Design tools to control transients in solvent extraction plants

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Abstract

The design of solvent extraction plants has been generally based on the assumption of steady operating conditions in both the settler and the mixer units. However, recent experience in both large and small plants has shown this not to be the case. Mixer transients include start up aqueous locking and partial phase separation in secondary mixers, which both lead to reduced plant capacity or overflow of upstream mixers. Analysis of the hydraulic pressure profiles has generated new design methods to eliminate the effects of the partial or complete phase separation. Changes to the method of specification and selection of primary and secondary mixers have been made to eliminate the onset of partial separation during operation. The observation of settler operation has shown that it is also a dynamic process, even under "steady-state" conditions. Four transient phenomena have been identified that lead to internal circulating flows, poor disengagement of fine secondary haze and erratic and high entrainment levels in the discharges. A first approximation analysis has shown that there are significant pressure differentials within settlers, and these change along the length and with the phase separation rates. A technique to control the effects of these transients was developed that allows plant performance to be improved beyond "standard" expectations. The design and selection of the settler internals is described.

Key words: Solvent extraction, Mixer-settlers, Phase separation, Entrainment, Mixer transients

Summary

The mixer transients are generally caused by fast-separating emulsions. The transients cause an accumulation of aqueous in the secondary mixers. It is most common during commissioning and with very clean organic solutions. Treatment relates to the design of the inlet to the secondary mixer to allow symptomatic relief of consequences of partial phase separation and the creation of a more stable emulsion that does not separate as quickly.

Methods were developed for controlling settler entrainments and mixer transients in copper mixer-settler units. For settlers, this involves the placing of a randomly packed medium within the settler. The medium serves to promote coalescence, thereby, reducing the subsequent entrainment. It also serves to prevent the propagation of transient and high velocities within the settler. These transients are thought to be the cause of the extreme peaks in entrainment often seen in settler exit streams. Many of the problems associated with entrainments can be attributed not to the average entrainment but to the very high peaks. Elimination of the peaks benefits the operation. The addition of coalescing medium to solvent extraction mixer-settlers is not only applicable to the copper industry but also to other metal extractions that use the same type of equipment.

The technique consists of the use of a low-cost, readily available, medium system installed within the settler unit. Previous attempts to use other types of performance promoters have suffered from poor economics, poor performance (due to blockages caused by byproduct crud accumulation), difficult operation and difficult housekeeping.

The main advantage of the media system is that it is low cost, highly effective, has very good economics, is tolerant of crud production and can easily be cleaned and kept in good condition. It can also be readily fitted to the existing settlers (of any design) to improve their performance.

Introduction

A method was developed for controlling entrainments in solvent extraction (SX) mixer-settler units. This involves the placing of a randomly packed medium within the settler. The medium serves to promote coalescence, thereby, reducing the subsequent entrainment. It also serves to prevent the propagation of transient and high velocities within the settler. These transients are thought to be the cause of the extreme peaks in entrainment often seen in settler exit streams (Miller et al., 1996a). Many of the problems associated with entrainments can be attributed not to the average entrainment but to the very high short-term peaks. Elimination of the peaks provides great benefit to the operation.

The development of the technique and the understanding of the underlying mechanisms have been an iterative process. The technique was initially applied to address an immediate problem that had reduced the capacity of a client plant (Miller, 1996b). Subsequent experience showed that a deeper understanding of the hydrodynamics inside commercial settlers was needed. Development of these models has led to further refinement of the application and selection criteria for the technique.

The problem of mixer transients has not been widely reported in the literature, except in the context of aqueous

Preprint number 01-148, presented at the SME Annual Meeting, Feb. 26-28, 2001, Denver, Colorado. Revised manuscript received and accepted for publication November 2001. Discussion of this peer-reviewed and approved paper is invited and must be submitted to SME Publications Dept. prior to Nov. 30, 2002.

accumulation in the secondary mixer or mixers after shut down (aqueous locking). The experience with clean organic solutions (particularly during commissioning) has shown that a similar problem exists with fast-breaking emulsions, creating an accumulation of aqueous that prevents full hydraulic capacity of the mixer train being achieved. This problem has also been addressed by a number of progressively more accurate models that have improved the understanding of the problem.

Adverse operation conditions in SX plants

Settlers. The exit separated phases from the mixer-settlers always contain some of the other phase due to incomplete separation (entrainment). This entrainment interferes with process efficiency, upsets other downstream processes and is an economic loss to the operation. The level of entrainment that can be accommodated by the subsequent process steps determines the capacity of a settler. Improvements in the performance of a settler reduces the loss of expensive reagents and allows for increased capacity of the plant.

Entrainment is minimized in *one* of the two exit streams by selection of the appropriate operating continuity in the mixers. Aqueous continuous operation has droplets of organic dispersed in an aqueous continuum; whereas organic continuous conditions have droplets of aqueous in an organic continuum. Aqueous continuity generally minimizes the entrainment of aqueous in organic, organic continuity minimizes the entrainment of organic in aqueous.

Adverse operating conditions in the SX process increase the level of entrainments, leaving the settlers reducing their capacity. Some of these situations were described by Miller (2000a). Generally, the adverse operations result in either longer phase-separation times (break times) or much faster break times, both of which result in higher exit entrainments and lower settler capacity. Reduction in these entrainments leads to recovery of plant productivity and increase of the settler capability.

