PID and Predictive PID Control Design for Crushing And Screening

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Abstract: Crushing and screening processes are often regulated using relay-based on/off controls or, when controlled by PID controllers, the control design is limited to a single crusher and its proximity only. The design does not necessarily consider process interactions, load disturbances and, in a wider perspective, does not treat a crushing and screening plant as an entity from feeding to product piles. To allow usage of conventional PID controllers, effort is placed on designing controls that decouple interactions and provide a good load disturbance rejection. The paper also introduces a new controller type, Predictive PID, and justifies its usage for regulating dead-time dominant integrating crushing processes.

Keywords: PID control, predictive, control design, crushing, screening.

1. INTRODUCTION

Crushing and screening processes conventionally have only simple on/off controls for starting and shutting down equipment such as conveyors, feeders and crushers. Even if there were variable speed driven actuators for conveyors and feeders, they are often driven at a pre-defined fixed value and changed rather infrequently. Tradionally, at its best, there are only a few automatic PID (Proportional-Integral-Derivative) controls allowing smooth and disturbance compensating regulation of process variables. However, these controls are limited to operate only within a particular machine unit, such as a crusher, not allowing co-ordination and control of consequtive machine units. The lack of plant-wide process controls in crushing and screening processes prevents from utilising the whole production capacity potential of the machinery.

In the past, a lot of effort has been put on modelling single crushers which, obviously, serves a solid base for designing process controls. Crusher modelling work propably publicly initiated by Whiten (1972) and (1984) has been continued e.g by Evertson (2000), Johansson (2009) and Itävuo (2011) and (2013). There has been some publications on expanding the process control idea out of the crusher itself such as Sbarbaro (2005) and Itävuo (2012). Yet, they have limited their research on a sub-process containing a single crusher with screens and conveyors. A typical crushing plant, however, contains several crushing and screening phases which should be always considered simultaneously on a control design board.

Designing PID and Predictive PID controllers for a whole crushing plant does not require models for processes to be controlled. However, interactions and non-linearities of process variables need to be considered and, consequently, design measures to decouple them need to be taken. However, when considering tuning of PID and Predictive PID controllers, process models are significantly valuable. The process models can be used for tuning controllers and for verifying their performance and robustness against modelling uncertainties and load disturbances. Controller design even benefits from simple low-frequency models that can be created through identification tests such as presented by Airikka (2012a). A nice collection of several transient-based identification tests are given by Åström and Hägglund (1995).

In one of his break-through papers, Hägglund (1996) introduced a Predictive PI (PPI) controller for compensating dead times in delay-dominant processes. The elegant idea is to replace a derivative part of a PID controller by a predictive part which is clearly better for predicting future than derivation itself. The other benefits of the PPI controller come through both its simple parametrisation compared to a Smith predictor and its applicability to integrating processes without any modifications. In this paper, the usage of derivation part together with a PPI controller is justified through performance improvement achievements in real implementations for crushing and screening processes. The derivation term does not work for compensating dead time but it works, as traditionally, for reacting on changes in the controlled process output helping the control loop act better than without. And, when applied to integrating and dead-time dominant processes, the PPID controller gives a bigger stabilisation space for the controller tuning parameters allowing tighter tuning.

The target for process control design is to provide with an automation system enabling efficient user operation with high and robust control performance for guaranteing a stable process with increased process through-put, product quality and uptime. To achieve that, several aspects must be considered such as selection of appropriate control strategy and controller types, design of appropriate sensors and manipulatable actuators that can be operated over their whole range. Also, the paper treats a real industrial case where a crushing and screening process was automated in plant-wide –wise to improve plant operations, capacity and efficiency.

2. PPI CONTROLLER

The PID controller can be given as

$$u(t) = k_p\left(e(t) + \frac{1}{t_i}\int_0^t e(\tau)d\tau - t_d\frac{dy(t)}{dt}\right) \tag{1}$$

where controller output $u(t)$, control error $e(t)$ between setpoint $r(t)$ and measurement $y(t)$ are all signals with respect to time t and PID controller tuning parameters are proportional gain k_p, integral time t_i and derivative time t_d.

