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Automatic Control of Mass Flowrate in an Ore Feeding System with Time-Delayed Gain-Variant Dynamics

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Abstract — Regulatory control systems are used to adjust an outcome variable to a specified reference value of interest. They are widely applied in process control and manufacturing automation and have fundamental importance for industrial productivity and operational safety. Nevertheless, the absence or inefficiency of regulatory control in many industrial operations is also a common issue, that must be overcome with proper use of Control Systems Engineering. In this context, this article presents the development and implementation of a strategy for regulatory control of ore mass flowrate in a mineral processing facility. Effective mass flowrate regulation is a primary necessity for productive operation of mineral plants, though it is often neglected in practice. The mineral facility concerned in this work, although being simple, posed some challenges for the development of a regulatory control, due to its significant dead time and gain-variant dynamics. Nevertheless, a suitable process modeling and controller design were attained, leading to excellent results in regulatory performance. This article is organized as follows: firstly, the industrial process and the control problem are introduced. Then, the development of a dynamic model for the process is described, with special emphasis to its inherent dead-time and gain-variant dynamics. Next, it is presented the controller design strategy, based on the application of the root locus method to time-delayed systems. Finally, the implementation of the control system and the results obtained are discussed. Preserving due proportions, the strategy developed in this work can be used as a general framework to design regulatory control for similar industrial facilities.

Index Terms — dead time, mineral processing, PID control, process control, regulatory control, root locus method, time-delay.

I. INTRODUCTION

BRAZIL is the world leading producer and exporter of iron ore, mainly due to operations of VALE, a global mining company with several operations in the country, delivering iron, copper, manganese, and nickel ores. Those operations are located mainly in the Northern state of Pará, and in the Southeastern state of Minas Gerais. From 2014 to 2019, the company's average yearly production of iron ore was about 342 million t/year. Most of such production was exported overseas, mainly to Asian and European countries.

Mineral processing plants implement several stages of ore beneficiation, necessary to convert raw ore from the mines into ore products. Those plants comprise several processing facilities as crushing, screening, grinding, flotation, and others, which are operated by means of a plantwide digital control system, either SCADA- or DCS-based, using many process control loops.

The most important process variable in a mineral processing plant is the *ore mass flowrate*, measured in t/h. The transportation of bulk ore between mineral facilities is usually done by belt conveyors, and the ore flowrate is measured by a *conveyor weigher* (also known as *dynamic weigher* or *belt scale*), which is a heavy instrument assembled directly in the conveyor structure [1]. The conveyor weigher has two primary sensors: piezoelectric loading cells to sense the unit weight of material on the belt, in t/m; and an incremental encoder to sense the belt moving speed, in m/s. By combining the unity load and the belt speed measurements, the weigher computes the material flowrate, in t/h, with a typical precision of $\pm 0.5\%$ of its full measurement range. The ore flowrate in a conveyor is not defined by the belt speed, but by the speed of the feeder that discharges the ore on the conveyor belt. Therefore, by changing the feeder (actuator) speed it is possible to control the ore flowrate using its measurements provided by the conveyor weigher (sensor).

This work describes the modeling and control of ore flowrate

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for the ore feeding system “RF-A” of the Fábrica Iron Ore Processing Plant, in the city of Ouro Preto, Minas Gerais state, Brazil, operated by VALE. The next section introduces the industrial process and the control problem.

II. THE INDUSTRIAL PROCESS

The industrial process concerned in this work is an ore feeding process, in which the ore stored in a silo must be transported to a mineral concentration facility at a specified flowrate. Figure 1 illustrates the industrial facilities involved. Raw ore originally stored in a stockyard is retrieved by a reclaimer machine and transported by a belt conveyor circuit to a storage silo, shown in Figure 2(a). At the bottom side of the silo there are two independent feeders, RF-A and RF-B, that retrieve the ore in the silo and discharge it on their respective belt conveyors, AB04-A and AB04-B, shown in Figure 2(b). Those conveyors then transport the ore to the concentration facility for further processing. The speed of each feeder defines the ore mass flowrate in the respective conveyor, which must be controlled to match the demand for ore of the concentration facility, specified by the plant operators. Each feeder is powered by an induction motor driven by a frequency inverter (variable frequency driver), so that the feeder speed – and thus its ore flowrate – can be varied within a specific range. By increasing the speed of the feeder, the ore flowrate on the conveyor increases, and vice-versa. Figure 3 shows the feeder RF-A.

The feeders have a singular mechanical design to retrieve ore from the silo. The feeder body is supported by a truck coupled

to the motor shaft through a connecting rod, so that the unidirectional rotary motion of the motor shaft generates a linear back-and-forth motion of the truck, similar to the motion of the rods that drive the wheels of a locomotive. This back-and-forth motion causes the feeder to discharge the ore from the silo on the conveyor belt. The discharging rate is proportional to the frequency of the back-and-forth motion of the feeder.

The feeder (actuator) and the conveyor weigher (sensor) are something away from each other. Therefore, after the ore is released by the feeder on the conveyor, it reaches the weigher position after a fixed *time delay* (also known as *dead time*) T_d , since the conveyor belt operates with constant speed. Time delays are one of the most common non-linearities in industrial process. Short time delays may have negligible impact on system performance; however, long time delays usually bring strong degrading effects on process stability and may lead to unstable and unsafe operational conditions. Therefore, time delays must be addressed with proper Control Systems Engineering techniques.

The best way to deal with time delays is preventing them during the design of an industrial plant. However, there will always be situations in which this is not technically feasible. In the case of a mineral processing plant, it is physically impossible to install a conveyor weigher immediately after the feeder position, due to mechanical assembling and maintenance restrictions required by a weigher. Thus, the solution is installing the weigher at the closest suitable distance from the feeder, in order to reduce the time delay the much possible. This

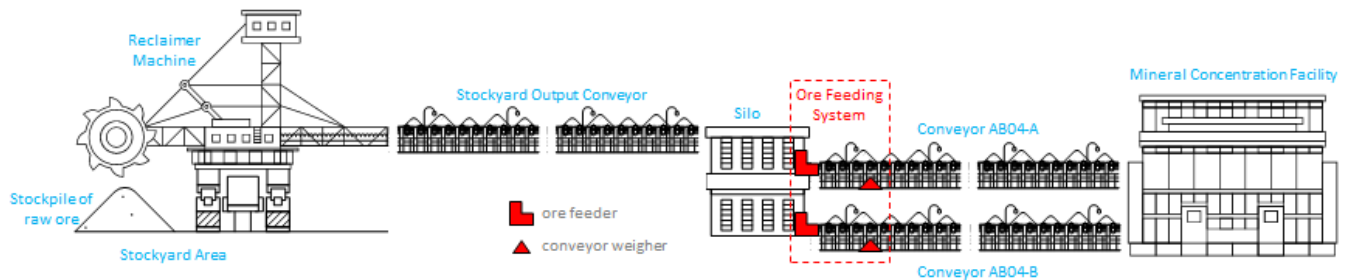
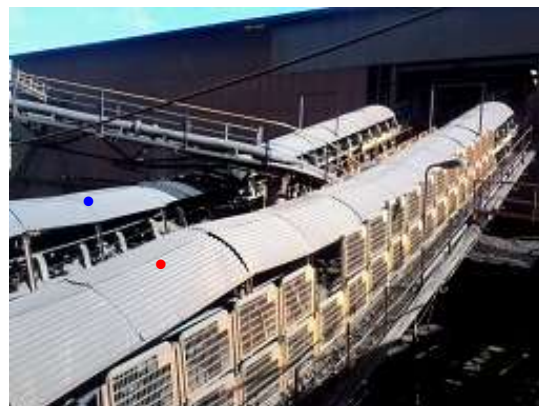


Fig. 1. Illustration of the industrial facilities.



(a)



(b)

Fig. 2. (a) Ore storage silo. (b) Conveyors AB04-A (●) and AB04-B (●).

