

On the Industrial Plant Performance & Operating Point Drifting Phenomenon

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Abstract— The effective performance of various processing plants is reliant on many contributing factors. Such factors include equipment availability and operating performance, operating practices, technical influence and the ability to control the process in a consistent and high performing manner. Operating objectives need to be established to allow for product quality and throughput targets to be achieved throughout. This paper elaborates on an assessment of various categories of plants to demonstrate the wide dominance of the theme and aims to assess the performance of specific industrial plants, namely, Coal Handling Preparation Plants and Bauxite Beneficiation Plants against these parameters. The assessment will focus on the process as well as the control schemes supporting these process objectives.

I. INTRODUCTION

The performance of a mineral processing plant, can be measured in many ways, whether it be production throughput, product quality within contract specifications or cost per tonne margins which may extend to process energy consumption (electrical or steam), equipment maintenance or failure or resourcing costs. A mineral processing plant, such as an coal handling processing plant or bauxite beneficiation plant, can be a complex process with multiple control schemes ranging from regulatory to advanced regulatory to; in some cases supervisory schemes, incorporating estimation and optimization techniques. Thus, it is important to establish and assess mineral processing plant's performance measures in both technical and business terms.

II. APPROACH

The assessment approach discussed in this paper will utilise historical data collection from the processing plants to provide the information upon which to base the analysis and conclusions of the assessment.

The assessment tasks cover a range of process control elements. The three assessment categories considered for this assessment are:

- Control loop performance – do the individual loops perform satisfactorily?
- Control scheme optimisation – are the unit schemes appropriate for processing requirements?
- Business drivers – where would revenue value be realised by improving process control performance?

Alarm management will not be discussed in this paper as the detail and extent will be significant, limiting the control scheme assessment discussion.

The paper will firstly detail the assessment and improvement recommendations of three Coal Handling Preparation Plants, detailed in Section III; followed by the assessments and recommendations of two Bauxite Beneficiation Plants, discussed in Section IV, finalized by concluding remarks, Section V.

III. COAL HANDLING PREPARATION PLANTS

The loops analysed at CHP Plant 1 consist mainly of those installed in the washery circuits. Thirty four (34) loops were analysed in total, comprising of sixteen (16) from CHP Plant 2 and eighteen (18) from CHP Plant 1. Specific criteria were applied to assess the performance of these individual control loops. Of the surveyed control loops in CHP Plant 1 and CHP Plant 2, the assessment found:

- **6% (2 out of 34) of loops were in manual or out of service.** With the exception of two CHP Plant 1 loops in manual mode due to faulty instrumentation, all other loops were found to be operating in automatic.
- **18% (6 out of 34) of loops are poorly tuned.** Comparison of error measures and autocorrelation results revealed high variability in dense-medium density control. There are specific control design and tuning techniques which can improve the performance of density control.
- **38% (13 out of 34) of loops are tuned with an incorrect philosophy.** Both CHP Plant 1 and CHP Plant 2, have an incorrect tuning philosophy applied

to the sump level control loops. It is recommended that the Sump level control, be tuned to absorb disturbances within the process, rather than to maintain a consistent level in the sump.

III.1 CHP PLANT 1 AND CHP PLANT 2: CONTROL LOOP TUNING PARAMETER ASSESSMENT

The role of a sump within the process plant flow sheet is to collect varying feed flows to a central repository for transfer, to prevent air entrainment and importantly provide an absorption function within the process. The absorption function is to remove or mitigate process disturbances within the area of influence. The disturbances reporting to the sump are represented by surges in sump feed rate. In response to the disturbance the pump discharge rate or other manipulated variable must not propagate the disturbance to the corresponding process areas. Figure 1 illustrates that within CHP Plant 2 a dense medium sump level has insignificant variation of level, which results in significant controller output variation, this introduces potential process disturbances in corresponding process areas.