The PID controller has three different terms with different tasks. The expressions past (integral control), present (proportional) and future (derivative) are often used when addressing differencies of the terms. The derivative term may have different implementation variants but typically it is executed as given by (1) but having a low-pass filtered measurement instead of a raw measurement $y(t)$. The reason for this is to avoid amplifying measurement noise in the controller output. The filter time constant is often an additional fourth tuning parameter for the PID controller.

The Predictive PI controller by Hägglund is given as

$$u(t) = k_p\left(e(t) + \frac{1}{t_i}\int_0^t e(\tau)d\tau\right) - k_{pred}\int_{t-L}^t u(\tau)d\tau \tag{2}$$

where k_{pred} is a predictive part tuning parameter for having an integrated controller output for a limited time horizon as an input. To implement a PPI controller, a good estimate of process time delay L is essential for successful control performance.

Hägglund has justified the PPI control in various ways. First, the PID controller's derivative term is not capable of predicting process output $y(t)$ properly due to non-linearity of the dead time. Second, the PPI controller is a simplification of a Smith predictor that was originally created to deal with processes with long dead times. Actually, under some assumptions, the PPI controller is an exact match to a Smith predictor designed for a FOPDT (First Order Plus Dead Time) process. The beauty of the PPI controller is that it has less tuning parameters than the Smith predictor.

Later, Normey-Rico modified the PPI controller by introducing a low-pass filter for the measurement to improve robustness against measurement noise. However, this does not change the core of the PPI controller as given by (2). It has been shown by Airikka (2013) that the PPI controller increases a stability region of admissible controller parameters for the controller's proportional and integral parts. The larger stability region allows tighter controller tuning resulting in a better control performance if only sufficient robustness is guaranteed through careful control design.

3. PREDICTIVE PID CONTROLLER

The Predictive PID (PPID) controller combines both PID and PPI controller resulting in a controller as below

$$u(t) = k_p\left(e(t) + \frac{1}{t_i}\int_0^t e(\tau)d\tau - t_d\frac{dy(t)}{dt}\right) - k_{pred}\int_{t-L}^t u(\tau)d\tau \tag{3}$$

The proportional controller reacts on an amplitude of the controller error and the integral controller reacts on the accumulated control error by driving the control error eventually to zero for a stable closed-loop system. However, neither of them reacts on the velocity of process output. The derivative controller is criticised by its insufficient capability to compensate dead times. Yet, it still qualifies as a controller which reacts on process output changes $dy(t)/dt$.

It is known that the stability region can be expanded by introducing the derivative part's phase lead resulting in higher phase cross-over and bandwidth frequencies. The systems that typically require phase lead are integrating systems with dead times. Now, by combining the derivative and the predictive controllers, the frequencies can be increased even further. Now, the derivative part would take care of the integrating nature of the system whereas the predictive part would deal with the dead time compensation. Later, it will be shown that this is pretty much what takes place when having a PPID controller for IPDT systems.

The obvious drawback of the PPID controller is its complexity. The PID controller has four tuning parameters (k_p, t_i, t_d, t_{df}) and the PPI controller only three (k_p, t_i, L) assuming that the prediction parameter k_{pred} is given as a function of integral time t_i as proposed by Hägglund (1996). When considering the prediction parameter as an independent tuning parameter, a number of the tuning parameters is also four for the PPI controller. However, the PPID controller has at least five tuning parameters (k_p, t_i, t_d, t_{df}, L).