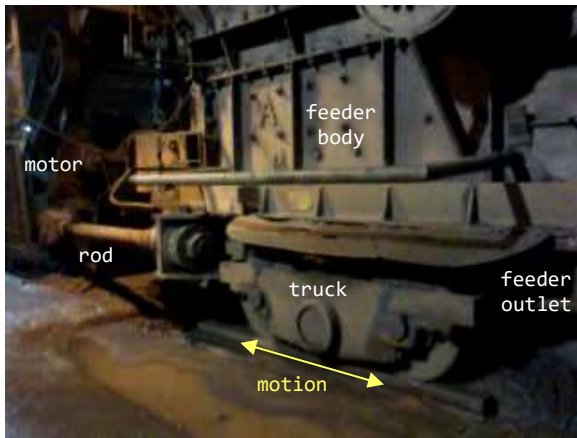


Fig. 3. Feeder RF-A.

care is crucial to allow an effective flowrate control. This work does not concern whether the weigher in conveyor AB-04A was installed at a proper position from feeder RF-A. Rather, it concerns to the estimation of the time delay T_d , required to develop a dynamic model of the ore feeding system. This is discussed ahead in Section 4.1.

III. ORIGINAL SYSTEM PERFORMANCE

The feeder RF-A was reported to operate under closed loop control. However, whenever the feeder started operating from a stopped condition, there were very high overshoots and excessive oscillations in the ore mass flowrate in conveyor AB-04A, as shown in Figure 4. The figure covers a period of 1 hour and shows three signals: the set-point (black), the measured flowrate (red), and the control action (green) sent to the feeder by the *programmable logic controller* (PLC) [2] responsible for all controls of the industrial facility. The signals values were retrieved from the Plant Information Management System (PIMS), which records continuously many process variables of the industrial plant, and their historical values can be easily retrieved for analysis.

In Figure 4, the set-point was fixed at 850 t/h. During the first

half of the plot, the ore flowrate was reasonably stabilized around the set-point, despite its inherent noise and fluctuations due to variations in process conditions. At time 13:42:10, the feeder stopped operating due to failure in a downstream equipment whose operation is *interlocked* to the operation of the feeder. In such a situation, the PLC stops all upstream equipments, including the feeder, to prevent damages to the failed equipment. When the feeder stopped, the ore flowrate dropped to zero, despite the flowrate signal indicated a small constant value of 49.6 t/h during all the time the feeder was stopped. This small false flowrate value was just a bias error in the conveyor weigher. When the flowrate signal dropped to near zero, the error between the set-point and the flowrate signal suddenly increased and caused the control signal to also increase up to a saturation level of 85 units. The system remained in this condition until the failure in the downstream equipment had been fixed and the production line could start running again. At time 13:46:12, the feeder started operating, and the control system sent to the feeder a constant control action of 40 units for 20 seconds and then released an aggressive variable control action causing a very bad transient flowrate response, with excessive oscillations.

It would be expected that the ore flowrate increased slowly towards the set-point value. However, it got too large overshoots, with a maximum peak of 1,499.8 t/h (76,45% higher than the set-point) and oscillated for an excessive settling time of about 450 seconds (7.5 minutes). This was a very aggressive control that must be prevented, especially in a process with significant time delay. Large overshoots, excessive oscillations, and long settling times are undesired behaviors because they stress the equipments, may lead to unsafe operational conditions, and introduce unduly variations in the next mineral processing stage, performed by the ore concentration facility.

In many cases, to prevent large overshoots and excessive oscillations in the flowrate, the plant operators often switched the control system to “manual mode” so that they can set a low

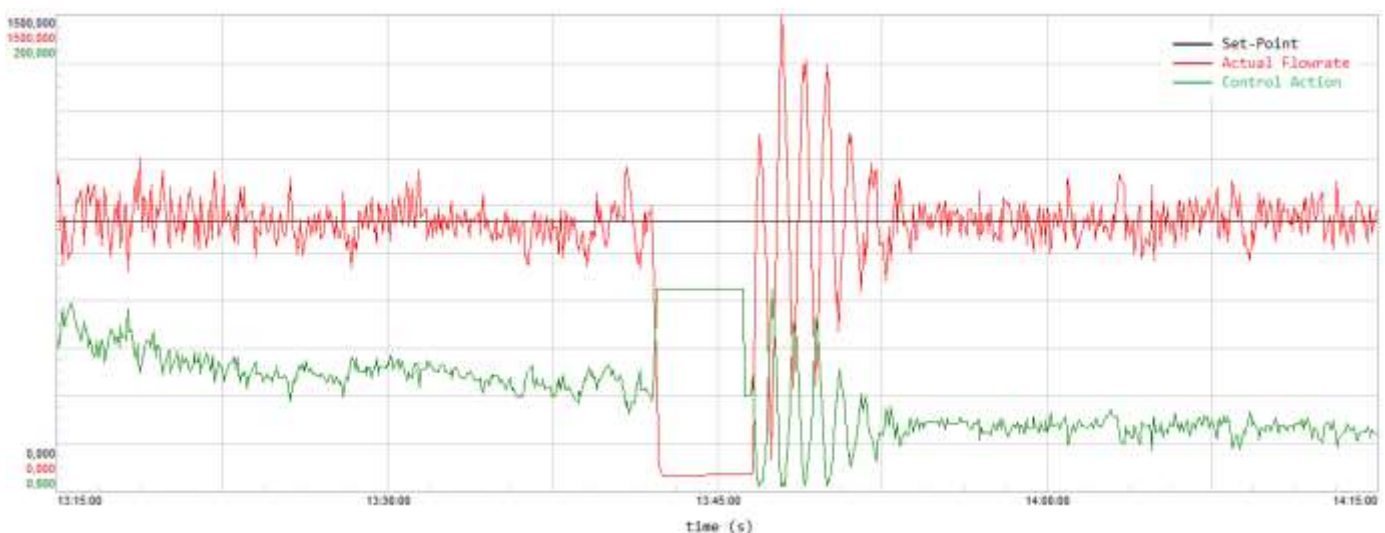


Fig. 4. Original performance of the ore feeding system.

control action to start operating the feeder, and then increase gradually the control action, trying to reach the set-point. The problem with this approach is that changing manual control actions usually take too long times to make the flowrate to reach the set-point, thus decreasing the productivity of the downstream concentration facility.

The bad process performance exemplified in Figure 4 is an example of how the incorrect use of a control system can result in extremely varying process conditions and stressing operation of the equipments, with potential risks of structural damages on them due to overload.

Many other bad performance occurrences were analyzed using recorded data from the control system. It was found that the ore feeding system already had been into closed loop control with a proportional-integral-derivative (PID) controller. However, the control loop was incorrectly implemented due to the following reasons: (1) The PID parameters were arbitrarily chosen, with absolutely no design criteria, by technicians with little or no knowledge of what a PID control loop is; (2) The effects of the time delay in the flowrate on the control loop were not considered; (3) The PID controller was implemented in its full form including the derivative action, when this must had been prevented due to the significant noise in the ore flowrate; and (4) No digital signal conditioning was implemented to reduce the noise in the measured flowrate.

This situation motivated this work, to understand, model, and effectively control the ore flowrate of the ore feeding system. The understanding of the system was focused on its performance, as described in this section. The development of a suitable model for the system, and the design of an effective controller are described in the next sections.

IV. DEVELOPING A SYSTEM MODEL

Obtaining a suitable mathematical model for a dynamic system or process is the foundation stone for most control design methods. A dynamic model represents how the variables in the system evolve along time when the system is subjected to a given initial condition or input. A *suitable* model is one that represents the main dynamical characteristics of the system, without being excessively complex for the purposes the model is intended to be used. Hence, the modeling involves a trade-off between the simplicity and the accuracy of the model. In deriving a reasonably simplified model, it may be necessary to ignore certain physical properties of the system [3].

The theory of system dynamics is very broad, and many modeling methods are available, including linear and non-linear methods, time response or frequency response methods, deterministic and stochastics methods, etc. The choice of a method depends on the system under study and the purpose of the model. In this work, developing a model for the ore feeding system was motivated by the need to *design an effective control loop* for the system, so that its *time response* (ore flowrate) achieve a good performance. By analyzing historical process data, it was observed that when the system is in open loop, without any controller, its response to a constant input is fast and overdamped, suggesting that the system could be properly modeled by a single-input-single-output (SISO) linear transfer

function model with time delay [3][4][5][6]:

$$G_p(s) = G(s)e^{-T_d s} = \frac{K \cdot p}{s+p} e^{-T_d s} \quad (1)$$

where T_d is the time delay, K is the open loop gain, and p is the pole of the transfer function.