Wade [2] establishes that many liquid level loops are not critical and one can tolerate fluctuation, even offset in levels, if it smooth's out the flow to a downstream process unit. An alternative tuning approach to the PID level control called "averaging liquid level control", with the assumption that the allowable tolerable excursions above or below the set point are equal, the controller gain (K_c) is given by:

$$K_c = \frac{100}{2\Delta L_{max}} \quad (i)$$

Where ΔL_{max} is given by the maximum allowable deviation from set point, in percent of full scale level measurement, resulting from a step disturbance from the inflow. Additional recommendations within CHP Plant 1 extend to the following features:

- The majority of the control should be achieved with the proportional action of the controller as illustrated by Wade [2]. The integral action should be insignificant with respect to the proportional component, due to the integrating process characteristic of the sump level.
- Adopting the formula detailed in (i), ΔL_{max} will be in the order of 30(%) transposing the proportional gain K_c to a low value, to allow significant variation of level to absorb as much of the feed flow variations as possible.

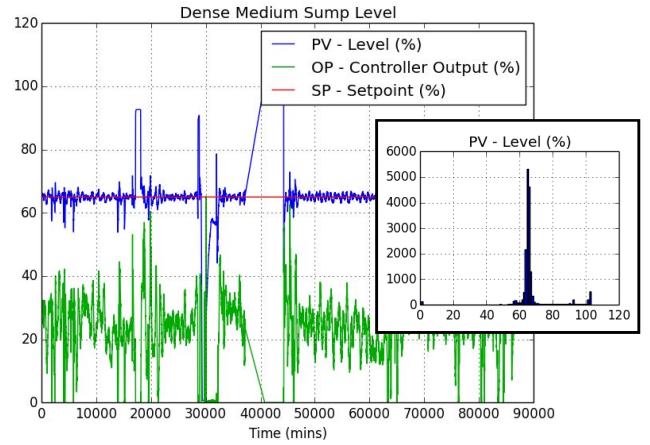


Figure 1: Secondary Dense Medium Sump level control

III.2 DENSE-MEDIUM DENSITY CONTROL IMPROVEMENT RECOMMENDATIONS

Medium density varies by up to 0.6 SG (peak-to-peak) due to poor control design or tuning. An illustration of this variability for one of the loops is shown in Figure 2. Considering the coal washability, Wills [3] illustrates that the variability in the Medium Density (SG) can affect the classification efficiency, which will directly impact coal recovery and ultimately revenue.

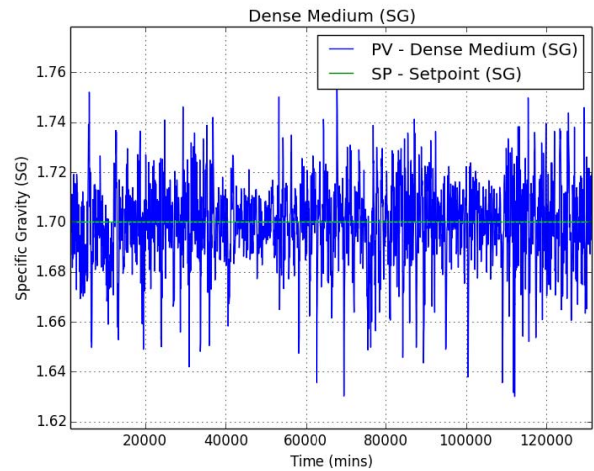


Figure 2: Variability of CHP Plant 1 Module 2 Secondary Dense Medium Cyclone medium-density

An analysis of the process interaction between the dense medium density and the dense medium sump level, illustrated an interaction between the two process variables. A cross-correlation indicated a correlation coefficient of -0.94. Illustrating that when the sump level increases, the negative effect occurs on the dense medium density, resulting in a decrease in specific gravity.

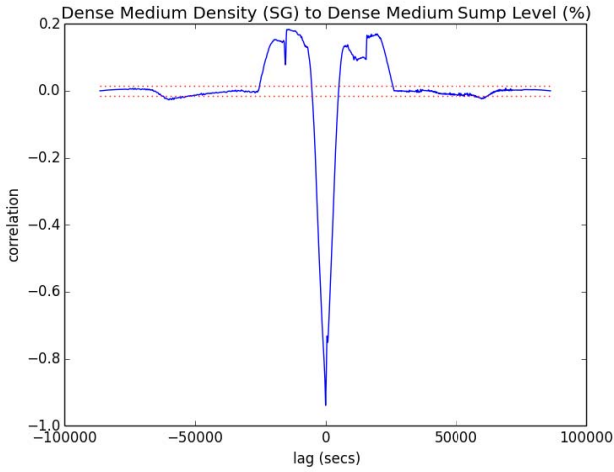


Figure 3: Cross-correlation of Dense Medium Density (SG) to Dense Medium Sump Level (%)

