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Tailings thickening challenges at Kamoa Copper

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Abstract

Kamoa Copper Mine is already the world's third-largest copper producer, and it has the ability to grow further. Start-up of the Phase 3 Kamoa concentrator has thrown up some interesting issues with the flocculation and thickening of the tailings, and these are discussed, along with the corrective actions taken. The plant was suffering from almost continuous 'sliming', where the bed rises within the thickener, eventually overflowing solids, compromising the process water for the whole flotation circuit.

The case highlights several problems commonly encountered with thickeners when starting-up, including product selection issues and initial operational difficulties. It shows how working through the issues systematically and applying experience from other operations helps. Simple static cylinder testing has been used extensively and becomes a valuable troubleshooting tool. It is a technique that, with a little training, most metallurgists can use.

Keywords: *thickener, cylinder testing, coagulant, flocculant, settling rates, clarity*

1 Introduction

The Kamoa Copper Deposit, located approximately 25 km west of Kolwezi in the Democratic Republic of Congo, is recognised as one of the largest high-grade copper deposits globally. It is part of the Central African Copperbelt, a region known for its significant copper mineralisation. With the recent completion of its Phase 3 expansion, the project has rapidly expanded its production capabilities, achieving a processing capacity of over 600,000 tonnes per annum of contained copper. This positions Kamoa Copper SA as one of the largest copper producers globally and the largest in Africa, reflecting its operational efficiency and growth potential. A map of the resource is included as Figure 1 showing both the Kakula and Kamoa deposits.

The treatment of copper sulphide ore is undertaken by flotation. Copper sulphide flotation is a critical process in the extraction of copper from ores, particularly from minerals such as chalcopyrite, bornite and chalcocite. The effectiveness of flotation depends on various factors including mineral composition, particle size, pH levels and the reagents used. As the copper sulphide ore is extracted from the mine via a primary crusher followed by a secondary crusher and high-pressure grinding rolls. The crushed ore is stockpiled in the atmospheric air before it goes to the grinding circuit, where it is reduced to a particle size of 53 microns before flotation.

Once floated, the rougher concentrate is then reground to 10 microns for the cleaner circuits.

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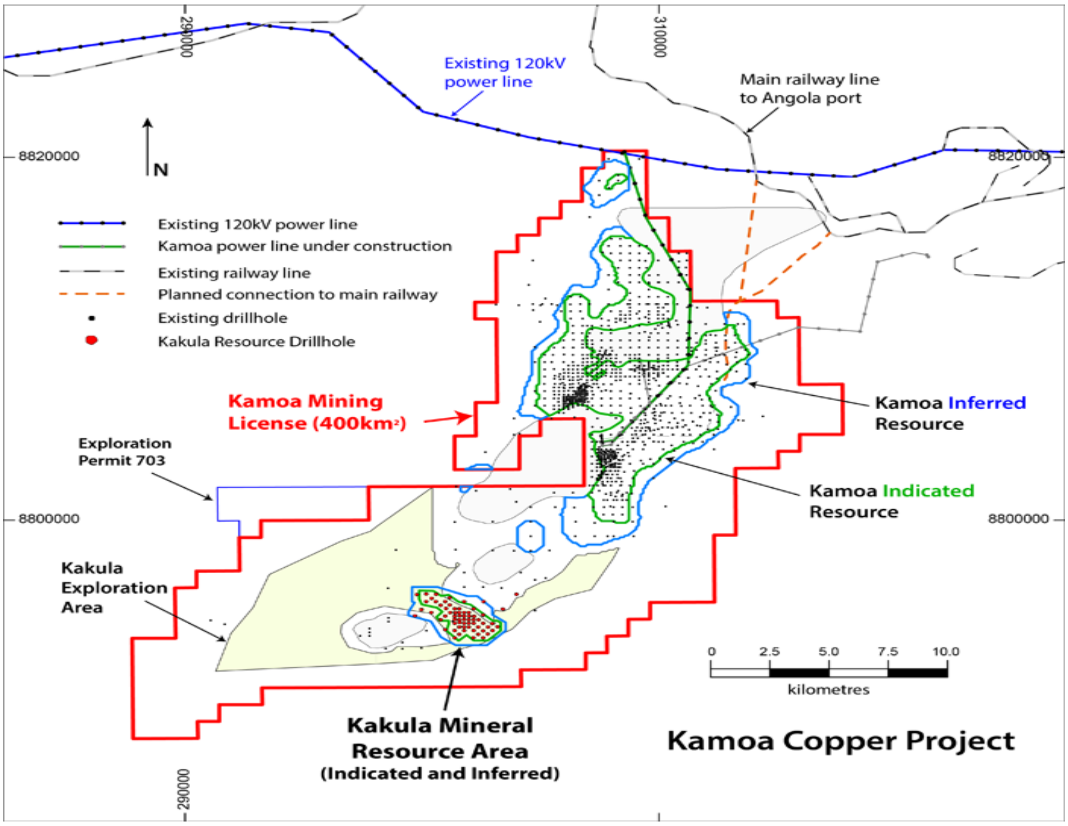

Figure 1 Kamoa Copper project

The plant has been built in stages, with Phase 1 being an initial flotation plant working on the Kakula deposit. Phase 2 was a parallel flotation line built alongside Phase 1, treating more of the Kakula feed. Phase 3 is an entirely new plant situated about 20 km distant and is for processing of the Kamoa ore.