3.1 First-Order Plus Dead Time (FOPDT) systems

Consider a FOPDT (First-Order Plus Dead Time) system characterised by its transfer function

$$g(s) = \frac{k}{Ts+1}e^{-Ls}\text{, for } k \neq 0, T > 0 \tag{4}$$

Figures 1-4 illustrate controller parameter stability regions for a FOPDT system having $k = 1$, $T = 1$ and $L = 5$ for different controllers (PI, PPI, PID and PPID). The system can obviously be considered as dead-time dominating for as $L/(L+T) \approx 0.83$. In each plot, there is an integral gain k_i (k_p/t_i) plotted against a proportional gain k_p (horizontal). The controller parameter values inside the closed region stabilize the system when there is no model mismatch.

Figure 1 illustrates PID controller stability regions (solid) versus a PI controller stability region (dotted). It shows that by introducing a PID controller does not actually increase the total stability region area for a FOPDT system but rather changes it shape. The stability region shape gets narrower for larger derivative times (large α). The largest admissible proportional gain remains somewhat of unchanged for any PID controller tuning but the largest integral gain increases proportionally to α and to derivative time t_d. Obviously, it is difficult to maintain stability if even small changes in controller gains may bring the closed loop system unstable.

Figure 2 plots Predictive PI controller stability regions (solid) against a PI controller stability region (dotted). It shows how the stability region increases proportionally to prediction gain k_{pred} allowing tighter tunings for the PPI controller than that of the PI controller. Yet, at some point, the stability region starts to shrink (for k_{pred} >> 1/T).

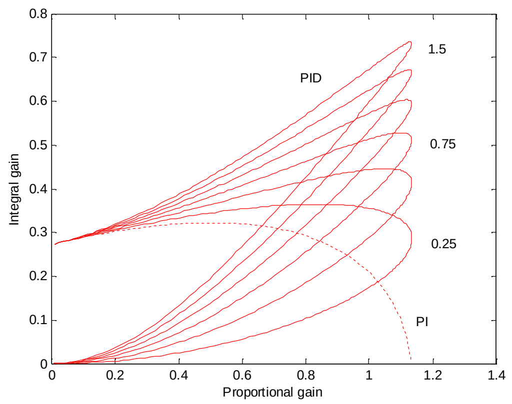

Figure 1. Stability regions of PI (dotted) and PID controller (solid) for $t_d = \alpha\, t_i$ with α = 0.25, 0.5, 0.75, 1, 1.25, 1.5.

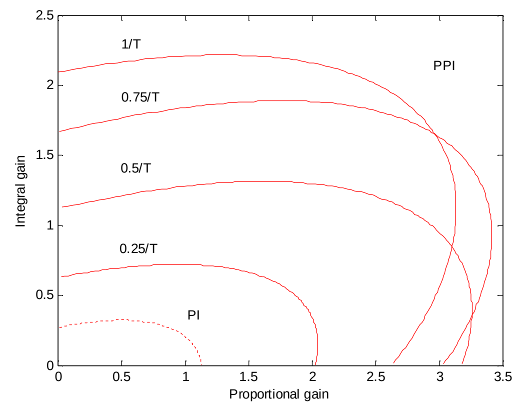

Figure 2. Stability regions of PI controller (dotted) and PPI controller (solid) for $k_{pred} = \lambda/T$ with λ = 0.25, 0.5, 0.75, 1.

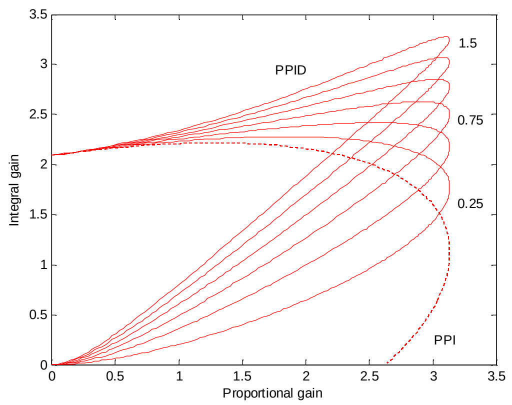

Figure 3. Stability regions of PPI controller (dotted) for k_{pred} = $1/T$ and PPID controller (solid) for $t_d = \alpha\, t_i$ with α = 0.25, 0.5, 0.75, 1, 1.25, 1.5 and $k_{pred} = 1/T$.

