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15th Triennial World Congress, Barcelona, Spain

LEVEL CONTROL STRATEGIES FOR FLOTATION CELLS

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Abstract: Flotation is a difficult process to run efficiently. One way to make flotation performance better is to improve cell level control. However, controlling pulp levels in flotation cells is a complex control task because of strong interactions between the levels in flotation cells. Therefore advanced controllers are needed to give good level control. This paper deals with a model of six flotation cells in series. Simulations are performed to compare different control strategies. Four control strategies are considered: one SISO controller and three different MIMO controllers. It is shown that level control performances of the MIMO controllers are significantly better than that of the classical SISO controller. *Copyright © 2001 IFAC*

Keywords: Decoupling, Multivariable control, Feedforward control, MIMO, SISO, Level control

1. INTRODUCTION

Flotation is one of the most commonly used mineral processing operations. The process is complicated and it is influenced by a large number of variables. One of most commonly controlled variables in a flotation cell is the slurry level. The level in the cell influences the concentration and yield of mineral and it is therefore important that the levels in the cells remain stable.

Level control of flotation cells is a very complex task due to high interactions between the process variables. A control action implemented at any point in the flotation circuit tends to be transmitted to both upstream and downstream units, and sometimes with amplification. Large variations in the flow rate to the first cell and varying composition of the raw ore also cause problems.

Flotation cells are conventionally controlled by isolated PI controllers. PI control works well when the cell being controlled is isolated. However, in a flotation circuit where interactions are strong, PI control does not meet the requirements of high control performance. Hence a considerable amount of research has been carried out over the last few years to develop better control techniques for flotation circuits.

The aim of this research is to study and compare different control strategies from the point of view of cell level control. Four different strategies are implemented and compared: one traditional SISO control strategy, and three MIMO control strategies. The strategies are compared by the means of special performance indices.

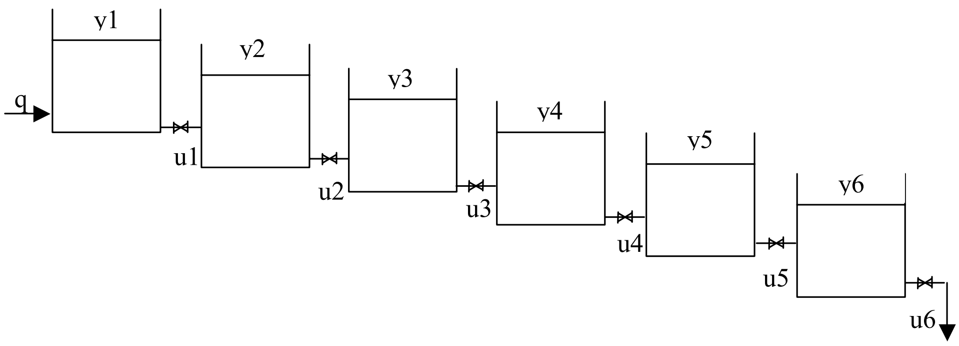

Fig. 1. Schematic diagram of flotation cells

2. MATHEMATICAL MODELING OF FLOTATION CELLS IN SERIES

In a flotation process several single cells are connected in series as shown in Fig. 1.

The change in volume of the pulp can be denoted for each cell in series using the following non-linear differential equations:

$$\frac{\partial y_1}{\partial t} = \frac{1}{L}\left(q - K_1 C_{v1} u_1 \sqrt{y_1 - y_2 + h_1}\right) \qquad (1)$$

$$\frac{\partial y_i}{\partial t} = \frac{1}{L}\left(K_{i-1} C_{v(i-1)} u_{i-1} \sqrt{y_{i-1} - y_i + h_{i-1}} - K_i C_{vi} u_i \sqrt{y_i - y_{i+1} + h_i}\right) \quad i = 2,..,5 \qquad (2)$$

$$\frac{\partial y_6}{\partial t} = \frac{1}{L}\left(K_5 C_{v5} u_5 \sqrt{y_5 - y_6 + h_5} - K_6 C_{v6} u_6 \sqrt{y_6 + h_6}\right) \qquad (3)$$

where
L = the area of the tanks
q = feed rate to the first cell
y_i = pulp level in the cell
h_i = physical difference in height between the cells
u_i = control signal
K = constant coefficient
C_v = valve coefficient

Eq. (1) is for the first cell in series, Eq. (2) for cells in the middle (i = 2,3,4 and 5) and Eq. 3 for the last cell in the series.