A. Estimation of the Time-Delay

One way to estimate the time delay T_d of the ore feeding system is simply measuring the elapsed time from the instant when the feeder (actuator) starts operating to the instant when the ore in the conveyor reaches the weigher (sensor) position. This is a purely visual experiment performed by two persons, one at the feeder position, and the another close to the weigher, communicating via radio to the Control Room. Another way to obtain an estimate of T_d is by checking, in the technical documentation of the equipments, the values of the conveyor belt speed and the travel length for the ore from the feeder to the weigher, and then calculate the time delay by dividing the length by the speed. The first way is affected by subjective perceptions of the people performing the experiment, and the second way may lead to unrealistic results if the equipment data in the technical documentation differs from their actual operating conditions, as is often the case in mineral processing plants, due to low concerns in updating technical documentation.

To avoid the above problems in determining a realistic estimate of the flowrate time delay, and considering that equipment operation data could be easily obtained from the control system (PLC), the method of *cross-correlation* [7] was used to estimate the time delay T_d , using historical set-point (process input) and flowrate (process output) signals. Cross-correlation is a useful method to estimate “time of travel” and is the basic technique used in radar systems for determination of target distance based on the “time of flight” of radar waves.

The cross-correlation of two discrete-time signals $x(k)$ and $y(k)$ is given by equation (2) [7]. It computes the correlation between the signals across all possible time shifts d between them. The cross-correlation attains its maximum value when the signals are the most synchronized as possible. Therefore, when one signal is delayed from the another, the maximum cross-correlation value occurs for a time shift equal to the time delay between the signals. This makes the cross-correlation very useful to estimate dead times in dynamic systems from its input and output signals.

$$R_{xy} = \frac{\sum(x(k)-\mu_x)(y(k-d)-\mu_y)}{\sqrt{\sum(x(k)-\mu_x)^2} \sqrt{\sum(y(k)-\mu_y)^2}} \quad (2)$$

where μ_x and μ_y are the mean values of x and y , respectively.

Figure 5(a) shows the control action and the corresponding ore flowrate signals that were cross-correlated to estimate the time delay between them. The signals were recorded with a sampling period $T_s = 0.5$ seconds for 1,350 samples (675 seconds = 11.25 minutes). Figure 5(b) shows the cross-

correlation between the signals, whose maximum value occurs for a time shift of 65 sampling periods, corresponding to 32.5 seconds (= 65 samples × 0.5 second/sample). Hence, the cross-correlation estimate for the time delay is $T_d = 32.5 \pm 0.5$ seconds. Notice that the estimate will always be a multiple of the sampling period T_s , and its precision is equal to $\pm T_s$.

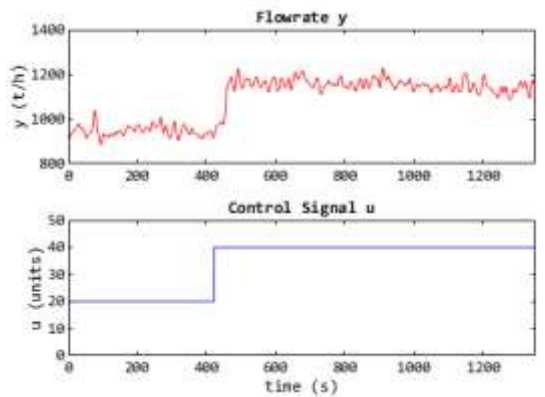
The transfer function model of the time delay T_d is given by the complex exponential function [3][4][5][6]:

$$e^{-T_d s} = e^{-32.5s} \tag{3}$$

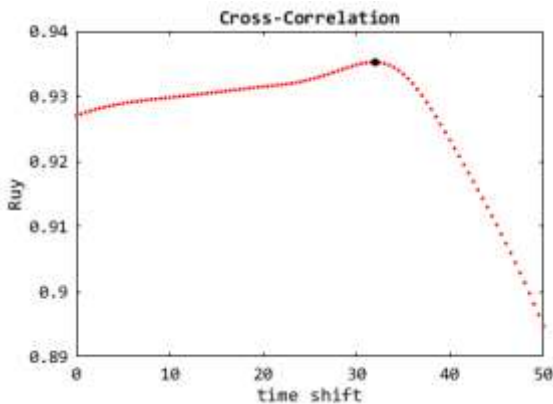
Notice that the time delay of the ore feeding system is high compared to its rising time. Any control design strategy for the system must carefully consider its time delay.

B. Estimation of the Open Loop Gain

After determining the time delay of the ore feeding system, the next step was to determine its first-order transfer function $G(s)$ in (1), which involves estimating the open loop gain K and the pole p . Notice from Figure 5(a) that when the system is in open loop, disregarding its time delay, its response (ore flowrate) to a constant input is fast and overdamped. Therefore, it was assumed that the dynamics of the system could be modeled by a linear first-order transfer function [3][4][5][6]:



(a)



(b)

Fig. 5. Cross-correlation analysis for time delay estimation.

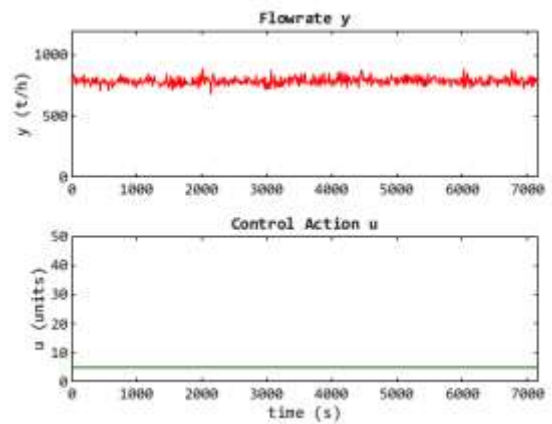
$$G(s) = \frac{K.p}{s+p} \tag{4}$$

where K is the open loop gain, and p is the pole of the transfer function. The open loop gain of the system can be accurately estimated from steady state operation of the system, as the ratio between its stationary response y_{ss} to a constant input u_o :

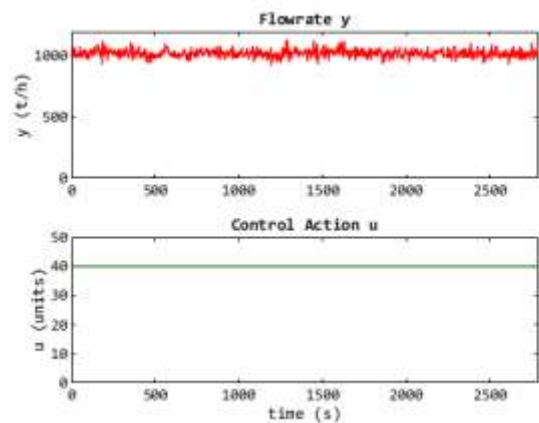
$$K = \frac{y_{ss}}{u_o} \tag{5}$$

Since the flowrate $y(t)$ has an inherent noise, the value of y_{ss} must be taken as the mean value of $y(t)$ on steady state condition. Figure 6 shows the variables $u_o(t)$ and $y(t)$ for two steady state tests on the ore feeding system. The signals were retrieved from the PIMS system with a sampling period of 1 second. For the first test, in Figure 6(a), the steady state values were $u_o = 5.0$ and $y_{ss} = 787,15$ t/h, so that the open loop gain calculated by (5) result as $K = 157.43$. For the second test, in Figure 6(b), the values were $u_o = 40.0$ and $y_{ss} = 1,022.34$ t/h, leading to $K = 25.56$, a very different value from the first test.

The significant difference between the values of K was not expected, and it was discussed with the plant operators. They informed that, when the system operates in manual mode (open loop), the manual control input u_o is usually chosen between 5% and 40%, corresponding to stationary



(a)



(b)

Fig. 6. Input and output signals from steady state tests.

flowrates y_{ss} of about 800 t/h and 1,100 t/h, respectively. However, the value of y_{ss} may vary slightly for the same value of u_o due to changes in the process conditions, which would be the reason for the variations in the open loop gain K . Hence, to get a deeper understanding on the behavior of K , several additional steady state tests were performed on the system for different values of u_o .