Some of the more common adverse conditions are as follows:

- Low operating temperatures, leading to high solution viscosities, slower break times and higher entrainments.
- Reduction of interfacial tension by adsorbed species (e.g., polymer flocculants and mineral oils) that increase emulsion stability, leading to longer break times and higher entrainments.
- Reduced interfacial tension from oxidation by species in the pregnant liquor solution (PLS) or electrolyte such as permanganate (Miller, 1995), which produce very stable emulsions, very long break times and very high entrainments.
- The need to operate in the "incorrect" phase continuity to minimize entrainment. This can be the case in cold conditions where the lower entrainment is with the alternate phase continuity.
- Operation with very fast breaking emulsions that do not form an emulsion band in the settler. The emulsion band serves to filter the secondary haze that is formed when primary droplets coalesce. The absence of any emulsion band means that the haze is not removed from the separated phases.
- Operation in organic continuity (of normally aqueous continuous units) to eliminate the production of fluffy, silica crud. Organic continuous operation produces a compact crud that separates from the two main phases and does not flow through the whole circuit.

The control of the exit entrainments from the settler, is the prime focus to minimize adverse effects on plant production. The remediation of any underlying problem can be addressed on a longer-term basis, provided the short-term, high-entrainment symptoms can be addressed. It was the need to address these short-term symptoms that led to the initial development and use of the medium in the field.

Mixers. The problem of mixer transient performance is generally caused by the preseparation of the phases before delivery to the settler. The accumulation of aqueous causes a higher than normal density in the mixer. Most SX secondary mixers are bottom fed through a baffle arrangement that presents emulsion into the up-pumping impeller. The increased mixer density requires an increased operating height in the upstream mixer to drive the emulsion under the feed baffle. The increased operating height can cause overflow of the upstream mixer either to the outside or into the top of the "separated" mixer. This can further reinforce the separated condition leading to a major loss of hydraulic throughput and chemical efficiency.

The newer design of primary pump mixers that minimize the secondary fine haze production are the most prone to this type of transient behavior. Fine haze can stabilize the emulsion to an extent that separation is not occurring in the mixer. The low power input and high hydraulic efficiency of the secondary mixers also exacerbate the problem.

Previous in-settler entrainment control techniques

A number of other techniques have been developed in an effort to either increase the settler capacity or to control exit entrainments (Miller, 2000b). None have achieved wide acceptance in the mineral processing industry to date. The reasons for this are not clear but operational anecdotes would indicate that intolerance to the presence of crud is a major factor.

The most promising was that developed by Lewis in 1977. A number of baffles were placed in the path of the emulsion flow to hold it back and to provide longer residence times for coalescence and separation. No further installations were made after the initial one at Cockle Creek. Some installations of the IMI settler were made (Ritcey, 1977), but no recent reports have been circulated. Control through the use of tightly woven filaments (Knitmesh®) has been abandoned due to blockage by crud, despite very high settler capacities when treating clean solutions.

Description of the coalescing medium system

The material used is a low-cost readily available PVC pipe cut into short lengths. This can be sourced in all parts of the world and provides an effective material to achieve settler performance improvement. It is kept in 1m^3 mesh bags to prevent escape and to allow removal for cleaning without shutting down the plant. The medium for installation in a large settler is illustrated in Fig. 1.

Selection criteria for coalescing medium. Selection criteria that were developed for stream specific coalescing were upscaled (on the basis of the required output performance) to develop similar criteria for in-settler coalescing medium. Field-trial measurements confirmed the expected performance in both "conventional" and Krebs-type mixer-settler units. A combination of two selection criteria has been found to give reliable results: a minimum residence time and a specific area flow rate (cubic meters per hour of settler total flow per square meter of medium surface area). Performance in all plant installations has confirmed this selection basis.

The medium sizes used to date have varied from 63 to 90 mm. The size selection is based on the useable settler volume to receive the medium and the availability and cost of the raw materials and cutting services. All sizes performed to expectation when the appropriate selection criteria were used.

Use of coalescing medium in mixer-settlers. The use of coalescing medium in copper mixer-settlers is a relatively new technique for the control of peak levels of entrainments in the exit streams. The aim of the installation is to allow flexibility in operation of the units without entailing the high risks of entrainment that accompany certain operating modes. In operation, the medium performs the functions of coalezcing, conrol of settler transients and control of silica crud production.

Coalescing: Under the coalescing function, the medium does the following:

- Coalesces fine bubble haze to reduce the overall level of entrainment.
- Coalesces the macro emulsion to increase the volumetric capacity of the settler.
- Holds back the dispersion band to enable operation in the "deep dispersion band" regime to minimize the entrainment of secondary haze formed during coalescence. A discussion of this style of operation is given in Miller (1996a).

Control of settler transients: For the control of settler transients, the medium does the following:

- Provides hydraulic baffling to prevent the propagation of high-speed wave fronts travelling through the dispersion band, which lead to high and erratic entrainment levels.
- Contains the large circulating and linear flows of the emulsion and separated aqueous and organic phases near the settler feed. A detailed discussion of these operating aspects is given in Miller (2000a).

Control of silica crud production: It is the coalescing function that allows the mixer operation to be altered from one operating continuity to another without entailing undesirable entrainments. For instance, it is possible to operate in organic continuity, to minimize fluffy (silica-based) crud production, without having large aqueous entrainment in the exiting organic. In fact, is was the need to minimize crud production in the Girilambone Copper Co. (GCC) plant that lead to the development of the technique (Readett, 1995). Many plants have seen the benefits of operating in organic continuity but have not had the benefit of entrainment control to ameliorate the effects on the downstream operations.