3. CONTROL STRATEGIES

The different control strategies are discussed and described in the following sections. These strategies are selected because they can be used with basic PI-controllers and without any additional instrumentation. Traditionally in flotation cell series there is only one flow measurement in the beginning of the series and level PI-controllers in every cell.

3.1 Feed forward controller

A flow feed-forward controller monitors disturbances in the inflow to the first cell and uses proportional action to close or open the valves of the cell in order to compensate for disturbances. Compensation is linearly dependent on the difference between the current inflow and the normal inflow. The measurement signal is filtered in order to prevent the feed-forward control from reacting to random variation in the flow. However, this kind of controller does not provide any extra performance improvement in the event of disturbances occurring somewhere else in the cell series. The model of the feed-forward controller is shown in Fig.2.

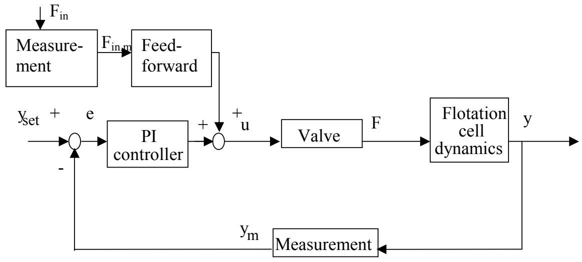

Fig. 2. Control diagram of feed forward controller.

3.2 Decoupling controller

A decoupling controller is based on differential equations (1), (2) and (3). The purpose of the decoupling controller is to eliminate the crosswise effects of control loops, and hence the stability of a single control circuit depends only on its own stability features. The basic model of the decoupling controller is shown in Fig. 3.

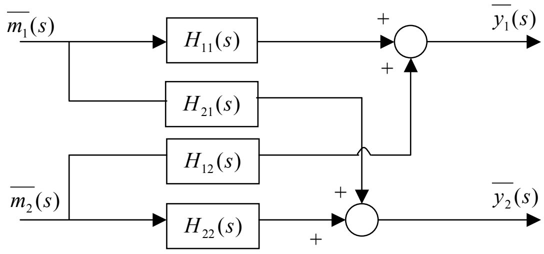

Fig. 3. Control diagram of a basic decoupling controller.

The mathematical criterion to be fulfilled for decoupling a tank *i* will be (Stenlund and Alexander, 2000)

$$\Delta F_{iin} - \Delta F_{iout} = 0 \qquad (4)$$

Where ΔF_{iin} is a change of inflow to tank *i.* Using the valve functions from Eq. (1), (2) and (3), the equation can be written as follows

$$F_i = K_i C_{vi} u_i \sqrt{(\Delta h_i)} \qquad (5)$$

Where Δh_i is the level difference over the valve. Substituting in Eq. (4), it becomes

$$K_{i-1} C_{i-1} (u_{i-1} + \Delta u_{i-1}) \sqrt{(\Delta h_{i-1} + \Delta(\Delta h_{i-1}))} - K_i C_i (u_i + \Delta u_i) \sqrt{(\Delta h_i + \Delta(\Delta h_i))} = 0 \qquad (6)$$

Solving this equation for the change in the control signal gives

$$\Delta u_i = \left(K_{i-1}C_{i-1}/K_iC_i\right)u'_{i-1}\sqrt{\left(\Delta h'_{i-1}\right)/\left(\Delta h'_i\right)} - u_i$$
$$= f\left(u'_{i-1}, \Delta h'_{i-1}, \Delta h'_i\right) \quad (7)$$

Eventually, the control signal for a tank i becomes

$$u_i = u_{PI} + f\left(u_{i-1}, \Delta h_{i-1}, \Delta h_i\right) \quad (8)$$

Where u_{PI} is the control signal from a PI controller. In order to handle the variations from the inflow, the feed forward is attached to the first tank.

3.3 Multivariable controller similar to Floatstar™

A multivariable controller (Schubert *et al.,* 1995) controls the total inventory of material in the upstream tanks. In this control strategy, controlling a valve is influenced not only by the difference between a set point and the measured level in the tank, but also the differences between set points and the measured levels in all the tanks in upstream. These variables are summed and fed to the PI controller of the cell. Furthermore, the variables are scaled by a suitable factor depending on the valve size, position and process.

In this strategy each control valve can be regarded as a sluice gate of a dam. When a damned inventory is too high in upstream, the valves are opened more than usual, even when there seems to be no need to take such an action on the basis of the levels in the neighbouring vessels. The control diagram is shown in Fig. 4.