Figure 7 shows the resulting pairs of values $\{u_o; y_{ss}\}$, obtained from 30 steady state tests for the ore feeding system, and the corresponding least-squares line fitted to them. Two unusual aspects were evidenced from those results. First, the points $\{u_o; y_{ss}\}$ had a significant spread around the least squares line, indicating that the open loop gain K changed from one test to another, as reported by the plant operators. In other words, the ore feeding system has a *time-variant open loop gain*. This is more evident for the points having $u_o = 15.0$, whose values of y_{ss} varies significantly. The same is observed for points having $u_o = 10.0$. The second unusual aspect is that the least square line has an intercept value of 777,39 when $u_o = 0$, meaning that a flowrate of 777,39 t/h would occur, in average, when there is no manual control action.

Another subtle evidence that the open loop gain K would be time-variant comes from the previous Figure 4. During the first half of the plot (before the system stopped), the control signal decreased slowly, in average, although the flowrate remained, in average, around the set-point. This means that the open loop gain increased, tending to amplify the effect of the control signal on the flowrate y , but since the (yet bad implemented) original PID controller worked to keep y adjusted to the set-point, the controller reduced its integral action, thus decreasing the control signal.

The unusual relationship observed between the manual control action and the resulting ore flowrate was discussed with the plant operators, and the PLC program was also verified. It was found that the control action sent by the PLC to the feeding system was the sum of a constant component, u_c , and a variable component, u_v . The constant control action u_c was intended to keep the feeder running at a minimal speed to avoid it stopping and prevent damages by frequent starts, just because of its back-and-forth design, described in Section 2. A direct limitation of such design is that the feeder could not be operated with speeds lower than 40% of its maximum speed, because the mechanical friction

between the movable parts of the feeder would increase drastically for low speeds, causing torque overload on the feeder motor. When the feeder starts running from a stopped condition, its high starting torque ensures that it reaches the minimum speed of 40% and keep running. The PLC limits the speed at this minimum value when the feeder is running.

On the other hand, the variable control action u_v corresponds to either the manual control action, when in “manual” mode, or the controller output, when in “automatic” mode. Therefore, the control actions in Figures 4, 5, 6, and 7 are u_v , not the total control action u sent to the feeding system. Moreover, the constant control action u_c was responsible for the average flowrate of 777,39 t/h (intercept of the least square line) when u_v is set to zero. This is the flowrate value corresponding to minimum operating speed of the feeder.

The value of the constant control action set in the PLC was found as $u_c = 120$, and the maximum allowable value for the total control action u was found as $u_{max} = 258$. Hence, u_c corresponds to 46.5% of the available range for the total control action u , and the variable control action u_v was left the remaining 54.5% of the range. This means that the use of such a significant constant control action u_c reduced the control rangeability of the variable control action u_v which is in fact responsible for the dynamic control of the process. A low control rangeability limits the effectiveness of an automatic control. A new discussion with the plant operators was carried out aiming to eliminate u_c or, at least, reduce its value. However, it was concluded that the elimination of u_c would bring the risk of severe damages to the feeder or its motor, so that the value $u_c = 120$ set in the PLC would not be changed without a deep evaluation of the entire system. Hence, the value of u_c was not modified, and the control system was developed under this constraint.

In the open loop gain equation (5), the value of u_o must be the process input, that is, the total control action sent to the feeding system. By adding the constant value of u_c to the variable values of u_v in Figure 7, a new scatterplot between y_{ss} and u_o ($u_o = u_v + u_c$) can be obtained with its corresponding least squares line (the red line), as shown in Figure 8. Notice that the angular coefficient of the least square line is exactly the same as in Figure 7, and its intercept became closer to zero when compared to Figure 7. Ideally the intercept should be zero because, a null control

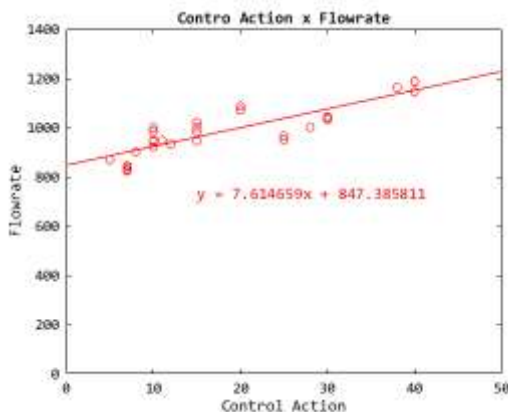


Fig. 7. Relationship between flowrate and control action.

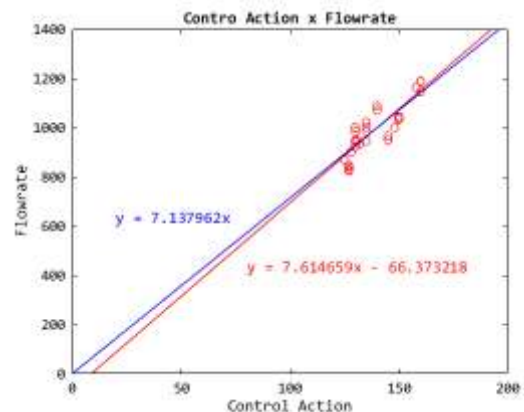


Fig. 8. Corrected relationship between flowrate and control action.

action u_o actually stops the feeder motor and no flowrate will result in the system. This did not happen for the red line because of random variations in the process data. However, it is possible to force the least squares line to pass at point $\{u_o = 0.0; y_{ss} = 0.0\}$, as shown by the blue line in Figure 8. Now, the intercept is null, and the angular coefficient of the line changed slightly to a new value, which represents the best estimate of the average closed loop gain, $K = 7.137962$, based on the test data. It is said average because of its time-variant behavior, which leads the steady state flowrate value y_{ss} to vary for a given value of u_o .

C. Estimation of the System Pole

After estimating the open loop gain K , the next step was to determine the pole p of the transfer function model (4). The simpler way to accomplish this is by performing a step response test with the system in open loop and record the control action $u(t)$ and the ore flowrate $y(t)$. Since the time delay T_d of the process is already known, the recorded flowrate signal $y(t)$ can be shifted backwards by T_d to be synchronized to $u(t)$, and then used to estimate the pole p through a system identification method [8].

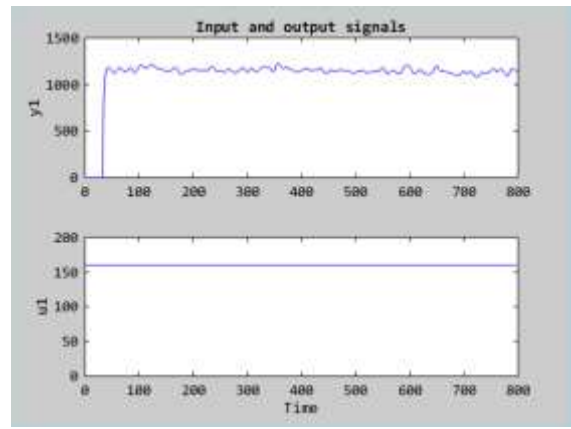
Figure 9 shows the control action and the flowrate signals obtained from a step response test on the feeding system. The feeder was completely stopped and then a constant manual control action $u(t) = 160$ was used as step input, and the resulting output flowrate $y(t)$ was recorded. These two signals were retrieved and used to estimate the pole p using a specific computational tool [9] to avoid wasting time coding system identification algorithms. The obtained estimate was $p = 0.531609$. Hence, from (1) the complete process model of the ore feeding system is:

$$G_p(s) = \frac{K \cdot p}{s+p} e^{-T_d s} = \frac{3.794605}{s+0.531609} e^{-32.5s} \quad (6)$$

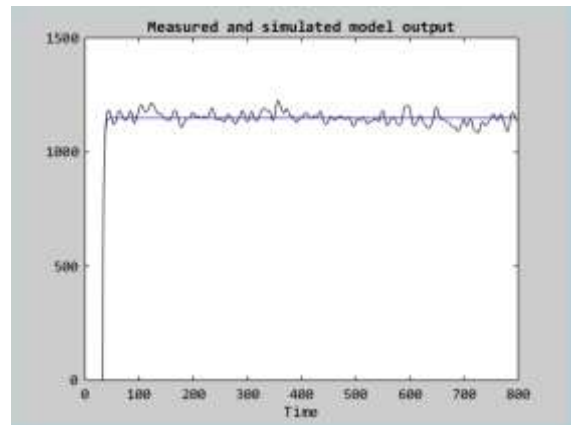
D. Reduction of Noise in the Process Variable

As shown in Figure 4, the ore flowrate has a significant high-frequency noise. When performing regulatory control, the error signal computed by the difference between the set-point r and a noisy output variable y will fully incorporate the existing noise in y . If such noisy error signal is used as input to an automatic controller, the resulting control action may also be noisy in some way.