When exposed to atmospheric air, copper sulphide ore undergoes partial oxidation and, once in contact with water, generates sulphuric acid. This reduces the pH and leads to the serious corrosion of pipes, motors and pumps, given that the plant was designed for treatment in a neutral environment. With Phase 3, the presence of sulphuric acid in the circuit reduced the pH of the solution from 7 to 4, impacting both the sulphide flotation efficiency and the solid-liquid separation during settling of tails and concentrates.

Whereas coagulant is the main key to controlling tailings thickener overflow clarity at Kakula Phases 1 and 2 due to the fine particles generated by regrinding, the fine particles formed at Kamoa Phase 3 do not affect the tailings thickener overflow clarity as much, thereby reducing coagulant consumption and allowing the plant to operate at with a much better overflow clarity, typically less than 20 nephelometric turbidity units (NTU) compared to the 50–100 NTU seen during Phases 1 and 2.

When compared to other copper flotation operations, the settling at Kamoa is characterised by the very fine grind and even finer regrinding. Typical particle size distributions (PSD) feeding the thickeners is shown in Table 1.

Table 1 Particle size distributions for Phases 1 and 3

			Phase 1	Phase 3
Feed rate	Tonnes per hour		665–682	850–870
Particle size distribution	% passing on (sieve µm)	Milling	84–86% (53 µm)	82–84% (53 µm)
		Regrind	50–55% (10 µm)	55–65% (10 µm)

2 Thickeners

The thickeners are conventional Outotec types fitted with vane feedwells. For Phases 1 and 2 they have a 38 m-diameter, with the first built to handle Phase 1 and a second identical unit added for Phase 2. Feeding is via a header tank for each thickener. The units are dedicated to each flotation line and cannot be cross-fed. Overflow from the thickener reports by gravity to the process water tank and forms the majority of the plant's process water. Tailings are pumped a short distance to the tailing's sump tank and from there to the tailings storage facility. For Phase 3 a single 48 m-unit was built.

On each flotation line there are separate sumps for the rougher circuit and cleaner circuit. The two streams are pumped over separately and only then combined in the header tank. From the header tank the slurry passes through the feedline into the feedwell. Coagulant addition is into the flotation sumps, and flocculant addition is to the header tank and feedwells. The Phase 1 Thickener is pictured in Figure 2.

Figure 2 Kakula Phase 1 thickener in operation

3 Sampling and test work

The test work was conducted with simple 1-litre static cylinder tests. Although more accurate information can be obtained from other methods such as a dynamic rig (a small laboratory-scale thickener), the static cylinder test offers the benefits of needing little equipment and being easy to run. Although most of the work was carried out by the flocculant supplier (SNF), this type of work is well within the scope of most metallurgists and laboratory personal. Figure 3 shows a typical batch testing using 7 cylinders together.