Figure 1 — In-settler medium in 1-m^3 bags for installation in a large copper settler (total volume 120 m^3).

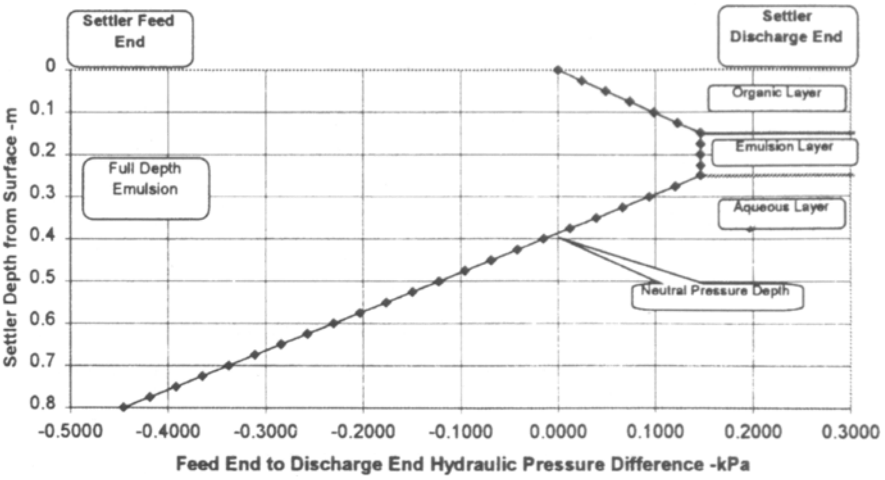

Figure 2 — Pressure differentials with an emulsion band present.

Development of a settler dynamic model

The effectiveness of the medium in controlling the exit entrainments led to the questioning of the actual mechanism, or mechanisms, that was occurring within the settler. Little work has been published on the operation and design of large industrial settlers since the 1970s. Because of this, there has been little fundamental development in the understanding of these units.

Hydraulic pressure considerations. Miller (2000a) describes the history and published observations of settler dynamic operation. The operation of large settlers was described as "violent" (particularly at the feed end) but no questions have been posed as to the reasons for this. Consideration of the macro pressure difference at levels throughout the settler shows that there is a changing differential pressure from the top surface to the base of the settler. This is illustrated in Fig. 2, where an emulsion band exists. Figure 3 shows the magnitude and direction of the flows that would be produced by these pressure differentials.

It can be appreciated that the induced flows will cause the dynamic nature observed in large settlers. The high forward

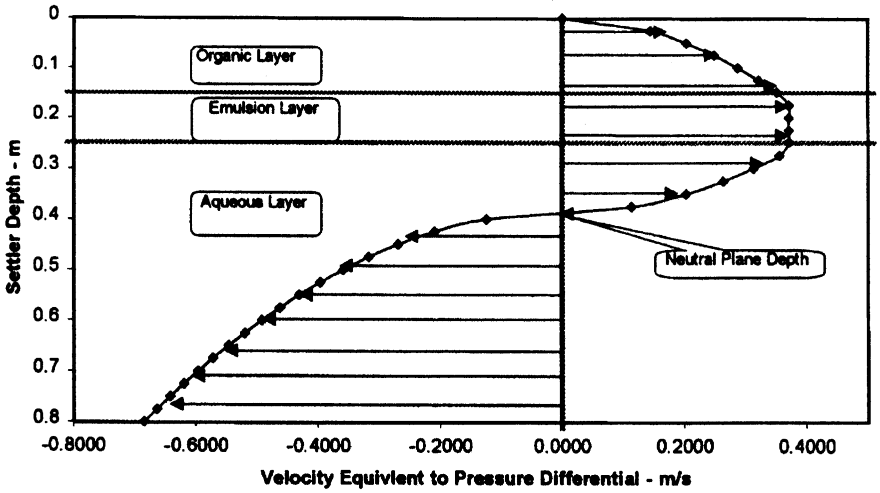

Figure 3 — Velocity profiles with an emulsion band present.

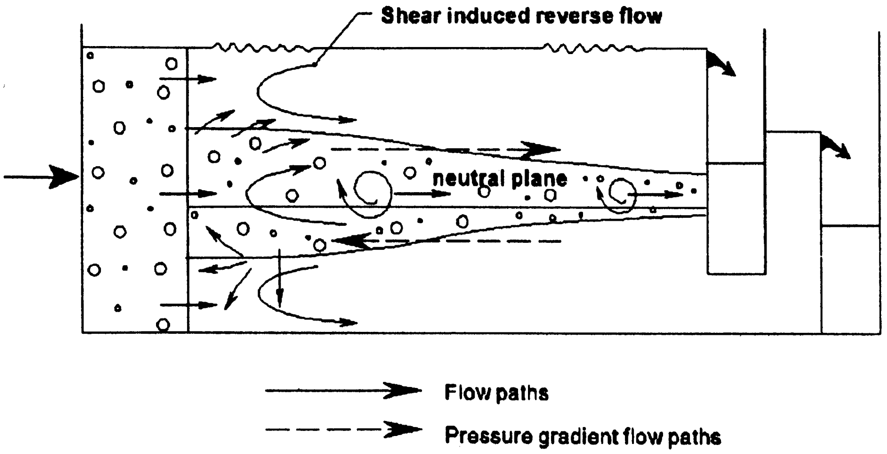

Figure 4 — Flow patterns in a settler with deep emulsion band at the discharge.