It was established that the Dense Medium Density and the Sump Level within the area of process influence, were controlled by an individual controller for each process variable. The interaction between loops can be reduced by the implementation of a control scheme that accommodates this characteristic, such as a decoupling approach. Figure 4 represents the block diagram for a typical decoupling two degree-of-freedom PID structure for an interacting tank system, where b_1 and b_2 are feedforward compensators.

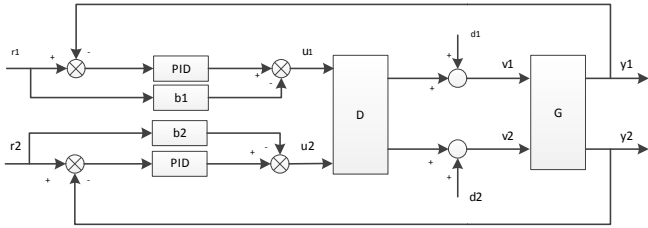


Figure 4: Two-input-two-output PID decoupling control system [4].

When considering the PI (PID) parameters, Tavakoli [5] suggests a gain margin greater than or equal to 3 and a phase margin of greater than or equal to 60° as the robustness constraints. Optimal PI tuning formulae resulting from the non-dimensional tuning method [5], capable of handling a first order plus deadtime, lag dominant and integrating process is illustrated in Eq. 2.

$$K_c K_p = \frac{T}{2\tau_d} + \frac{1}{14}, \quad \frac{T_i}{T} = \min\left(1 + \frac{1}{7} \frac{\tau_d}{T}, 9 \frac{\tau_d}{T}\right) \quad (\text{ii})$$

To resolve for D , in Figure 4, Wade [1] proposes an inverted decoupling approach. Note the signal direction through the decoupling elements, illustrated in Figure 5. Equations representing the decoupling circuit are given in Eq. 3.

$$m_1 = c_1 + D_{12}m_2; \quad m_2 = c_2 + D_{21}m_1 \quad (\text{iii})$$

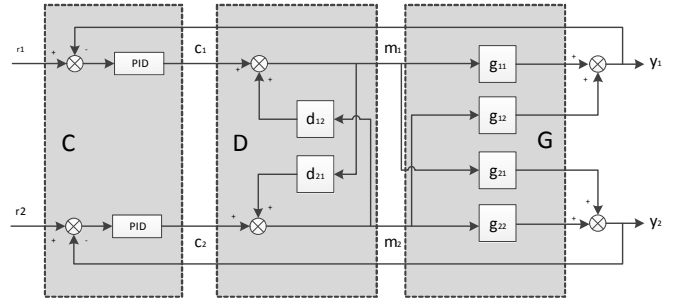


Figure 5: Inverted decoupling control system of a TITO process [7]

III.3 CHP PLANT 1: THICKENER UNDERFLOW CONTROL SCHEME PERFORMANCE ASSESSMENT

A thickener process concentrates solids in slurry generally for the purposes of: recovering water for use in the processing plant; and, reducing the volume of slurry that needs to be pumped/handled in the tailings (or waste fines) storage facility. In order to achieve the residence time for the desired amount of solids settling, the thickener vessel is large (15-40+m diameter). The solids-bearing underflow pumping rate is an important variable that can be controlled whilst the bulk of the overflow is via clarified overflow around the top edge of the vessel.

Operationally, the two aims are to: (1) avoid solids reporting to the clarified overflow stream; and (2) minimise (up to a limit) the volume of liquor, typically water, reporting to the underflow.

The main complexity in the thickener process is that the process of solids settling is complex for most applications. To assist with increasing the speed of solids settling, flocculant material can be added to the feed stream in order to assist aggregating solids particles.