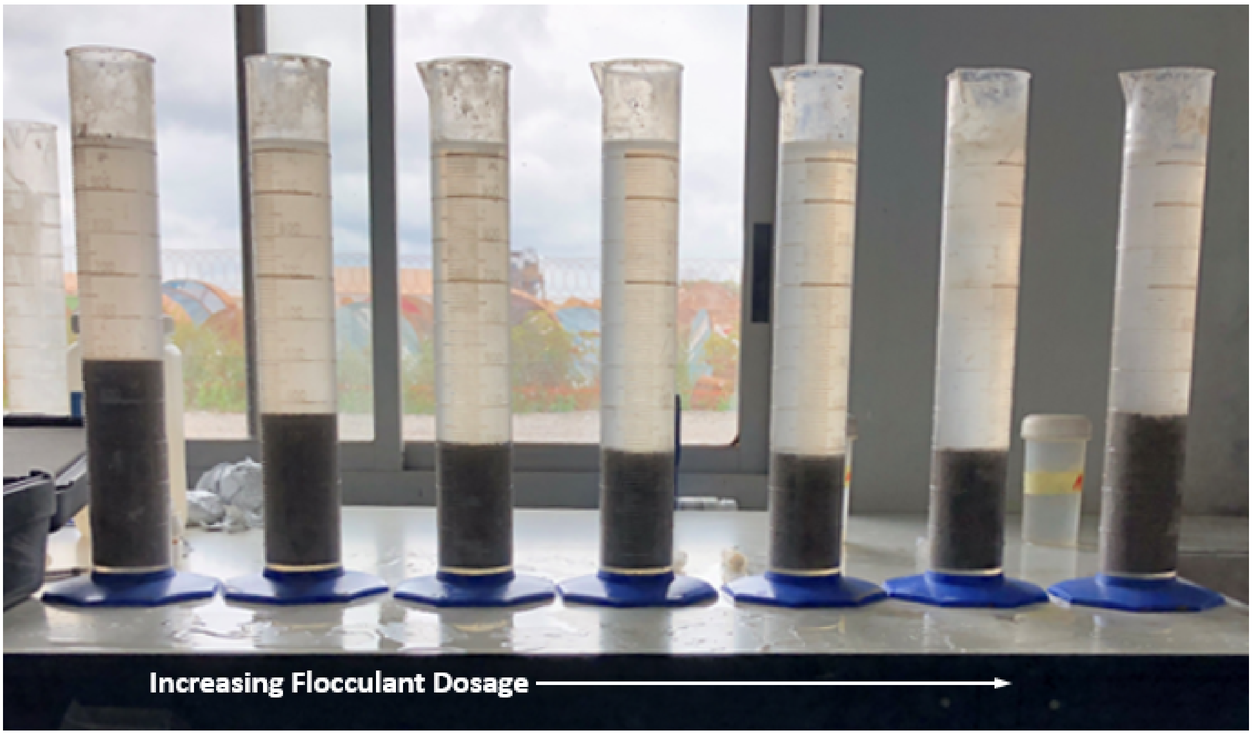

Figure 3 Example of cylinder-based test work. Note that in this case, flocculant dosage has little effect on supernatant clarity

One limitation is that these tests give good indications of performance with respect to settling rates and clarity but do not produce any meaningful information on the effects on underflow density or rheology.

In terms of using the test work as a troubleshooting tool, Figure 4, which was taken during the start-up, shows a disconnect between the actual overflow clarity achieved on the plant when compared to what the same reagents give in a laboratory-scale test.

Figure 4 Laboratory simulation (left cylinder) versus overflow from the operating thickener (right cylinder) using the same reagent dosages

When results like this are obtained it gives confidence that the underlying reagent choice is good and that the corrective action needs to focus on the addition conditions rather than swapping reagents. In most cases like this, with poor operational results but good laboratory results, we find that the right amount of flocculant/coagulant is not being added at the correct place.

4 Operational details and experience

4.1 Phase 1 Kakula overview and experience

The flow sheet for Phase 1 is shown in Figure 5 and is largely unchanged since commissioning.

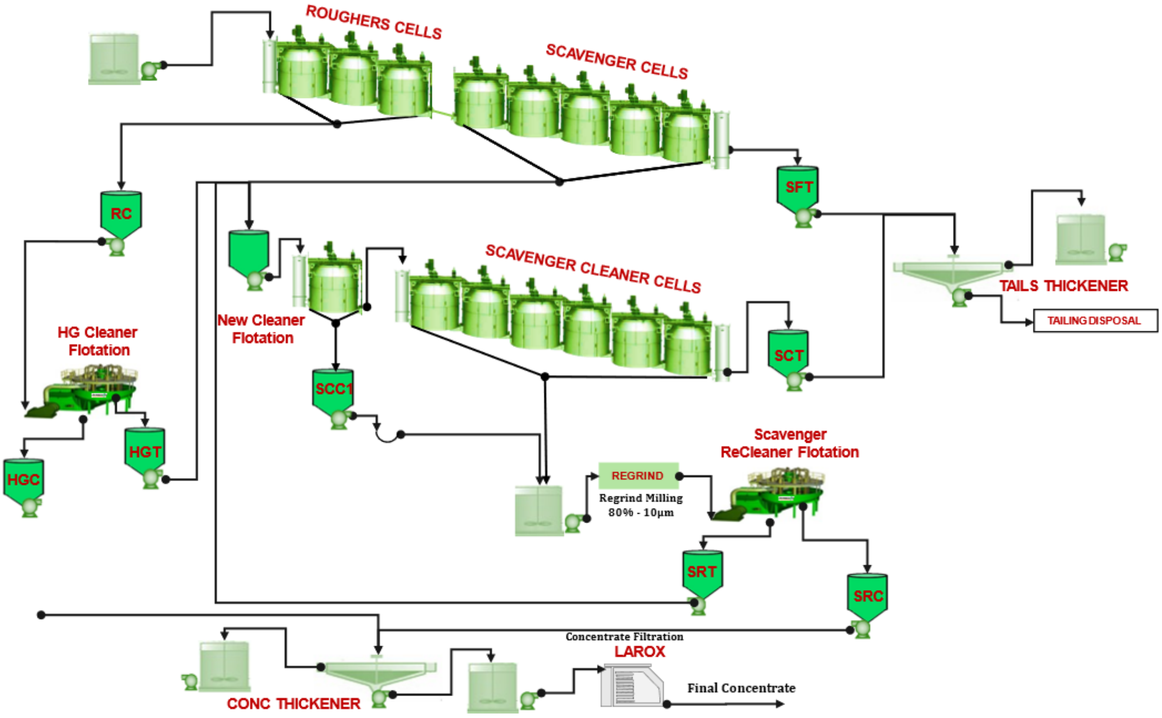

Figure 5 Kakula Flotation flow sheet Phase 1 and 2

Following the scoping test work, Phase 1 was constructed with both flocculant and coagulant handling systems to enable both chemistries to be used.

Coagulant addition was to the flotation tails sump tanks, and it was added before the transfer pumps to maximise the mixing. The flocculant and coagulant are added by a ring main system and each thickener receives the flocculant through a manual control valve and flow meter. The flocculant uses a split addition scheme with a portion added to the header tank and a portion added to the feedwell. Unfortunately control over the split is rudimentary and the lack of control valves makes accurate control more difficult. The system is relatively manual but is being steadily upgraded, with the target of automatic control on a g/t basis.