Figure 3 shows Predictive PID controller stability regions (solid) against a PPI controller stability region (dotted). Actually, the outcome is rather similar to that seen when comparing PI and PID controllers: the stability region basically changes its shape for α having the constant largest proportional gain but allowing slightly bigger largest integral gain. Figure 4 collects all the controller types (PI dotted, PID dashed, PPI dash-dotted, PPID solid) by showing their stability regions for $k_{pred} = 1/T$ and $t_d = 0.25\, t_i$. The predictive controller expands the stability region drastically but the impact of introducing the derivative controller to PI or PPI controller is rather small.

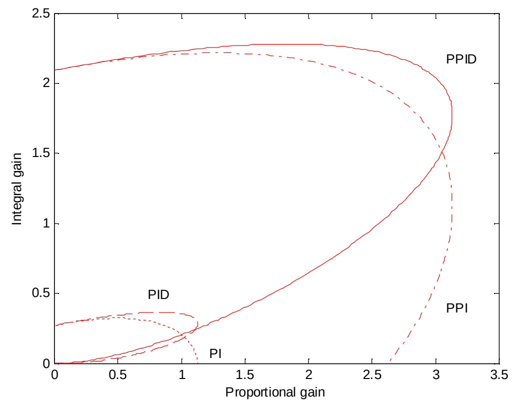

Figure 4. Stability region comparison between PI, PID, PPI and PPID controllers for $k_{pred} = 1/T$ and $t_d = 0.25\, t_i$.

Figure 5 shows setpoint and load disturbance responses for PI, PID, PPI and PPID controllers when applied to a FOPDT system. Basically, there is no improvement when introducing the derivative controller (PI→PID or PPI→PPID) but a clear improvement can be seen when introducing the prediction part (PI→PPI or PID→PPID).

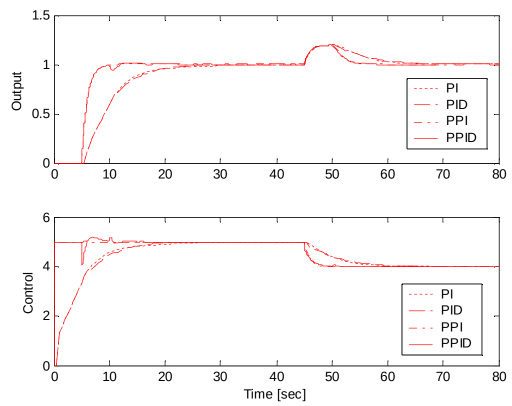

Figure 5. *Upper*: setpoint and load disturbance responses for PI (solid), PID (dotted), PPI (dashed) and PPID (bold) controller. *Lower*: control signals.

3.2 Integrating Plus Dead Time (IPDT) systems

Consider an IPDT (Integrating Plus Dead Time) system characterised by its transfer function

$$g(s) = \frac{a}{Ls} e^{-Ls}, \text{ for } L > 0, a \neq 0 \quad (5)$$

Figures 6-9 illustrate controller parameter stability regions for a IPDT system having $a = 5$ and $L = 5$ for different controllers. In each plot, there is an integral gain k_i (k_p/t_i) plotted against a proportional gain k_p (horizontal). The controller parameter values inside the closed region stabilize the system when there is no model mismatch.

Figure 6 illustrates PID controller stability regions (solid) versus a PI controller stability region (dotted). Again, introducing a PID controller shapes the stability region by making it narrower for large derivative times t_d or α. However, there are two differences compared to a similar case for a FOPDT system. First, the largest admissible proportional gain is increased slightly by a PID controller. Second, the stability region is much narrower for large t_d.

Figure 7 plots predictive PI controller stability regions (solid) against a PI controller stability region (dotted). It shows how the stability region increases proportionally to the prediction gain k_{pred} allowing, once again, larger controller gains and consequently faster closed loop. Controversially to a FOPDT system, the stability region dos not start to shrink for any value of $k_{pred} > 1/L$.