The simpler way to reduce the noise of a signal is computing its moving average. However, this method may lead to ineffective smoothing results just because it does not consider the frequency characteristics of the signal. A more efficient way to reduce high-frequency noise is designing a *low-pass filter* [10] with cut-off frequency close to the least relevant frequency of the signal. To determine the frequency components of a signal, it is necessary to perform a *density power spectrum* analysis [11][12][13]. To analyze the power spectrum of a signal, it is important to sample the signal with suitable sampling period to catch the relevant variations of the signal. As shown in Figure 6, the flowrate signal to be analyzed is essentially a DC signal with an



(a)



(b)

Fig. 9. (a) Actual control action and flowrate used to estimate the pole p . (b) Simulated flowrate for the estimated pole p .

additive noise. Hence, it is not necessary to sample it with high sampling rates.

The flowrate *noise* can be estimated by taking a steady-state flowrate signal and then subtracting its mean value. Figure 10 shows the power spectrum of the noise for the flowrate signal shown in Figure 6(a), suggesting that they contribute mostly to the noise. It suggests that a good choice for the cut-off frequency for a low-pass filter should fall within 0.04 Hz and 0.1 Hz. To avoid complex filter designs, it was decided to use a compound low-pass filter formed by

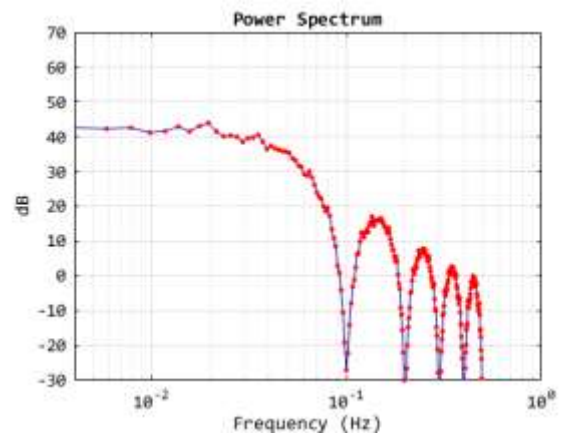


Fig. 10. Power spectrum of the flowrate noise.

n replicated first-order low-pass filters in series, which provides good signal filtering performance for process control applications. The transfer function of the filter is:

$$F(s) = \left(\frac{\omega_c}{s + \omega_c} \right)^n \quad (7)$$

where $\omega_c = 2\pi f_c$ is the filter cut-off frequency in rad/sec, and n is the order of the filter. Each pair of values $\{\omega_c, n\}$ defines a specific filter. After investigating several values for ω_c and n , the filter parameters were chosen as $n = 3$ and $f_c = 0.05$ Hz ($\omega_c = 0.314159$ rad/sec). The filter transfer function is then:

$$\frac{Y_F(s)}{Y(s)} = F(s) = \left(\frac{0.314159}{s + 0.314159} \right)^3 \quad (8)$$

E. Block Diagram of the System

The block diagram of the ore feeding system is shown in Figure 11. Following the industrial automation nomenclature, the signals SP, PV, and MV mean, respectively, the set-point, the process variable (variable to be controlled), and the manipulated variable (control action). As discussed in Section IV.B, in automatic operation mode, the control variable u is given by the sum of a variable component u_v , provided by the control system (PLC), and a constant component u_c to ensure the minimum speed required by the feeder. In manual operation mode, u_v is given by a “manual MV” control action provided by the plant operator from the Supervisory System. The total control action u is sent to the variable speed driver (VFD) which generates the AC current i to drive the speed v of the induction motor, which by its turn defines a hypothetical ore flowrate y_1 free of noise. The noisy flowrate y_2 is conditioned by the existence (1 = True ; 0 = False) of ore in the silo, and then is time-delayed resulting in the actual flowrate y_a , measured by the conveyor weigher. The weigher is wired to an analog input port of the PLC to sample the measured flowrate y , which is then smoothed by the low-pass filter (LPF). The filtered flowrate y_F is subtracted from the set-point signal SP to generate the error signal E as input to a PI controller (PID), discussed in the next section. The blocks in yellow represent together the ore feeding system, and are entirely modelled by the transfer function (6).

Having built a representative linear time-invariant model for the ore feeding system, regardless of its time-variant open loop gain K , and designed a suitable low-pass filter for conditioning the measured flowrate signal, the next step was to develop the regulatory control.

V. DEVELOPING A CONTROL STRATEGY

Before developing a control strategy for a system to improve its performance, it is important to understand what a “good performance” of the system would be, and to define a performance criterion to quantify it. Recall from Figure 4 that the bad performance of the system relates to high overshoots, excessive oscillations, and long settling time of the flowrate. Hence, the performance criterion was defined with regards to the time response of the system for a constant input signal, in terms of maximum overshoot M_p , settling time t_s , and average steady state error e_{ss} :

$$\begin{cases} M_p \approx 8 \% \\ t_s \approx 200 \text{ sec} \\ e_{ss} \approx 0 \text{ t/h} \end{cases} \quad (9)$$

The maximum overshoot and the settling time are performance metrics for the transient response of the system, whereas the steady state error is a performance metrics for its stationary response. The values in (9) were defined from discussions with people from the Operations and Engineering staffs of the industrial plant.

Assuming that the closed loop dominant dynamics of the ore feeding system can be approximated by an ordinary second-order transfer function with damping factor ξ and natural frequency ω_n [3][4][5][6], the maximum overshoot M_p and the settling time t_s in (9) are attained if the ore feeding system has the following closed loop dominant poles:

$$s_i = -\xi\omega_n \pm j\omega_n\sqrt{1 - \xi^2} \quad (10.a)$$

$$s_i = -0.02 \pm j0.0249 \quad (10.b)$$

The next step was to choose a controller type with the ability to make the closed loop system to have dominant closed loop poles the mostly close to (10.b). Since the control problem is of a regulatory type, and considering that it must be implemented in a PLC, the natural choice was the

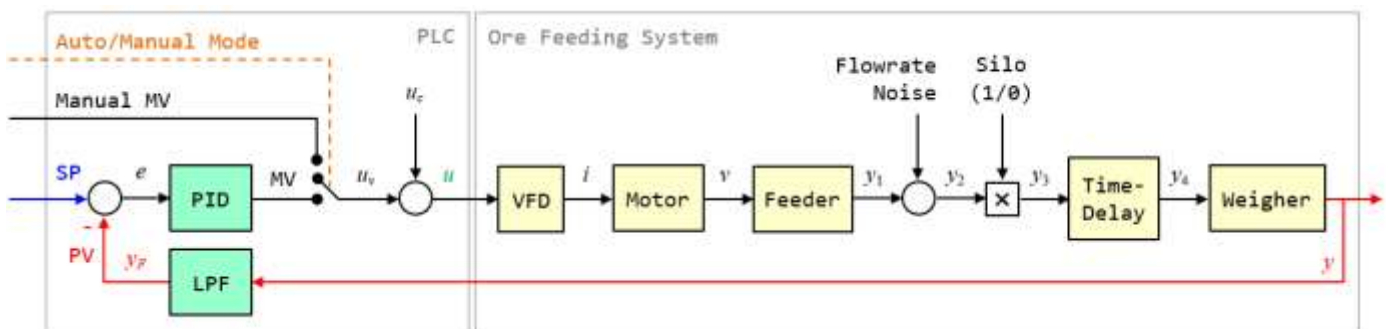


Fig. 11. Block diagram of the process control system.

proportional-integral-derivative (PID) controller [3][4][5][6][14], which is simple and robust to modeling uncertainties, and has proved to be versatile for many regulatory control applications in the industry. Moreover, it could be natively implemented in the PLC.