On start-up of Phase 1 there were several issues mainly related to the flocculant and coagulant handling and distribution system set-up. These included the coagulant being added to the rougher tailings only, meaning that the finer cleaner tails did not have a chance to contact the coagulant before the header tank. This had an impact on the overflow clarity that could be achieved. The process was changed to allow coagulant feed to both the rougher tailings and cleaner tailings sumps so that all the solids could then be contacted with coagulant before the flocculant addition. Figure 6 shows the difference in performance before and after the changes with the overflow clarity improving significantly.

Figure 6 Tailings thickener overflow before and after changes

Most other problems related to dosing control and ensuring that the right amount of coagulant and flocculant were being added.

Laboratory-scale testing was used extensively to confirm the product selection and set addition conditions and dosages. The work showed that the initial product choice from the scoping work (Flopam 910 SH flocculant and DB45VHM) worked well, and the focus quickly moved to getting the application of the products correct. Figure 7 shows how supernatant clarity is influenced by changing the coagulant dosage.

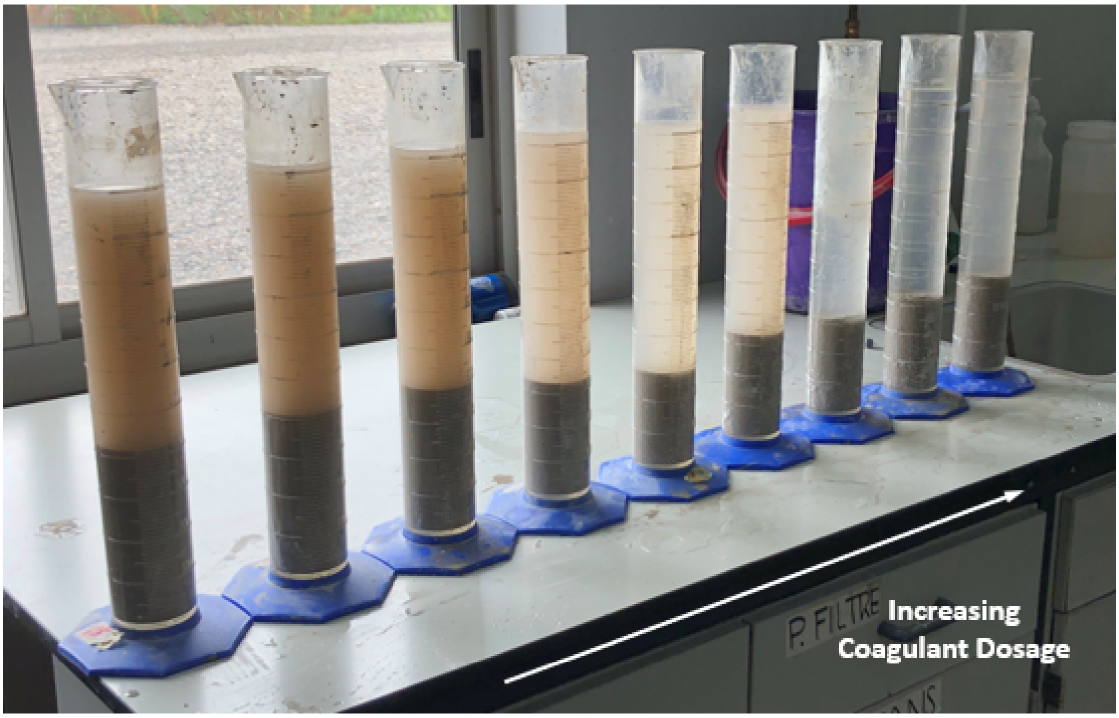

Figure 7 Effect of coagulant on Phase 1 settling, 0 g/t on left-hand side and 80 g/t on right-hand side

Changes since start-up have been an increase in the coagulant dosage and that the plant now tends to run a higher dosage of the coagulant than the flocculant. The initial assumption was for 50–60 g/t of flocculant and 30 g/t of coagulant; it is now running at around 45–50 g/t of flocculant and 50 g/t of coagulant, and overflow clarity has improved significantly.

The thickeners were set up with a ‘split dosage’ for the flocculant, with a portion of the dosage added to the header tank and the remainder added via three points in the feedwell. This scheme was simulated in the laboratory and is shown in Figure 8.