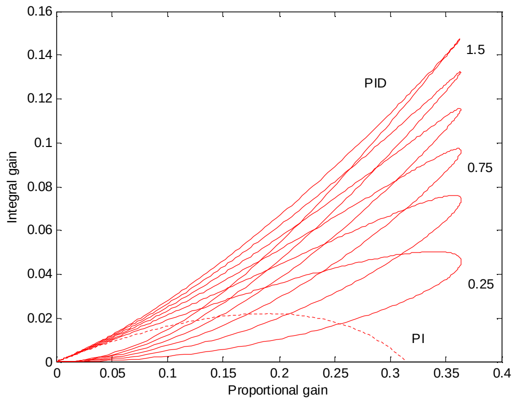

Figure 6. Stability regions of PI controller (dotted) and PID controller (solid) for $t_d = \alpha\, t_i$ with $\alpha = 0.25, 0.5, 0.75, 1, 1.25, 1.5$.

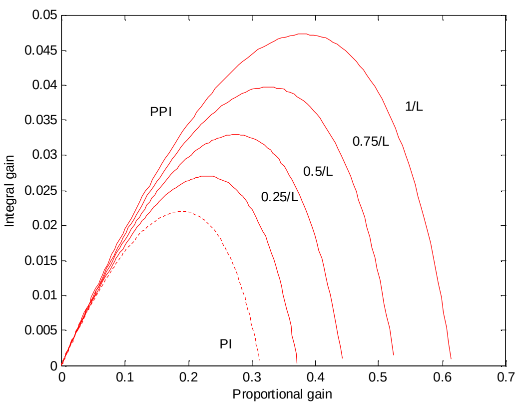

Figure 7. Stability regions of PI controller (dotted) and PPI controller (solid) for $k_{pred} = \lambda/L$ with $\lambda = 0.25, 0.5, 0.75, 1$.

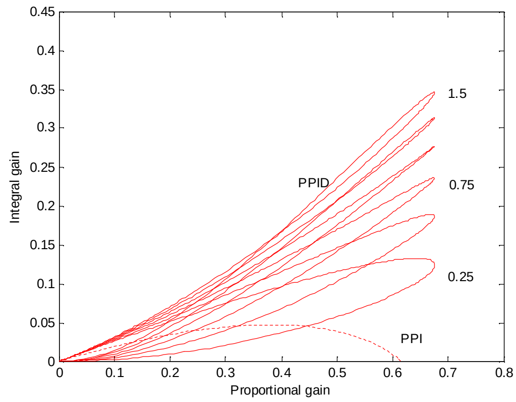

Figure 8. Stability regions of PPI controller (dotted) for $k_{pred} = 1/L$ and PPID controller (solid) for $t_d = \alpha\, t_i$ with $\alpha = 0.25, 0.5, 0.75, 1, 1.25, 1.5$ and $k_{pred} = 1/L$.

Figure 8 shows Predictive PID controller stability regions (solid) against a PPI controller stability region (dotted). The stability region grows quickly providing with higher integral gains for large α but having the constant largest proportional gain. Figure 9 collects all the controller types (PI dotted, PID dashed, PPI dash-dotted, PPID solid) by showing their stability regions for $k_{pred} = 1/L$ and $t_d = 0.25\ t_i$. Both predictive and derivative controller expand the stability region drastically but the shape gets trickier as soon as the derivative term is introduced.