Since the ore feeding system has time-variant dynamics, a simple proportional controller is not sufficient for regulatory control of the flowrate. An integral control action would be able to compensate for the varying open loop gain of the system. A derivative action should be disregarded because it would amplify the noise in the flowrate, even though the flowrate signal is previously smoothed by the low-pass filter. Therefore, a *proportional-integral* (PI) controller was chosen for design and implementation.

A. Controller Design with the Root Locus Method

The transfer function model of the continuous PI controller is [3][4][5][6][14]:

$$C(s) = \frac{K_P \cdot s + K_I}{s} \tag{11}$$

where K_P and K_I are, respectively, the proportional and integral gains of the controller. Many different methods are available to design values for K_P and K_I according to the control problem and the process model. In this work, since the ore feeding system has a time delay, and a suitable transfer function model was developed, the root locus method [3][4][5][6] was chosen to design the controller parameters, due to two main reasons: it is an accurate and robust method to shape the closed loop dynamics of a system; and it can consider the process time delay directly, without the need for approximations, that would impair the design accuracy. Using the root locus method, the values of K_P and K_I are determined so that the *magnitude and phase conditions* for the close loop system are satisfied. The values were found as $K_P = 0.0161$ and $K_I = 0.0021$. Hence, from (11), the controller transfer function is:

$$C(s) = \frac{K_P s + K_I}{s} = \frac{0.0161s + 0.0021}{s} \tag{12}$$

Figure 12(a) shows the root locus plot for the time-delayed process $G_p(s)$ in open loop. The four branches of the root locus that do not touch the real axis are due to the time delay exponential transfer function. The black mark “x” is the pole of $G_p(s)$ at $s = p = -0.531609$. The green marks “x” are the complex poles (10.b) required to meet the performance criterion (9). Since the root locus of the process does not pass at s_i , the system needs a closed loop controller with the ability to change the root locus to pass at s_i . Figure 12(b) shows the root locus of the closed loop system. The red mark “x” is the pole of the controller $C(s)$ at $s = 0$, and the mark “o” is the zero of the controller at $-K_I/K_P = -0.1331$. The pink mark “x” corresponds to the *three* poles of the low-pass filter $F(s)$ in (8). Notice that the root locus now passes at the desired poles s_i .

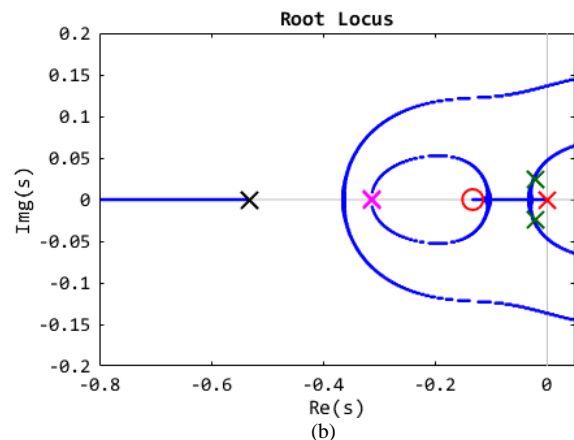
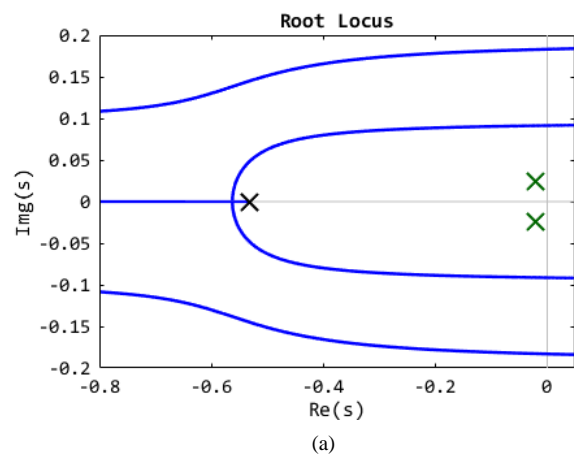


Fig. 12. (a) Root locus of the time-delayed feeding system. (b) Root locus of the closed loop system.

B. Simulating the Closed Loop System

The complete model of the ore feeding system, including the process model (6), the low-pass filter (8), and the controller (12), was implemented in a specific software [9] for simulation. Figure 13 shows the simulated flowrate for a set-point of 1,000 t/h. Its overshoot was 84.54 t/h, corresponding to 8.45% above its steady state value, and met the performance criteria for M_p in (9). The settling time was 203 seconds, and also met the performance criteria for t_s . The steady state error e_{ss} also met its corresponding specification. Hence, the controller was regarded suitable. Notice that this simulated performance represents a great improvement compared to the bad performance of the real system, shown in Figure 4.

VI. IMPLEMENTATION OF THE CONTROL SYSTEM

After designing the controller and confirming its ability to improve the system performance through simulations, the next step was the practical implementation of the control system in the PLC of the industrial plant, an Allen-Bradley ControlLogix™ [2]. The continuous-time models of the low-pass filter (8) and the controller (12) must be converted to a discrete-time form for implementation in the PLC. The sampling period used in the PLC for all control applications was fixed to $T_s = 0.1$ sec. The I/O interfaces of the PLC operate with fixed values between sampling instants, so that

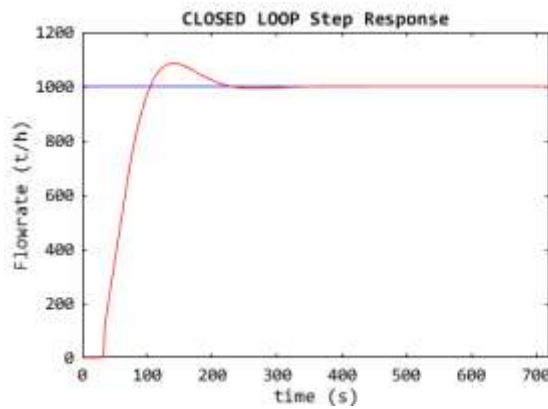


Fig. 13. Simulated closed loop response of the system model.

the *zero-order-hold* (ZOH) discrete equivalent [15] must be used. However, it was not necessary to compute the ZOH discrete equivalent of the continuous PI controller model (12), as done for the low-pass filter, just because the ControlLogix™ PLC has a native function block to implement a PID controller, given its parameters. Such function block is called PIDE [16], shown in Figure 14. It is very powerful and can implement many operational aspects of a PID controller as: direct or indirect realization of the PID equation, derivative smoothing, bumpless transfer, control action dead band, etc. Detailed information on the PIDE function block can be found in its related technical documentation. Due to operational requirements of the industrial plant, the PIDE function block of the ore feeding system was configured for *bumpless transfer* [4][14][16], to allow proper switching between automatic and manual operation modes.

VII. RESULTS AND BENEFITS

The control system implemented led to excellent regulatory performance of the ore feeding system. Figure 15

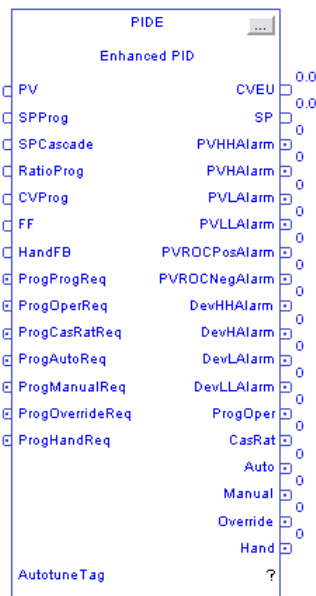


Fig. 14. The PIDE instruction block from the PLC.

shows the real curves of set-point r (blue), measured flowrate y (red), and variable control action u_c (green) for the closed loop system, for a period of 1 hour, from data recorded by the PIMS system. Initially, the feeding system was stopped, and the control signal was saturated at $u_v = 85$. The system started operating at time 00:57:08, when a constant control action $u_{v(manual)} = 40$ was set for 20 seconds, only to speed up the system flowrate. At time 00:57:29, the control was switched to the automatic control action u_v from the PI controller using *bumpless transfer*, and the control signal u_v increased up to a transient peak and then dropped smoothly to regulate the flowrate at the set-point level. Compared to Figure 4, the noise in the control signal was drastically reduced, due to the use of the low-pass filter, meaning a less stressing action on the feeder. The flowrate noise was also reduced due to the smoother control action.