The controller objectives can be summarised as:

- Ensure solids inventory does not build up excessively. The inventory can be measured / inferred via: bed level, bed pressure, underflow density, rake torque.
- Ensure that the optimal amount of flocculant is being added, both “too much” and “too little” can be detrimental to settling performance, in addition to the expense in the case of “too much.” The settling performance can be measured / inferred from settling time analysers, clarometers, hindered layer measurements, overflow turbidity probes.
- Ensure that equipment limits are being observed. In particular that rake torque is not excessive, nor underflow pumping under stress.
- Ensure other process objectives are met. For example, there is often a minimum underflow pumping rate/velocity (to avoid solids settling in long tailings lines.)

The tailings thickener underflow density control loop has low utilisation due to poor performance in automatic resulting in manual mode selection. The underflow density is low reducing the capacity to transport solids to the tailings dam. The thickener is adequately instrumented and, with a robust underflow control scheme, should be capable of regulating density to maximise solids transport and minimize the probability of tailings line blockage.

Plant throughput restrictions are incurred by the thickener performance, particularly during coal seams that exhibit ease of washability. Improved thickener underflow density control will relieve this constraint by managing the thickener constraints in 'real-time' while reducing the water loss to tails.

Tailings dam operating objectives support a higher underflow density which prompts stacking of the tails at a steeper incline angle in the dam.

III.4 CHP PLANT 1 AND CHP PLANT 2: RECLAIMER CONTROL IMPROVEMENT RECOMMENDATIONS

Analysis of the reclaimer present at CHP Plant 1 showed performance issues in the following:

- **Inconsistent reclaim-rate over the length of each Long-Travel (LT) run** over some periods. Figure 6 shows the Clean Coal reclaim rate for a 100% LT speed setting. The fact that the reclaimer can continuously sustain 100% speed suggests that deeper cuts should be set so that, for the same average reclaim rate, the speed is lower. This provides some headroom in LT speed which an automatic control scheme can use to try to stabilise the tonnage during each LT, thereby recouping lost capacity due to reduced tonnage at either end.

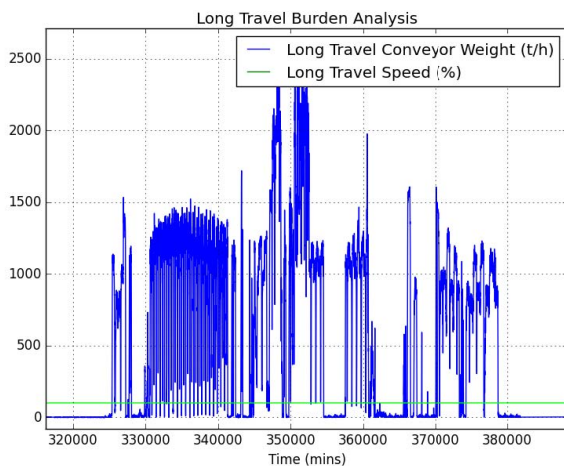


Figure 6: CHP Plant 1 Clean Coal reclaim-rate (tph)

- **Alternating average reclaim-rate depending on LT direction.** Figure 7 shows alternate LTs with different average rates, most likely due to velocity of the reclaimer relative to the moving conveyor that is

being discharged onto. This causes a different tonnage density, per metre, depending on the LT direction. This effect is well-known in other reclaimer types and control schemes can directly address this issue to maintain constant output burden.

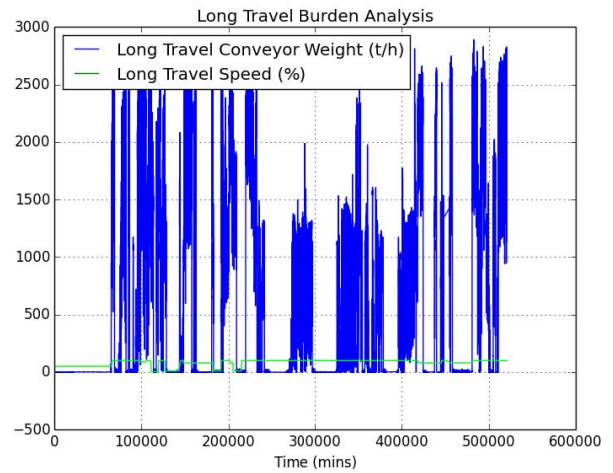


Figure 7: CHP Plant 1 Clean Coal alternating reclaim-rate (tph)

- It is recommended that the reclaimer control scheme, include a constraint control approach where the reclaim rate can be maximised without exceeding any of the constraints or operating limits of the machine. For example, the control scheme should incorporate a constraint for the scraper-chain drive power or current, to prevent drive over current events and control machine stress including scraper wear, triggering protection systems.