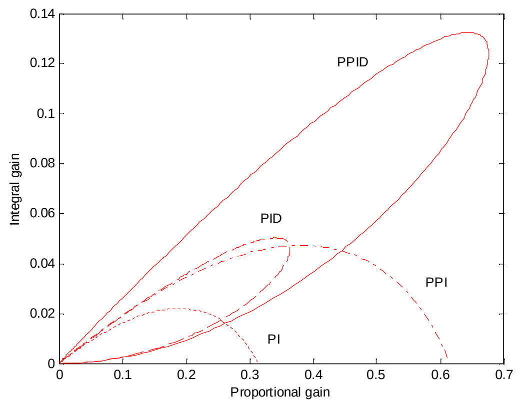

Figure 9. Stability region comparison between PI, PID, PPI and PPID controllers for $k_{pred} = 1/L$ and $t_d = 0.25\ t_i$.

Figure 10 shows output responses and control signals for a step setpoint and a load disturbance change for a IPDT process when being regulated PI, PID, PPI and PPID controllers. Similarly, figure 11 illustrates the same responses for an IPDT process. Both derivative part and prediction part separately improve setpoint following and disturbance attenuation contributing most when working together as an PPID controller.

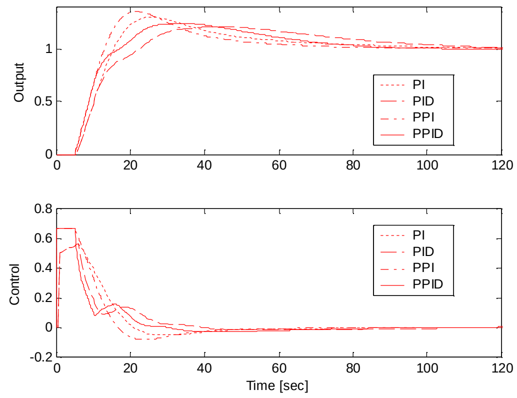

Figure 10. *Upper*: setpoint responses for PI (solid), PID (dotted), PPI (dashed) and PPID (bold) controller. *Lower*: control signals.

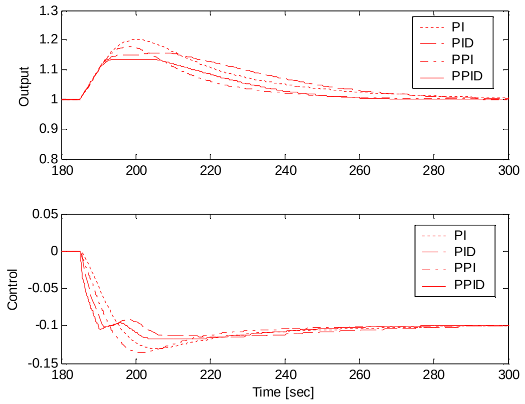

Figure 11. *Upper*: load disturbance responses for PI (solid), PID (dotted), PPI (dashed) and PPID (bold) controller. *Lower*: control signals.

4. CRUSHING AND SCREENING PROCESSES

Crushing and screening processes are dominant processes in aggregate production and play a vital role in minerals processing as well. The fundamental purpose of crushing and screening is to crush feeding material (rocks for aggregate production) into smaller particles and, finally, after consecutive crushing and screening phases into aggregate fractions having different targeted particle sizes in mm. Screening is essential for separating particles of different sizes. Oversized non-crushed material is re-circulated back to a previous crusher and undersized material can continue to the next crusher without crushing. For transporting material, there are feeders and conveyors that link crushers.

For measuring rock flows, ultrasonic sensors and belt scales can be introduced. Based on the both conveyor and feeder flows and, also, feed chute and crusher cavity level measurements, feeding actuators can be manipulated in closed loop. The processes can be mostly characterised by low-frequency FOPDT and IPDT system models with sufficient accuracy for a good control performance. Due to relatively long dead times caused by long conveyors and residence times in feed chutes and crusher cavities, dead time compensation needs to be considered in process control design. Compensation based on prediction is a good asset for a PI or a PID control loop that otherwise is capable of tackling the controlled process.