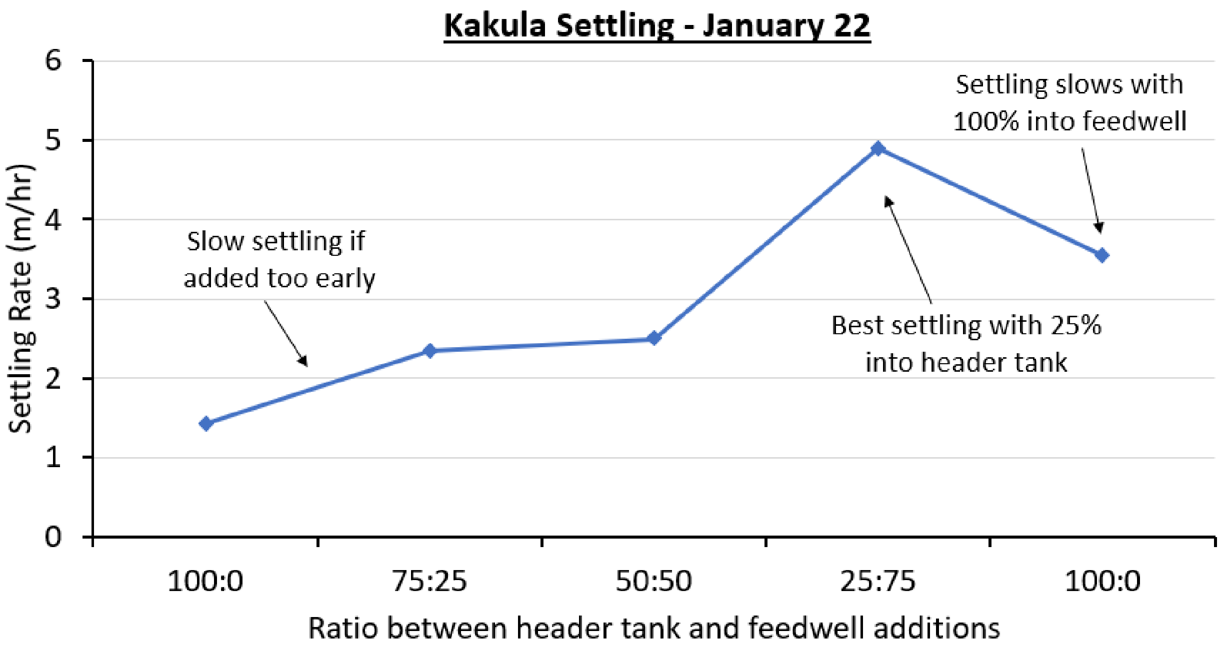

Figure 8 Effect of changing the flocculant addition points

The graph shows that how the flocculant is added to the thickeners has an impact on the performance. The difference in settling rates between the optimal addition scheme (25% header tank and 75% feedwell) and a sub-optimal scheme is larger than that for changing between products.

Typical performances of Phase 1 and Phase 3 are shown in Table 2.

Table 2 Comparison between Phases 1 and 3

			Phase 1	Phase 3
Feed tonnes per hour	Tph		2 × 665–680	850–870
PSD	% passing on (sieve)	Milling	84 –86 % (53μm)	82–84 % (53μm)
		Regrind mill	50–55 % (10μm)	55–65 % (10μm)
Feed solids	Tph	Concentrate	65–80	65–80
		Tails	590–610	780–800
Thickener size	m	Concentrate	2× Ø21 high rate	Ø28 high rate
		Tails	2× Ø38 high rate	Ø47 high rate
Underflow density	t/m^3	Concentrate	1.89–1.92	1.57–1.65
		Tails	1.46–1.5	1.39–1.45
Flocculant dosage	g/t	Concentrate	30	30–40
		Tails	45–50	40–45
Coagulant dosage	g/t	tails	50	20–25
Overflow clarity	NTU	Concentrate	250–300 NTU	150–250 NTU
		Tails	40–70 NTU	30–45 NTU

Phase 2 was a direct copy of Phase 1, built alongside and sharing the same coagulant and flocculant dosing systems. It was built with the modified coagulant addition feeding to both the rougher tailings sump and the cleaner tailings sump.

After several months of working systematically through the problems the dosages were brought under control and the clarity improved significantly.

4.2 Phase 3 Kamoa overview and experience

Phase 3 was built to handle the Kamoa plant output, again with a very fine PSD but a higher tonnage on a single flotation line. It uses a similar Outotec Vane thickener design, but the size was increased to a 48 m diameter to handle the increased tonnage.

Following the experience with Phase 1, both coagulant and flocculant were again specified as reagents, and a large clarifier was added to improve the process water quality and prevent recirculation of fine solids back into the flotation circuit. The Phase 3 flow sheet is shown as Figure 9.