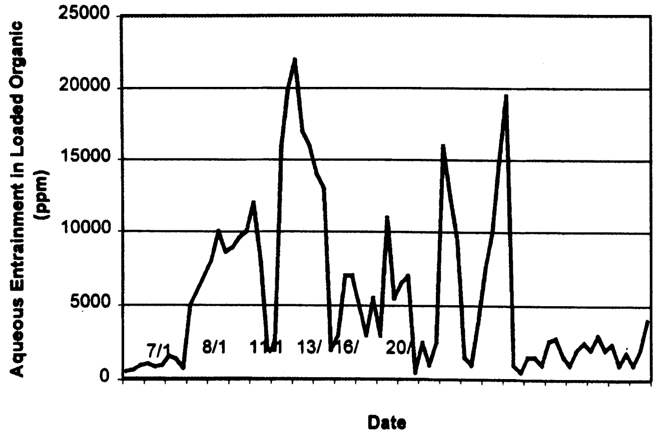

Figure 5 — Settler exit entrainments with dispersion band at the discharge.

velocity in the emulsion layer (or deeper organic layer) causes a vertical circulating flow pattern in the organic layer, which is most easily recognized as a "reverse" flow of organic (from the discharge end to the feed end) in the first half of the settler. A similar but more forceful vertical circulation pattern is also set up in the aqueous layer (Miller, 2000b). It should be noted that the maximum forward velocities calculated from the model are an order of magnitude greater than the design-phase space velocity for the settler.

Phase separation flows. The phase separation flows are vertically upward and downward from the emulsion layer. Where these are fast, i.e., at the feed end of the settler, they can have a significant effect on the internal flow patterns within the emulsion layer itself. With a combination of vertical separation flows and differential pressure-induced horizontal flows, high-speed circulating flows can be set up within the emulsion layer. These are illustrated in Fig. 4.

When the emulsion layer extends all the way to the discharge end, it can form a bounded channel in which the circulating flows can propagate linearly as roll fronts. The speed of the roll front propagation is likely to be high and likely to reflect the velocities shown in Fig. 3. When the roll fronts dissipate their energy against the discharge end wall, they cause the high and erratic entrainments shown in Fig. 5.

Fast separating emulsions. The separation of emulsions with fast breaking characteristics is another area of settler dynamic operation. The coalescing process involves the expulsion of a fine secondary droplet, as residual droplet surface tension causes the most distant portion of the droplet to contract away from the coalescing interface. The residual emulsion layer normally filters this secondary haze because the separated phases pass through it to the bulk phase in the settler. If the emulsion band is not present within the settler, then no filtering can take place and the haze is carried to the discharge. This type of operation is illustrated in Fig. 6. The discharge entrainment is well above the design expectation and is highly erratic in both its magnitude and temporal pattern.

Effectiveness of the coalescing medium systems

Entrainment control. The medium has shown that it can reduce the operating entrainment of organic in the raffinate and electrolyte to around 20 ppm. The normal expectation for raffinate entrainment loss is 50 ppm or higher. A 20-ppm loss is a realistic expectation when using the medium. The medium installation has a payback of less than four months for final extraction stages.

The medium in the first series extraction stage (E1) can reduce the average entrainment of PLS in the loaded organic from around 800 ppm to a steady 250 ppm. This is based on the experience at a number of operations using it. A reduction in the PLS entrainment will give a corresponding reduction in the amount of acid and cobalt lost in the EW iron control bleed. Again, this has very short payback periods.

Hydraulic capacity. Another benefit of the medium is the increased hydraulic capacity that can be handled successfully by the settler. An approximate 20% "overcapacity" is possible using this technique. This is demonstrated in Fig. 7, which shows the increase in capacity of a copper SX plant over time as medium is added to the five settlers in the circuit. In this case, the increase in capacity was from 50% of design to 100% of design. The adverse operating conditions were a result of permanganate degradation of the SX reagent and poor settler size-selection criteria.

Control of transients. The transient operation in settlers can show itself in high and erratic levels of entrainment in the exit streams. The medium eliminates these transients and returns the operation to within the design. The effectiveness is illustrated in Fig. 8, which shows the settler entrainment before and after installation of the settler medium.

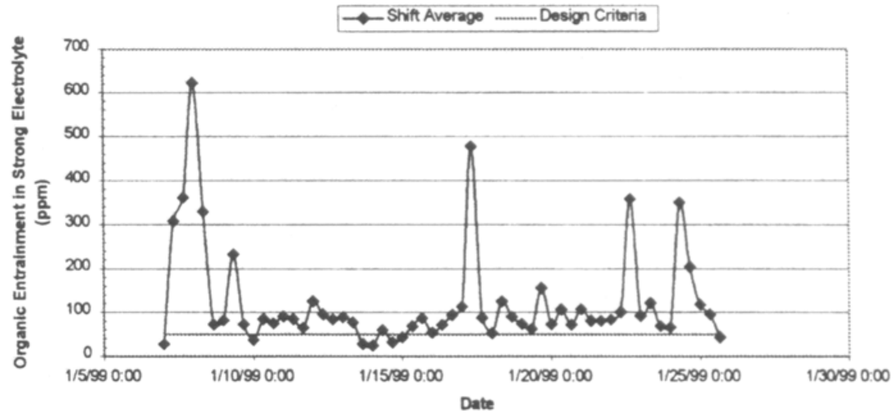

Figure 6 — Settler exit entrainments with no dispersion band.