5. INDUSTRIAL USE CASE OF PPID CONTROL

An industrial use case of a PPI control is on a three-stage mobile crushing plant for aggregate production. There are three track-based Lokotrack crushing units each of which equipped with a machine control unit having monitoring and controls over the particular crushing unit but with no interaction to other two crushing units. There is a primary crushing unit with a manipulable feeder for pushing rocks to the primary jaw crusher received from an excavator feeding the feeder. The secondary crushing unit with a gyratory crusher receives pre-crushed material from the primary unit. The secondary unit has also a manipulable feeder with a small storage volume and a variable speed driven conveyor receiving pre-crushed material from secondary feeder and non-crushed material after the secondary crusher for re-crushing in the secondary crusher.

Table 13. *Upper*: load disturbance responses for PI (solid), PID (dotted), PPI (dashed) and PPID (bold) controller. *Lower*: control signals.

	Secondary crusher - feed chute control loop	**Tertiary crusher - cavity level control loop**
Controlled variable (CV)	Feed chute level	Tertiary crusher level
Manipulated variable (MV)	Feeder 1 speed	Feeder 2 speed + Conveyor 2 speed
Disturbance variables (DV)	Feed chute output flow (unmeasured), Feeder 2 speed (known), Conveyor 1 level (measured)	Conveyor 2 level (unmeasured), Feed chute level (measured), Secondary crusher CSS (known)
Feed-forward variables	Feeder 2 speed	Feed chute level
Limiting controls	Jaw crusher level max control, secondary crusher level max control	Secondary crusher level max control, Conveyor 2 level max control

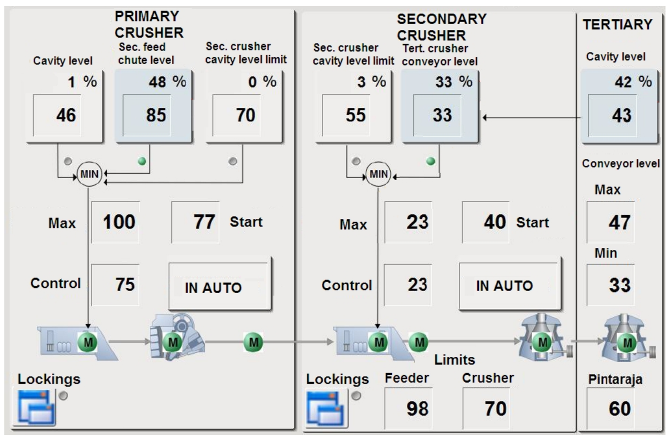

Figure 14. Excavator cabin control room has a touchscreen with above control diagram display for entering setpoints and process limits.

The outcome of the secondary crusher goes through a screen where non-crushed rocks are returned back to the secondary crusher. The crushed material continues to the tertiary crushing unit. The tertiary crushing unit has a long uplifting conveyor with a screen and, finally, a gyratory crusher for producing final aggregate fractions. There are ultrasonic material level sensors mounted on each crusher cavity level but also on the uplifting tertiary conveyor. The crusher cavity levels are calibrated to have a measurement range 0 - 100% where 0% equals to a conveyor running empty. The secondary crushing unit has the variable-speed driven secondary feeder which is manipulated for regulating the tertiary crusher cavity level. The transportation delay is close 60 secs due to several conveyors, secondary crusher and screens between the secondary feeder and the tertiary crusher. Process static gain between the manipulated feeder speed (MV, 0-100 %) and the controlled tertiary crusher cavity level (CV, 0-100%) is variant but less than 1.

The secondary crusher cavity level and the tertiary crushing unit's conveyor level are used as constraints for manipulating the secondary feeder speed. The limiting controller structure is used for tackling material level limitations given as setpoints for these controllers. The limiting controllers are of PI controller type. The process between MV and CV is an IPDT process. The PID controller would do the trick but due to rather long relative dead times compared to other process dynamics, the system needs dead time compensation. Thus, a PPID controller was selected as the main controller. Due to insignificant residence time of the system, the IPDT model with static gain with dead time is capable of encapsulating the essence of the process. The IPDT model is obtained after executing a step change in the secondary feeder speed. The obtained model parameters for the controller design were k = 0.1 and L = 58 sec.