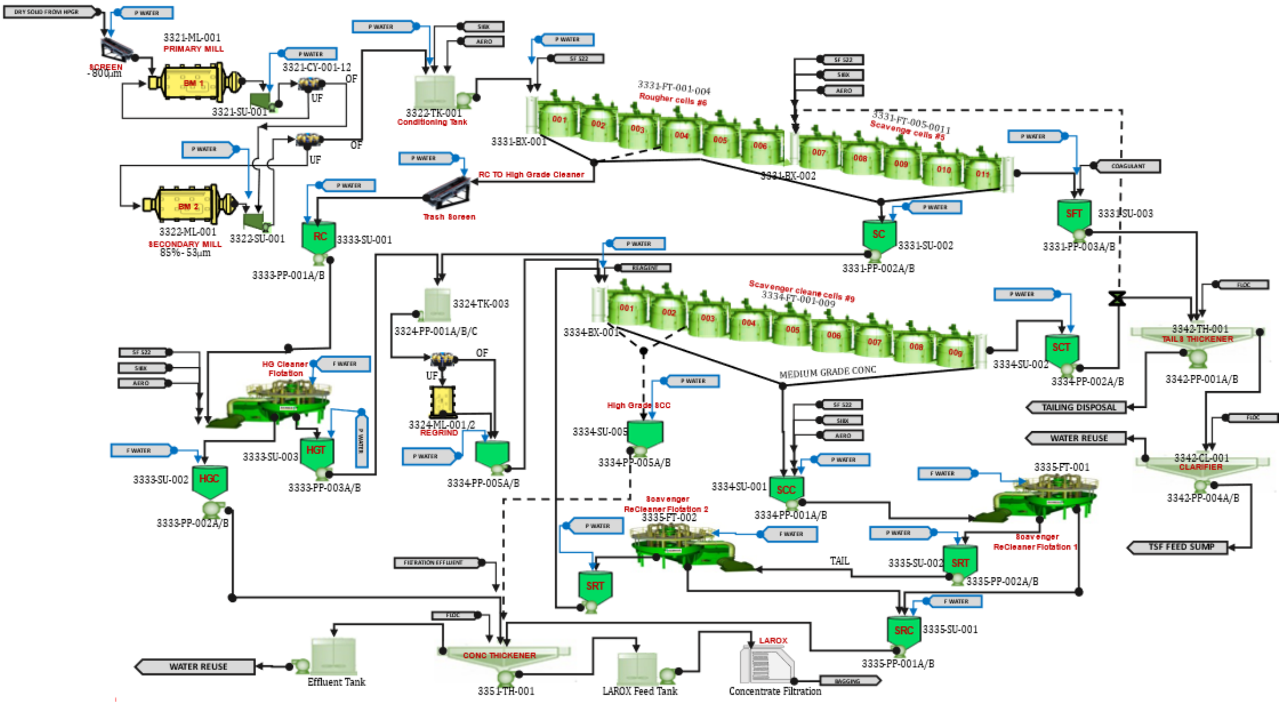

Figure 9 Phase 3 Kamoa flow sheet. Note the addition of a clarifier after the thickener

The start-up on Phase 3 was characterised by long periods of sliming of the thickener: a situation where the bed rises and eventually overflows the thickener into the launder and the process water tank. The sliming limited the tonnage being fed into the plant to around 650 tonnes per hour, with the target of more than 800 tonnes. Concerns were raised regarding the thickener performance and a further potential drop in feed once the regrind mills were started.

The problem was traced back to two causes:

- higher than expected density of the slurry feeding into the thickener
- a lack of flocculant being added into the thickener feedwell.

The level of feed solids reporting to the header tank was higher than anticipated and, with the low overall volumetric flow rates, the feedwell dilution channels struggled to pull-in enough water to effectively dilute this material and is shown as Figure 10.

Figure 10 Feedwell dilution channel in operation

Figure 11 shows a laboratory-scale simulation of changing the solids concentration.

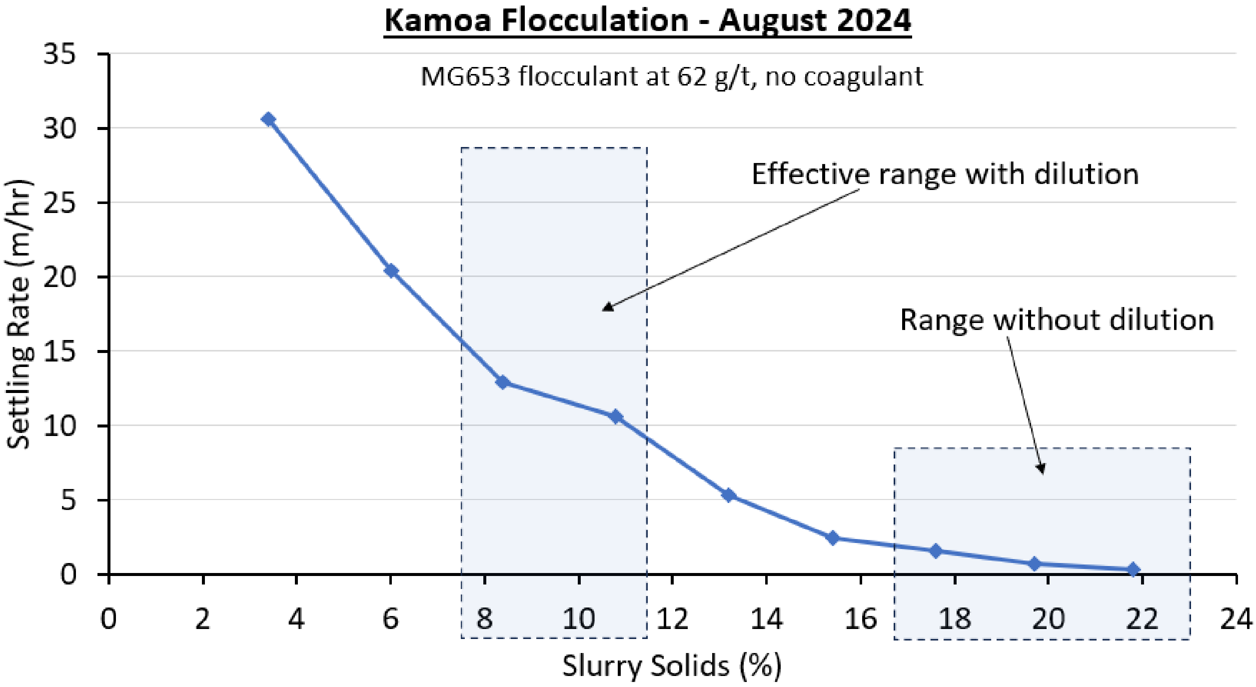

Figure 11 Effect of changing feed solids on settling rates

In the graph above we can see very slow settling rates from the high solids slurries (higher than 12%). The settling rates increase as the slurry is diluted and this is the principle behind high-rate thickening. Corrective action in this case was to increase the dilution of the slurry so a new high-capacity line bringing tailings return water directly into the header box was quickly installed. This meant that at any point the dilution of the slurry could be increased to help the settling and provide a valuable tool that most other thickeners do not have.