Since the PIMS system records continuously many process variables of the industrial plant, their historical values can be easily retrieved for further analysis. This makes possible to validate the system model by simulating the ore flowrate using recorded actual set-point values as input to the system model, and then comparing the simulated flowrate to the recorded actual flowrate. This allows to verify the adherence between the responses of the system model and the real system.

Figure 16 shows the same actual set-point (blue) and flowrate (red) values from Figure 15 together with the simulated flowrate (black) for the closed loop system model. The simulated flowrate has a strong agreement with the actual flowrate, indicating that the system model represents with very good accuracy the dynamics of the real system. It is important to recall that the system model is *time-invariant* whereas the actual system is *time-variant* with regards to its open loop gain. Therefore, slight differences between the behaviors of the real and simulated systems are naturally expected.

The actual and simulated control signals corresponding to the flowrates in Figure 16 are shown in Figure 17. Although the signals have similar waveforms, the actual control signal has higher values than its simulated counterpart because the open loop gain of the real system would have been **lower** than the *average open loop gain* K of the system model, so that a **higher** control signal was needed by the real system to regulate the flowrate at the set-point. Along the time interval 9,000 ~ 32,000 seconds, the average values of the control signals were: $u_{actual} = 145.4914$ and $u_{simulated} = 112.0829$. Hence, according to equation (5), the open loop gain of the real system was $K_{actual} = y_{ss}/u_{actual} \approx 800/145.4914 = 5.4986$, whereas the simulated system has a gain $K_{simulated} = y_{ss}/u_{simulated} \approx 800/112.0762 = 7.137975$. As expected, $K_{simulated}$ is virtually equal to the average open loop gain K of the model (6), however, the open loop gain of the real system, K_{actual} , is lower than K . Again, this is a clear evidence that the real system has a time-variant open loop gain. Nevertheless, and more important, the PI controller was effectively able to regulate the flowrate at the set-point, with smooth transient response when the system starts operating, in opposite to the aggressive transient observed in Figure 4.

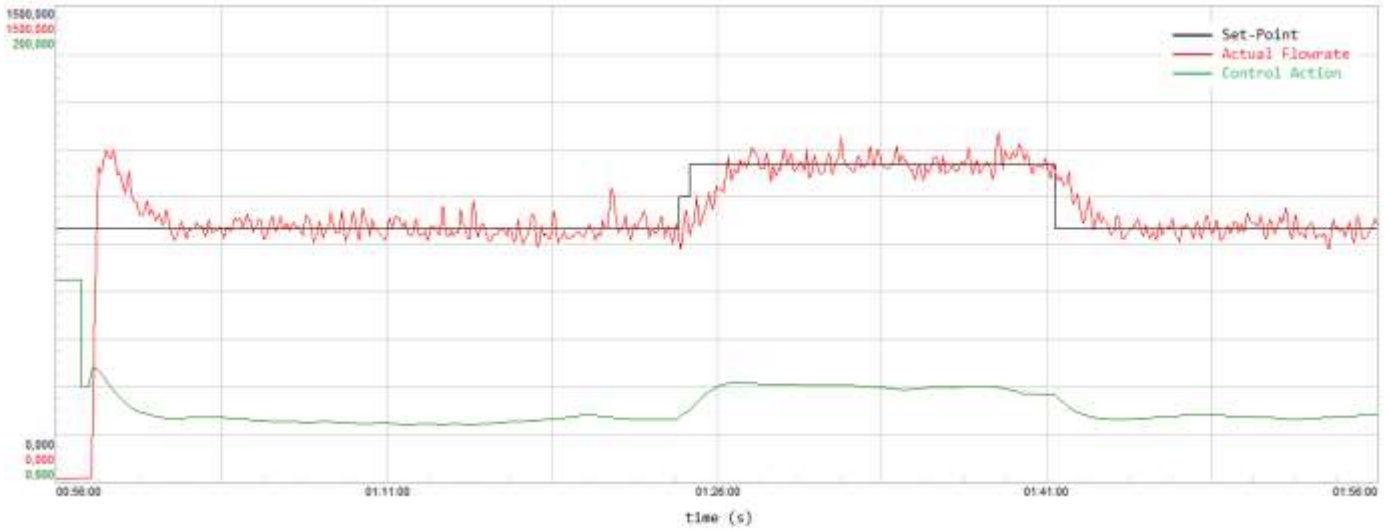


Figure 15: Performance of the ore feeding system with closed loop control.

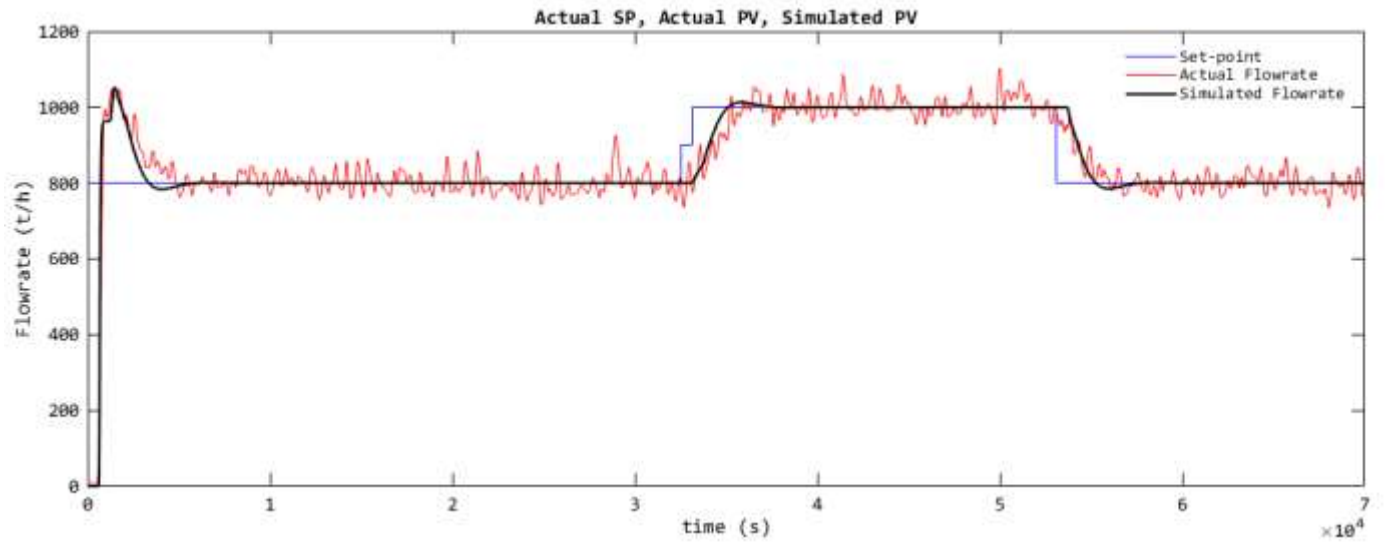


Figure 16: Actual and simulated closed loop responses, for the actual set-point.