III.5 CHP PLANT 3: CONTROL LOOP PERFORMANCE ASSESSMENT

54 control loops were analysed for CHP Plant 3. The summary of the performance assessment of individual loops concludes:

- 72% of the loops were operating correctly, with 28% of the loops poorly tuned (15 out of 54).
- No control loops indicated as out-of-service.

In general the control loops are well utilised and function as intended.

IV. BAUXITE BENEFICIATION PLANTS

IV.1 BB PLANT 1 AND THE BB PLANT 2: CONTROL LOOP PERFORMANCE ASSESSMENT

The state of process control is good, within BB Plant 1 and BB Plant 2. Further opportunities to deliver value using process control still exist, many of which relate to the process control schemes rather than systems.

Some indicators of the state of process control are listed below:

- All of the observed localised control schemes (e.g. tank level, material handling) operate in automatic.
- 58% of surveyed control loops were operating correctly; 42% require attention and/or further investigation.

A higher number of opportunities exist for scheme improvement at BB Plant 2 compared with the simpler, highly automated, BB Plant 1.

BB Plant 1, 11 control loops were analysed and for the BB Plant 2 13 control loops were analysed. Results from this analysis were:

- BB Plant 1, 55% of the control loops were operating correctly.
- BB Plant 2, 62% of the control loops were operating correctly.
- The control loops are well-utilised in automatic and function as intended, however several would benefit from modifications to their loop equipment, configuration or tuning.

Specific opportunities, some of which require further investigation, are detailed in the following sections of this paper including recommended actions.

IV.2 BB PLANT 1: CONTROL LOOP IMPROVEMENT RECOMMENDATIONS

1. It was observed that at higher feedrate setpoints the apron feeder can saturate at 100% speed. This is a known problem with site personnel currently investigating solutions. It is recommended that a review of ore flow control loop tuning be undertaken if the flow characteristics of the apron feeder or transfer chutes are changed.
2. Shuttle chute position trim controller was at its limit of adjustment for a significant period (21%) of the time analysed. It is recommended that a calibration of the screen motor current calculation that determines screen loading be undertaken.
3. Process water supply pressure controller was saturated 67% of the time analysed. When saturated, supply pressure is below setpoint. Further investigation is recommended to determine if this is affecting supply volumes and what effect this could be having on plant performance.

IV.3 BB PLANT 2: CONTROL LOOP IMPROVEMENT RECOMMENDATIONS

1. The poor quality of slurry flow measurements is affecting both the cyclone feed controller and recycle flow control loops.
2. Classifier Water Addition performance observes include:
 - a. All classifier water addition controllers are subjected to a flow disturbance, generated when

the trommel tank water addition valve position is doubled to maintain trommel tank level. An example of this characteristic for the flow to Classifier 3 is illustrated in Figure 8.

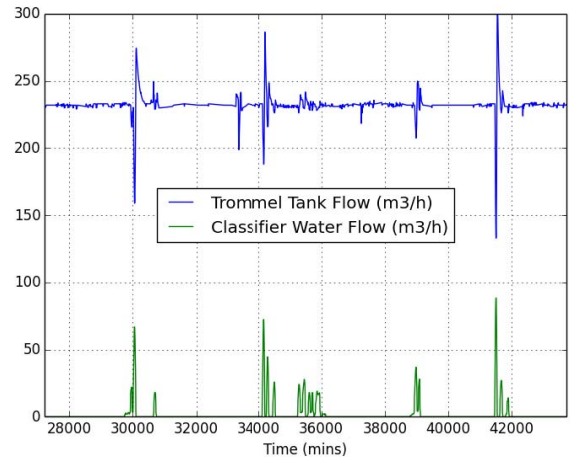


Figure 8: Disturbance caused by trommel tank valve

The duration of the disturbance is approximately 2 minutes. The significance of the potential impact, on classification performance is unknown. The existing trommel tank water addition configuration may be required to be altered to prevent this control-induced disturbance.