Between building Phases 1 and Phase 3, the flocculant distribution pipework design on top of the thickener was changed. Phase 1 utilised 100 mm diameter lines to take the diluted flocculant solutions out to the feedwell, but on Phase 3 only 50 mm diameter lines were installed, which limited the flow.

A simple laboratory-scale dosage curve was then carried out where we looked at how the settling rates changed with flocculant dosage and product type and is shown as Figure 12.

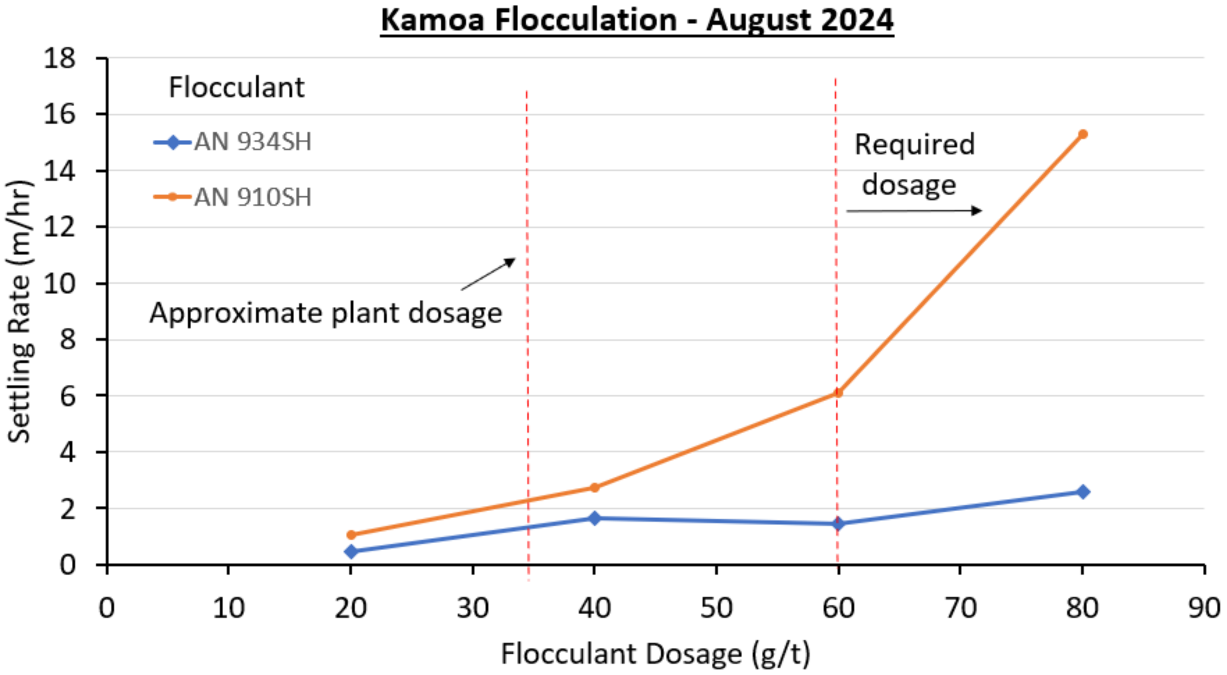

Figure 12 Cylinder test work comparing flocculant responses under different dosages

The graph shows several important trends:

- Settling rates drop with low dosages. At 35 g/t we are seeing around 2 m/hr settling, which is too low to successfully operate the thickener.
- With the AN 934SH flocculant (as specified and used on Phase 1), the settling rate stayed low regardless of dosage.
- A lower anionic charge type, AN 910SH, showed improved performance.

With a 50 mm diameter line feeding flocculant to the feedwell, the maximum amount of flocculant that could be added was only around 270 l/min giving a dosage of around 35 g/t. Due to the restrictions the majority of this solution ended up being added into the header tank, with less than a quarter reporting into the feedwell. Testing on the plant showed that if the valve taking flocculant to the header tank was closed, giving dosing to the feedwell only, overall flocculant flow dropped from 270 l/min to less than 100 l/min.

It was also found that the control valve installed for the tailing's thickener flocculant was only 25 mm diameter, further restricting the overall flocculant flow. The valve for the corresponding concentrate thickener was 50 mm diameter and these were eventually swapped over.

To quickly get more flocculant to the thickener a line was added from the flocculant make-down system sump pump to feed directly to the feedwell, and flocculant solution was dropped into the bund to be pumped directly. Although highly manual in operation, this worked well. Feed tonnage into the plant was increased and the regrind mills started with the thickener remaining stable. This confirmed that a lack of flocculant dosage was the problem.