Effect of medium in the settler. The effect of placing medium within the settler is shown in Fig. 9. The dynamic section of the settler is now constrained by the medium and all pressure differentials, circulating flows and major separation flows, occur in a short portion of the unit. The remainder of the settler is quiescent, and no dynamic operation is observed. The emulsion band is eliminated in the discharge end of the settler and there is no channel in which roll fronts can propagate. The medium also acts to damp any circulating flows in the bulk of the settler volume, as well as dissipating the energy and flow patterns of any roll fronts that impact it.

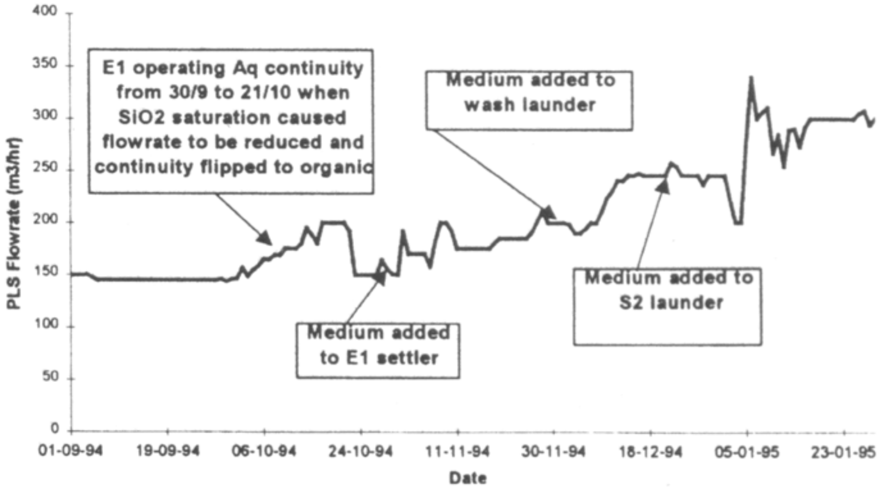

Figure 7 — Plant capability increase with addition of ECA medium.

Operating flexibility. Another benefit is the opportunity to operate the mixers in any continuity without uncontrolled entrainments resulting. As a corollary, any uncontrolled change in continuity will not result in a major process problem from the massive entrainments that would have otherwise resulted. For example, if E1 flips from aqueous to organic continuous, the PLS entrainment in loaded organic can jump from 500 ppm to 25 g/L within minutes and cause major problems in the EW. Similarly final extraction stages changing from organic to aqueous continuous would cause major losses of organic to raffinate. Installation of the medium minimizes the entrainment of organic in the raffinate and prevents significant economic loss as a result.

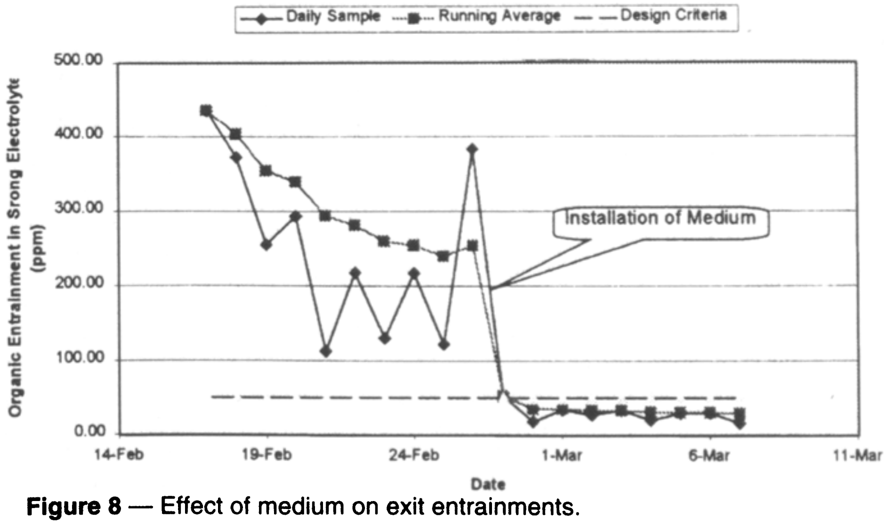

Figure 8 — Effect of medium on exit entrainments.

A further benefit of the medium is the ability to run the mixers in organic continuity for control of the effects of silica in the PLS. This often helps to eliminate "fluffy crud" and crud runs caused by high levels of silica. This has allowed two operations to regain control of their SX plants under adverse silica conditions (Miller, 1996).

Operation with settler medium

The medium does retain a small amount of crud, and other solids do build up on it. These contribute to a loss of effectiveness due to the modification of the surface tension of the medium. The buildup also increases the operating pressure drop, sometimes to a level that limits the settler hydraulic throughput. To prevent the occurrence of these situations a periodic cleaning program is implemented.

