuss. Faraday. Soc.*, *42*, 267 (1966).
5. F. J. Mangranite Jr., C. R. Leitz, and P. E. Galick, in *Organic Polymeric Flocculants: Effect of Charge Density, Molecular Weight and Particle Concentrations, Flocculation, Sedimentation and Consolidation* (B. M. Moudgil and P. Somasundaran, Eds.), Engineering Foundation, 1985, p. 139.
6. R. Koster, B. Burg, and M. Kaiser, *Aufbereit. Tech.*, *33*(5), 267 (1992).
7. S. J. Puttock, A. G. Fane, C. J. D. Fell, R. G. Robins, and M. S. Wainwright, *Int. J. Miner. Process.*, *17*, 205 (1996).
8. S. K. Nicol, *Proc. Australas. Inst. Min. Metall.*, *260*, 37 (1976).

9. H. B. Gala, S. H. Chiang, and W. W. Wen, *Proceedings of the 3rd World Filtration Congress*, Filtration Society, London, UK, 1982, p. 754.
10. S. K. Mishra. *Proceedings of the Conference on Industrial Practice on Fine Coal Processing*, Littleton, CO, 1988, 213.
11. G. Healy and S. R. Richard, *Fuel*, *69*, 53 (1990).
12. R. G. Holdich, *Filtr. Sep.*, *27*, 435 (1990).
13. Rank Brothers Ltd., *Staff Operators Manual for the Mark II Microelectrophoresis Apparatus*, Cambridge, UK.
14. R. J. Hunter, *Zetapotential in Colloid Science: Principles and Application*, Academic Press, London, 1981.
15. V. N. Moraru, F. D. Ovcharenko, Z. M. Yaremko, and R. L. E. Moraru, *Kolloid Zh.*, *42*, 870 (1980) (in Russian).
16. S. J. Puttock, M. S. Wainwright, R. G. Robins, C. J. D. Fell, and A. G. Fane. *Improved Dewatering in the Filtration of Alumina Trihydrate Trace Additions of Cationic Surfactant*, Personal Communication.
17. C. C. Mwaba, *Miner. Eng.*, *4*(1), 49 (1991).
18. G. Stroh and W. Stahl, *Proceedings of the European Symposium on Solid–Liquid Separation*, Cologne, Germany, 1992, p. 1.
19. R. J. Wakeman, International Conference on Separation Processes Technology, November 1979, Plymouth, UK.

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