Unlike in Phase 1, the product choice for Phase 3 was found to be problematic. During the first few weeks of operation, stockpiled material, which had oxidised and started to product an amount of acid mine drainage

(AMD) was being fed into the circuit. The pH in the circuit had dropped significantly, reaching as low as pH 4. Under these conditions the anionic-type flocculants (Flopam AN 934VHM and, to a degree, AN 910SH) lost their effectiveness. Laboratory-scale test work quickly showed a non-ionic type, MG653 (more commonly used for acid leach circuits), performed better and an amount was quickly procured. Figure 13 shows the settling rates for different types of flocculants.

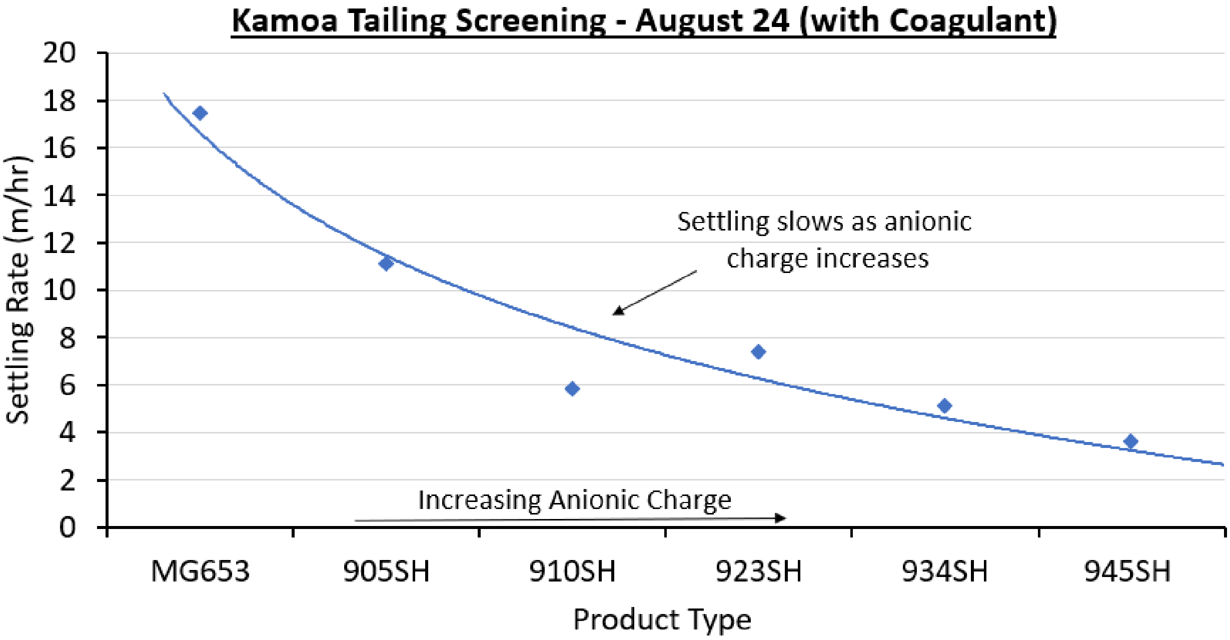

Figure 13 Tailings screening test work showing lower-charge products are favoured because of the lower-than-expected pH levels

The low pH caused by the AMD started to give other problems in the process, with corrosion and leaks becoming an issue. To counteract this, lime was added into the milling stage and the circuit now runs closer to pH 8. This again has changed the charge requirement of the flocculant, and the choice has reverted to the anionic types. This is only a temporary situation and, once the stockpiled ore has been consumed and the circuits pH stabilised, the flocculant choice will be re-evaluated.

Interestingly, all the test work carried out on Phase 3 gave very good supernatant clarity without the addition of coagulant. If the flocculant dosage is increased by a small amount the coagulant can be removed entirely. This would represent a significant cost saving and a simplification of the process.

5 Learnings and conclusions

Designing, building and commissioning large mineral processing plants is a difficult task under any circumstances. From a flocculation and thickening perspective, this is made more difficult because the test work on which the design will be based is normally limited by the samples available for test work, the solid phase samples are normally available in reasonable quantities put actual process water does not yet exist. Crucially, this work inevitably uses different water to what the plant will eventually be running with, and process water (after much recirculation) is difficult to simulate. However, in the case of Phase 1 the original work was surprisingly close to the products and dosages eventually used. For Phase 3 we saw significant changes from the specified reagents and the product choice is still being refined as conditions change. The potential removal of the coagulant dosage represents a simplification of the system and significant cost savings. If the reverse was the case, and a coagulant was not specified but becomes needed, adding this would be both difficult and expensive. The main lessons from this are to quickly re-check the screening work once the plant is up and running using the available process water and to not assume that the original choices are still the best.

The issue with the low pH AMD-generating material highlights the need to be flexible and to change product if needed. In the case of Kamoa, this meant changing back as conditions were further modified.

Both plants suffered from teething problems during their start-ups, but the test work highlighted the underlying causes and corrective actions were taken. With Phase 3 the actions were taken much more quickly as confidence had been built in the troubleshooting process. Although most of the issues were relatively minor and related to the distribution/addition of flocculant and its control, these issues made the difference between the thickeners running successfully or the whole plant being constrained. With the application of flocculants, the small details do count. If the right product is selected and the correct amount added under the right conditions, the reagent inevitably works.

In terms of ongoing work, the focus is now on improving the control, and optimising the dosages, of both the flocculant and coagulant.

Acknowledgement

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