Periodic cleaning. Cleaning of the medium is required whenever the effectiveness is reduced or when the hydraulic pres-

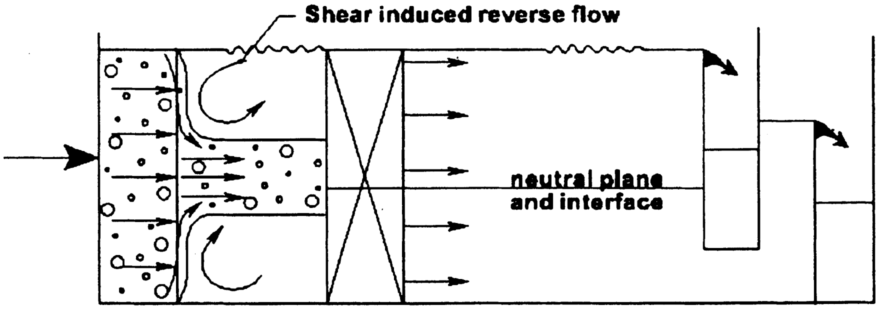

Figure 9 — Settler operation with coalescing medium.

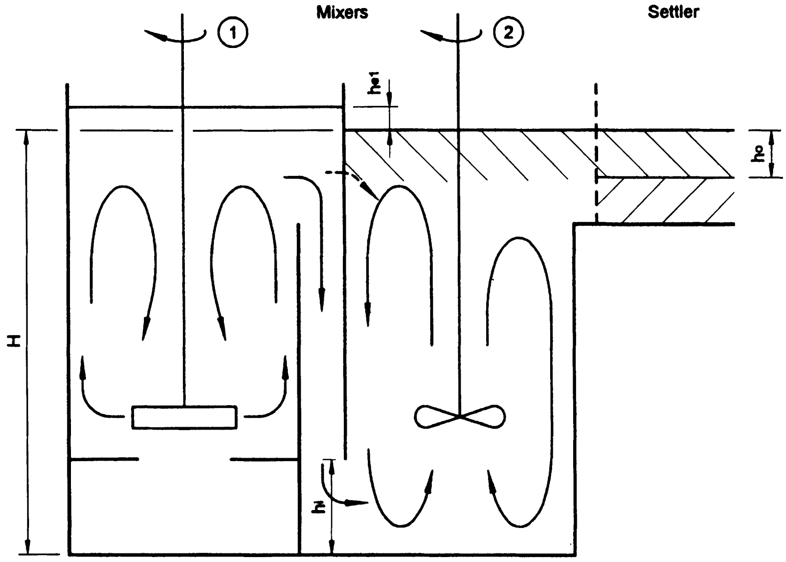

Figure 10— Mixer schematic design.

sure drop is too high. The loss of effectiveness is generally the first symptom to be noticed, and incidences of extreme pressure drop are unusual. The cleaning program is integrated with the regular preventative maintenance plant shut downs. Each settler can be cleaned on a regular basis during these periods without the need for a specific shut down for cleaning.

Cleaning is carried out with the flow to the unit reduced or stopped. This prevents the solids removed from the medium from continuing through the circuit and generating a crud run. Hosing with a high-pressure water stream is normally sufficient to clean the medium.

This can be done in situ with the settler drained of operating fluids or by removing the bags of medium and cleaning them outside the settler. The first technique does not require the removal of the bags from the settler but has the risk of contributing solids to the settler streams on restart. The water used for cleaning is recovered and treated in the crud treatment plant for recovery of the organic and recycling of the water into the process.

Cleaning should not be attempted while the settler is in full-flow operation. By definition, the release of the solids and crud from the medium will contribute to a crud run and further process upsets. If excess crud in the SX circuit is a continuing problem requiring a more regular cleaning schedule, then serious consideration can be given to altering the operating continuity to organic continuous to change the crud type from fluffy to a more compacting one. Cleaning of the medium to ensure its maximum effectiveness should precede alteration from aqueous to organic continuity. Continued good plant operation relies on the medium effectiveness being as high as possible for long periods.

Aqueous locking models

Increased Mixer 1 operating height. The geometry of a series of mixers is shown in Fig. 10. The first model developed assumes that the measured secondary mixer O:A approximates the average O:A in the mixer.

This is probably close to the mark, because mixers generally have a bottom entry and because they overflow at the top with the sample point close to the halfway mark up the tank wall. At hydraulic equilibrium, the feed emulsion density (ρ_{ei}) must equal outflow emulsion density (ρ_{eo}) (which is the same as the emulsion density from the first mixer ρ_{e1}) independent of the maximum emulsion density in the mixer (ρ_{emax}).

Assuming that the continuum of emulsion densities (from the bottom to top of the second mixer) can be approximated by the measured emulsion density ($\rho_{e2} = \rho_{emax}$). The *driving* pressure on the *inlet of new emulsion* into the second mixer can be equated to the back pressure on the inlet of new emulsion into the second mixer due to the increase in the average emulsion density.

$$h_{e1} = h_i[(\rho_a - \rho_o)(R_1 - R_2)]/[(R_2 + 1)(R_1\rho_o + \rho_a)] \quad (1)$$

This relationship was calculated and plotted as a function of the advance O:A (R_1) and the measured O:A in the second mixer (R_2), for copper stripping mixers, and is shown in Figure 11. This type of relationship shows a dependence on both the advance O:A and the measured O:A in the second mixer. It is independent of the total mixer height. It is inherently useful, as it can be directly related to field measured quantities. At one plant studied, the magnitude of the predicted head increase was within the expected plant performance range, as it still showed that the plant would be operable albeit with a reduced splash allowance under the conditions experienced.

It is also of interest to note that the increased head in Mixer 1 reduces as the advance O:A decreases, i.e., as ρ_e increases. Again, this is consistent with plant experience where the strip mixer has been operated with low O:As to achieve emulsion throughput of the second mixer. The fact that the head in Mixer 1 decreases as the advance O:A decreases is intuitive, as one would expect less head would be needed with a greater inflow density from Mixer 1.