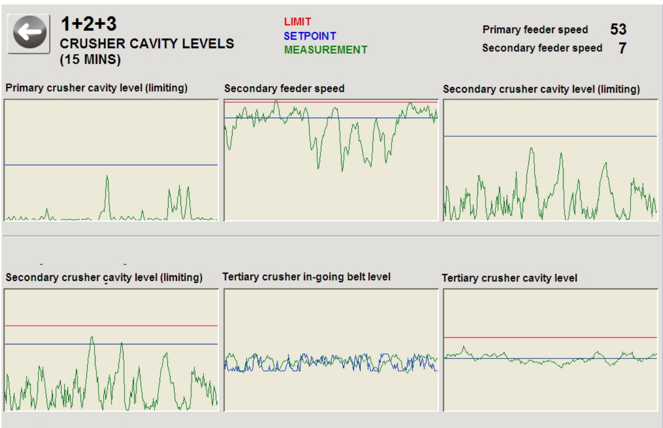

Figure 15. Crusher cavity level trends for 15 minutes on an excavator touchscreen.

REFERENCES

Airikka, P. (2012a), Simple Continuous-Time Identification Method for PID Controlled Crushing Processes, *IFAC Workshop on Automation in the Mining, Minerals and Metal Industries*, Gifu, Japan.

Airikka, P. (2012b), Event-Based Predictive PI Control for Crushing Plants, *IFAC Workshop on Multivehicle Systems*, Espoo, Finland.

Airikka, P. (2012c), Dynamic Flow and Level Simulation of a Mobile Crushing Plant for Process Control Design, *9th International Mineral Processing Seminar (Procemin 2012)*, Santiago, Chile.

Airikka, P., Aro, P. (2012d), Feed Control of a Mobile Rock Crushing Plant, *9th International Mineral Processing Seminar (Procemin 2012)*, Santiago, Chile.

Airikka, P., (2013), Stability Analysis of a Predictive PI Controller, *18th Mediterranean Conference on Control and Automation (MED'2013)*, Crete, Greece.

Evertsson, C.M., (2000), Cone Crusher Performance. *Ph.D thesis*, Chalmers University of Technology, Sweden.

Hägglund T. (1996), An industrial dead-time compensating PI controller, *Control Eng. Practice*, Vol. 4, No. 6, pp. 749-756.

Itävuo, P., Jaatinen, A., Vilkko, M. (2011), Simulation and advanced control of transient behaviour in gyratory cone crushers. *8th International Mineral Processing Seminar (Procemin 2011)*. Santiago, Chile, pp. 63–72.

Itävuo, P., Vilkko, M., Jaatinen, A. (2012), Specic Energy Consumption Based Cone Crusher Control, *IFAC Workshop on Automation in the Mining, Minerals and Metal Industries*, Gifu, Japan.

Itävuo, P., Vilkko, M., Jaatinen, A., Viilo, K. (2013), *Dynamic modeling and simulation of cone crushing circuits*, Minerals Engineering 43–44 (2013) 29–35.

Johansson, A. (2009), Modelling and simulation of cone crushers. *IFAC Workshop on Automation in the Mining, Minerals and Metal Industries*, Viña del Mar, Chile.

Sbarbaro, D. (2005), Control of crushing circuits with variable speed drives, *16th IFAC World Congress*, Praque, Czech Republic.

Whiten, W.J. (1972), *The Simulation of Crushing Plants with Models Developed using Multiple Spline Regression*, Journal of the South African Institute of Mining andMetallurgy, May, pp. 257-264.

Whiten, W.J. (1984), Models and control techniques for crushing plants, *In the book of Control '84 Mineral/Metallurgical Processing, Chapter 22*, AIME, pp. 217–225.

Åström K.J, T. Hägglund (1995), *PID Controllers: Theory, Design and Tuning*, 2nd ed., Instrument Society of America, USA.