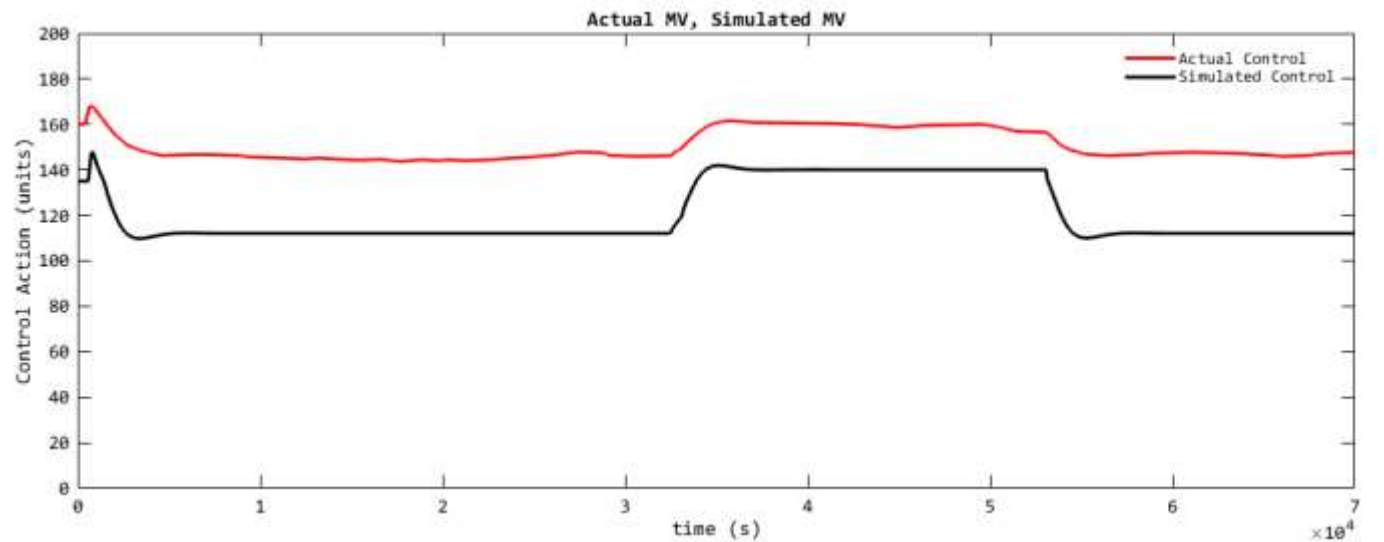


Fig. 17. Actual and simulated closed loop control signals.

All results showed previously were focused on the time response of the system. It's also desirable to translate those results in a more general benefit as the productivity. Using data recorded by the PIMS system, it was possible to compare the performance of the ore feeding system for operations before and after the implementation of the new control system. The regulatory performance metrics was chosen as the *standard deviation of the flowrate under steady state conditions for a constant set-point and no disturbances in the flowrate*. The lower the standard deviation, the best the regulatory performance, and vice-versa.

The period “before” the implementation of the control system comprised 58 days, from which 36 instances of steady state operations were collected from the PIMS system. The period “after” the implementation comprised 238 days, from which 117 instances of steady state operations were collected. The period “after” had a greater number of observed instances simply because more effort was done to monitor the control system after its implementation, without lack of generality. The standard deviation (variability) of the ore flowrate was computed for each observed instance. Figure 18 shows the histogram of variability for both periods, “before” and “after”. The *average flowrate variability* for the histogram “before” was 38.63 t/h, whereas for the histogram “after”, it was 21.74 t/h. Therefore, the control system brought a **productivity increase of 16.89 t/h** for the ore feeding system. This is a clear example of how a well-designed control system can bring strong benefits.

VIII. CONCLUSION

The purpose of this work was to develop an effective regulatory control for an ore feeding system, that could be implemented in the existing control hardware (PLC) of the industrial facility. The goal was attained successfully, although some technical challenges must have been overcome, especially regarding the time-variant open loop gain of the process. Although many real processes in the industry are complex, they may be properly represented by simple dynamic models for use in control design.

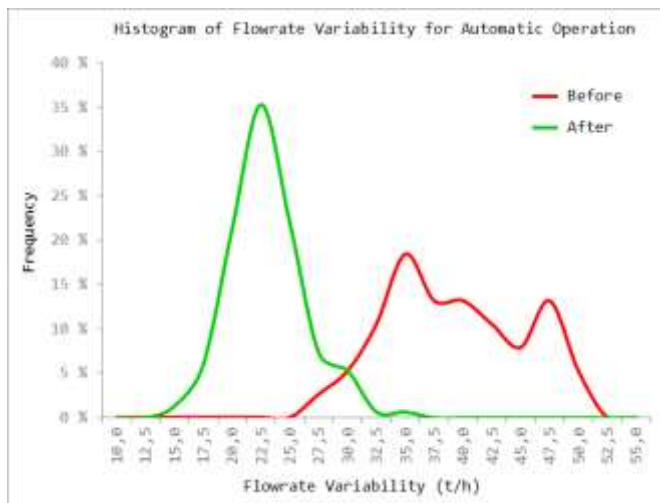


Fig. 18. Reduction in flowrate variability.

The root locus method proved to be powerful and accurate for designing the PI controller, which proved to be effective to control the ore flowrate, eliminating the need for advanced control techniques that would be difficult or even impossible to implement in a PLC, and would require additional investments.

As a suggestion to improve the process model, the level of ore inside the silo, measured by a radar-based instrument, can be correlated to the measured flowrate, to determine the relationship between them. If a precise relationship between these variables is identified, it will be possible to design more versatile control strategies to improve the disturbance rejection ability of the system, like cascade control, feedforward control or adaptive gain-scheduling control, still using the same control hardware (PLC). Such improvement was not done during this project because of the longer time required to develop and implement a more complex control strategy.

The modeling and control strategies presented in this work have been proven in practice several times on different mineral facilities that require regulatory control of ore flowrate in belt conveyors. The control strategy was also replicated to feeder RF-B, leading to similar results than the obtained in this work for feeder RF-A.

Finally, the plant managers were recommended to start a retrofitting project to replace the existing feeders with back-and-forth motion by new ones with unidirectional rotary motion. A unidirectional feeder provides a more uniform flowrate with reduced noise, improves the mechanical reliability of the system, and allows increased control rangeability from the elimination of the constant control action necessary to ensure the minimum operating speed of the existing feeders with back-and-forth motion.

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Controle Automático de Vazão Mássica em um Sistema de Alimentação de Minério com Atraso de Transporte e Ganho Variante

Resumo — Sistemas de controle regulatório objetivam ajustar uma variável de interesse a um valor de referência especificado, sendo largamente utilizados no controle de processos industriais e na automação de manufatura, e têm importância fundamental para a produtividade industrial e segurança operacional. Apesar disso, a inexistência ou ineficiência de controles regulatórios em muitas operações industriais ainda é bastante comum, devendo ser superada com o uso adequado de Engenharia de Controle. Neste contexto, este artigo apresenta o desenvolvimento e implementação de uma estratégia de controle regulatório de vazão mássica em uma planta de processamento mineral. O controle efetivo de vazão mássica em plantas minerais ainda é muito negligenciado na prática, apesar de ser uma necessidade primária para uma operação produtiva e eficiente de tais plantas. A instalação de processamento mineral considerada neste trabalho, apesar de simples, trouxe alguns desafios para o desenvolvimento do controle regulatório, devido possuir uma dinâmica com longo atraso de transporte (tempo morto) e ganho variante no tempo. Para lidar com esses desafios, foi realizada uma modelagem do processo, seguida

pelo projeto de um controlador usando o método do Lugar Geométrico das Raízes (LGR), levando a excelentes resultados no desempenho regulatório. O artigo é organizado da seguinte forma: inicialmente são introduzidos o processo industrial e o problema de controle. Então, é descrito o desenvolvimento de um modelo dinâmico para o processo, com ênfase ao seu inerente atraso de transporte e ganho variante no tempo. Em seguida, é apresentada a estratégia de projeto do controlador, considerando o modelo desenvolvido para o processo, aplicando-se o método LGR para sistemas com atraso de transporte. Finalmente é apresentada a implementação do sistema de controle e os resultados obtidos. Guardadas as devidas proporções, a estratégia desenvolvida neste trabalho pode ser usada como uma estrutura geral de projeto de controladores para plantas minerais similares.

Palavras-chave — atraso de transporte, controle PID, controle regulatório, Lugar Geométrico das Raízes, processamento mineral, tempo morto.