Entry-depth calculation. The analysis of the maximum entry depth requires a slightly different approach. In this case, the level of the inlet to the secondary mixer needs to be assessed in terms of the increased head in Mixer 1 that is required to overcome the separated phases in Mixer 2. The worst situation arises when the settler and last mixer are in hydraulic contact after shut down of the plant (true "aqueous locking"). The separated organic flows into the settler and a corresponding volume of aqueous flows back from the settler into the mixer. As a result, the mixer has increased hydraulic head from the deeper layer of aqueous in it.

The depth of organic is the same as that in the settler, and the increased head required to overcome the accumulated aqueous (with its higher average density) needs to be defined. This could be a proportion of the freeboard in the upstream mixer if there is no high-level connection between mixers. Alternatively, it could be the level of the mixer interconnection, e.g., the top of the downcomer baffle. Any further increase above this level would allow emulsion to flow in and progress over the top of the second mixer and out into the settler, without any emulsion actually being mixed by the second impeller. This further exacerbates the problem with even more aqueous dropping out of the bypassing emulsion and accumulating in the mixer.

When emulsion is *just flowing* into the second mixer, the pressure at the *top* of the inlet area must be equal from both the upstream mixer and internally to the mixer of interest. In this mixer, the total organic depth is small in relation to both the total depth and the depth of aqueous.

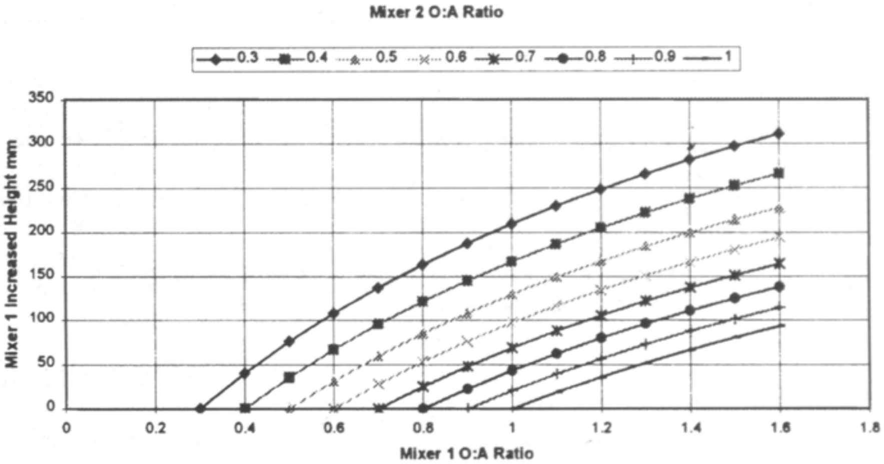

Figure 11 — Increased strip Mixer 1 height with varying O:A in Mixers 1 and 2 (inlet depth 2.5 m).

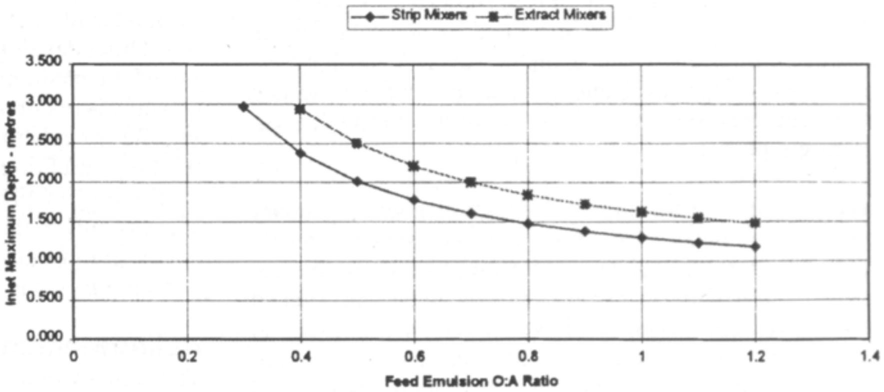

Figure 12 — Second mixer maximum inlet depth from surface for mixer start up without aqueous lock.

Pressure at the inlet from the head in Mixer 2 is

$$P = g(h_{os}\rho_o + (h_i - h_{os})\rho_a) \qquad (2)$$

Pressure from the head in Mixer 1 is

$$P = g\rho_{e1}(h_{e1} + h_i) \qquad (3)$$

Equating, rearranging and expressing in terms of R_1 gives

$$h_i = h_{os}(R_1 + 1)/R_1 + h_{e1}(R_1\rho_o + \rho_a)/R_1(\rho_a - \rho_o) \qquad (4)$$

Values of h_i generated by this equation have been plotted as a function of the advance O:A in Fig. 12. It is evident that the higher O:A ratios require shallower inlet depths and the strip mixers require shallower inlets than the extract mixers due to the higher aqueous density in the strip circuit. It is also evident that the inlet depth is independent of the total mixer depth.

The recent trend in SX plant design is to use up-pumping secondary mixers rather than the traditional down-pumping ones used in slurry tanks. With these impellers, the flow pattern at a shallow inlet from the up stream mixer is actually down the wall, as shown in Fig. 10. As a result, there has been no evidence of short-circuiting in those mixers with "shallow" inlet depths, and mixer efficiencies have remained above 95% in such tanks. The flow of the bulk of the fluid down the walls of the mixer is an unlooked for advantage of the "normal" specification of up pumping mixers.