- b. Variability exists in performance between the six water addition controllers. Currently all controllers have identical tuning parameters. It is recommended that a loop tuning effort be undertaken to account for differences in installed valve performance and to accommodate the location along the header.
- c. Screen underflow tank two, level controller is over-tuned. It is recommended that it is retuned to realise the surge tank capability. Underflow tank one level control performance is significantly higher.
- d. Tailings tank level control is oscillatory. It is recommended that a review of loop tuning or cascading the level to a flow controller.

IV.4 BB PLANT 2: CLASSIFIER WEIGHT CONTROL

Although the existing classifier mass control operates well and illustrates stability, the actual mass (in %) is intentionally allowed to range over 65-68.5% as the plant feed rate changes, illustrated in Figure 9. Given that the water-only weight of the classifier is ~40%, this range represents about 14% in solids mass. The Classifier is a critical separation process, and given underflow problems with the discharge valves, it is preferable that the classifiers run as stably as possible at a specified setpoint.

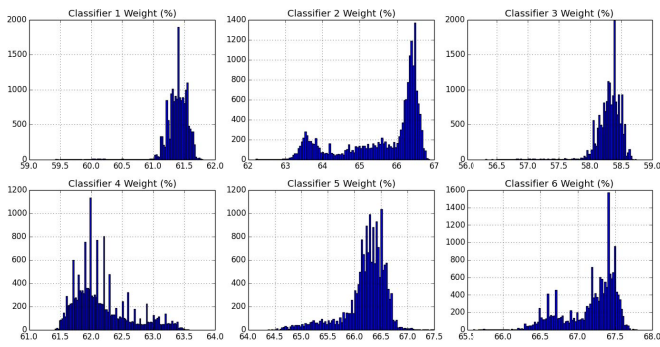


Figure 9: Classifier Weight Distribution(s)

Operation at a single setpoint could be achieved by:

- replacing the existing mass operating ranges (which selects 0, 1, 2 or 4 valves to be operating) with a PI controller; and,
- adding the option of 3 valves-open to the existing valve control logic.

Another advantage of restructuring to a PI controller format will be the ability to have a measure (i.e. the controller output) to allow comparison between the discharge valve loading on each classifier as a potential measure of imbalance.

Once the variability of the process variable is reduced, the question then becomes; what is the optimal operating setpoint? A lower setpoint, with the corresponding reduced solids residence time, may aid to prevent the reported “rat-holing” behaviour, at the expense of a slightly lower density discharge. However, this is mostly a process performance issue which is beyond the scope of this paper.

V. CONCLUDING REMARKS

In summary, one might say, what can process control do to improve or restore the performance of a mineral processing plant? Well, to firstly answer this question, one must understand the condition of the control schemes that are present within the process. This condition recognition aims to benchmark the control schemes. Once a base line is established an improvement program can commence with focus areas determined by the Control Scheme Performance Assessment results.

The specific improvements in control schemes (e.g. reclaimer, thickener control) and control loop performance (e.g. poor tuning) would have been identified by continuous performance monitoring. Continuous performance monitoring focuses on:

- Control loop performance,
- Control scheme performance.

This approach highlights performance deficiencies or deviations from optimum levels, thus enabling these opportunities to be pursued based on business benefit. By consistently tracking actual performance against set targets or measures, process control performance can be sustained and continually improved on an ongoing basis. It institutes:

- A continuous improvement focus driven by performance data analysis, identification and application of effective control schemes and techniques to optimise performance,
- A systematic, condition-based approach to process control maintenance, using a standardised methodology and tools to detect and diagnose performance deviations which otherwise get overlooked.

However, the effort required for continuous performance monitoring and supporting its systems on-site may be a significant burden for existing site staff if that task is to be performed manually. Thus, research into a methodology for continuous monitoring, assessment and adjustment of the plant operating points in real time will enable optimal technical performance, at a plant wide level, and thus, enhance the plant business objectives. This research is currently outside the scope of this paper, although future work will yield publications in this area.

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