Aqueous balance pipe. Aqueous locking can also be addressed by providing a mechanism for removal of the accumulated aqueous. In the past, this has taken the form of a "balance" pipe between the secondary mixer or mixers and the primary mixer (inlet below the false floor). The connection is opened after the primary mixer is operating and the contents of the secondary mixer or mixers are drawn into the primary mixer. In this way, the aqueous accumulation is mixed with advancing phases into a (hopefully) stable emulsion that flows through the mixers into the settler. The driving force for the flow of aqueous into the primary mixer is a combination of the higher hydraulic head in the secondary mixer or mixers and the low pressure caused by the pump mixer under the false floor in the primary mixer.

A number of design criteria can be used to size the connection between the mixers. One of the more important is that of maintaining the preferred operating continuity. This generally means that the flow of aqueous should be less than ten percent of the normal aqueous flow. Under these circumstances, the mixer O:A will fall to 0.9:1, but the operating continuity should not be effected to any great extent. With most mixer designs having a residence time of one minute for the emulsion, the time to empty the mixer of excess aqueous will lie somewhere between 10 and 20 minutes before the second mixer is full of only emulsion once more. Under these circumstances, it is seen that a high-level interconnection is still required to prevent short circuiting or overflow occurring during the period of removing the accumulated aqueous.

Another method of removing the aqueous accumulation is to start the whole plant at reduced organic flows and use the

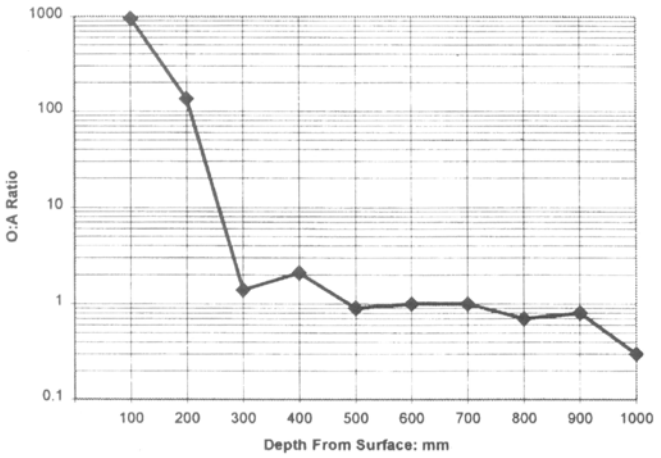

Figure 13 — Strip mixer O:A with depth.

capacity of the pump mixer to remove the aqueous accumulation with no advance aqueous. This needs a larger connection between the two mixers to reduce the time needed to regain operating equilibrium. Once again, however, it is still necessary to have a high-level interconnection between the two mixers to prevent overflow or short circuiting during the start up transient period.

Mixer designs. The development of high-efficiency, primary pump mixer impellers has eliminated the need for severe baffling of square tanks and their only purpose is to reduce the swirling motion of the bulk emulsion. Advantage can be taken of the square cross section of the mixer (where used) to minimize the baffling required. Two baffles at 180° are sufficient to eliminate the swirling. These should not extend through the surface as they will inhibit the smooth flow of the emulsion to the subsequent process vessel, as well as providing resistance to the re-entrainment of separated organic at start up.

Use of egg crate baffles in the primary mixers has shown that they can cause high operating levels if there is full baffle depth along the length of the baffle blade or blades. The enclosed space or spaces act as regions where the velocity is essentially nil and as such convert the internal mixer solution velocity into a static head. This head is held within the closed baffle space as a higher operating level than would otherwise occur. Internal mixer velocities of around 2 m/s are not uncommon, which translates into a static head of approximately 200 mm. In many cases, this can be sufficient to cause overflows of the primary mixer at either the nuisance or the severe level. With the static conditions within the baffle depth, phase separation occurs and organic accumulates in this region. Figure 13 shows the sampled O:A ratios as a function of the depth from the operating surface of a secondary mixer. The depth within the baffle (zero to 300 mm) is almost all organic, whilst below the baffle (300 to 900 mm) the O:A ratio is around the advance O:A. The O:A ratio drops further with depth and begins to indicate an aqueous locked mixer with very low O:A ratios at greater depths.

The analysis of the high-level (shallow) down-comer entry has highlighted the fact that no down-comer is actually required in a tank that is using an up-pumping impeller. A short baffle may be appropriate to stop any short-circuiting across the top of the mixer and to present the feed into the down flowing stream created by the mixer. This design approach will increase the *active* volume of the mixer, reduce the cost of mixer internals and simplify the construction of the tank.

Potential for future developments

There are a number of developments that are in progress to improve the technology and expand the areas of application. These involve the modeling of the settler transient processes by computational fluid dynamics (CFD). Use of this technique will provide optimization of the placement and selection of the coalescing medium characteristics to achieve the design goals. This modeling could also lead to the ability to design smaller, cheaper plant units using higher unit capacities. Other developments are being investigated in the cleaning of the medium and the improvement in life cycle before cleaning is required. All of these provide potential for increasing the acceptability of the technology and improving the working conditions of the plant operators.

The use of CFD for modeling the mixer circulating pattern should confirm the field experience that the mixer efficiency is not compromised by the high level inlet when an up pumping mixer is utilized.

Acknowledgments

The author would like to express appreciation to the directors and staff of the clients' operations for their assistance and cooperation during execution of these projects and development of the procedures and protocols for use of the techniques described in this paper.

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