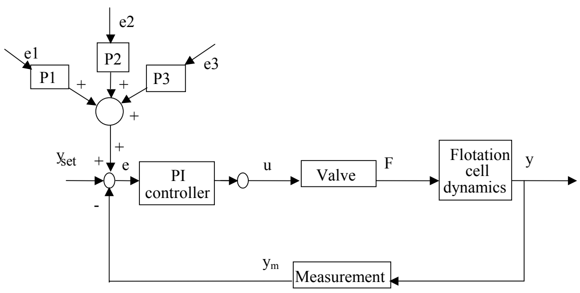

Fig. 4. Control diagram of a multivariable controller similar to Floatstar™ (tank 4).

3.4 Feed-forward multivariable controller

In the previously described controller, the error signals of the upstream cells are fed to the PI controller. In this strategy the error signals are added directly to the control signal of the valve. All error signals from the upstream cell are filtered in order to prevent the valve acting too rapidly to disturbances. In this technique the PI controller controls only its own level of the cell. Flow feed forward from the feed to the first tank has been added to every cell in order to handle the inflow variations from. The control diagram is shown in Fig 5.

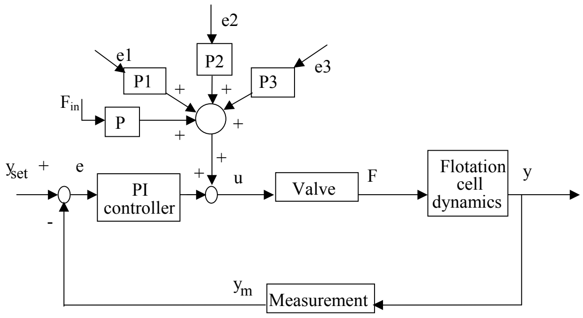

Fig. 5. Control diagram of the feed-forward multivariable controller (tank 4).

4. SIMULATIONS

In the simulations a configuration of six TC-50 cells in series was studied in accordance with the ideal tank assumption. Therefore the effects of boosters and launders were not considered. The valves were 100% oversized according to the ISA standard, and the retention time in each cell was 1.5 minutes. Control strategies included conventional PI controllers with feed-forward control, decoupling controller, a multivariable controller similar to Floatstar™ and a feed-forward multivariable controller. The simulation results of a + 3 cm change in the set points of the cell levels at times 100, 150, 200, 250, 300 and 350s are presented in the following. Making +/- 20% change in the feed to the first flotation cell was also simulated with different strategies. The set point of the cell level is lowest in the first cell, and the set point values increase on moving towards the last cell in the series, where the operating range of the level controller is smaller. The simulation schemes were constructed with Matlab 6.0.0 Simulink software.

The controllers were tuned and compared using the following indices. The IAE index (integral of the absolute value of the error) integrates the absolute value of errors, and even-handedly weights all the deviations. ISE (integral of the square error) gives more weight to big deviations from the set point.

$$ISE = \int_{t=t_1}^{t_2} \left(y(t) - y_{sp}(t)\right)^2 dt \quad (9)$$

$$IAE = \int_{t=t_1}^{t_2} \left|y(t) - y_{sp}(t)\right| dt \quad (10)$$

5. SIMULATION RESULTS

The simulations of the configurations of six TC-50 cells in series resulted in parameters for the PI controllers. Integration times in the traditional system with a feed-forward controller were between 15 – 50s and proportional gains between 0.8-1.2. Because MIMO control strategies respond better to disturbances, the PI parameters were set faster. Integration times in all the PI controllers were set to 15s and gain to 1. In the decoupling controller the PI-parameters were between 15-50s and 1-1.4, correspondingly.

The responses of the feed-forward controller to disturbances in pulp feed and to set point changes are presented in Fig. 6. As can be seen, the –20% change in pulp feed is affecting all the cell levels in the series. The set point changes in a cell also have undesirable effects on the adjacent cells. There is always a considerably large perturbation in the level of the next cell every time a set point change is made in the system.

The responses of the decoupling controller are illustrated in Fig 7. The decoupling controller is a MIMO controller, and it also takes into account the interactions between cells. As can been seen from the graphs, the decoupling controller effectively eliminates disturbances arising from changes in the pulp feed. Furthermore, set point changes in the cells do not affect to the other cells.

The responses of configurations in which a controller similar to Floatstar™ and the feed-forward multivariable controller are used are shown in Figs. 8 and 9. The controller similar to Floatstar™ seems to be slightly more robust than the other controller, especially during set point changes.

The IAE and ISE indices, which depict the performance of controllers, are shown in Tables 1, 2, 3 and 4. As was to be expected, the traditional SISO control with flow feed forward had the poorest figures in all cases.

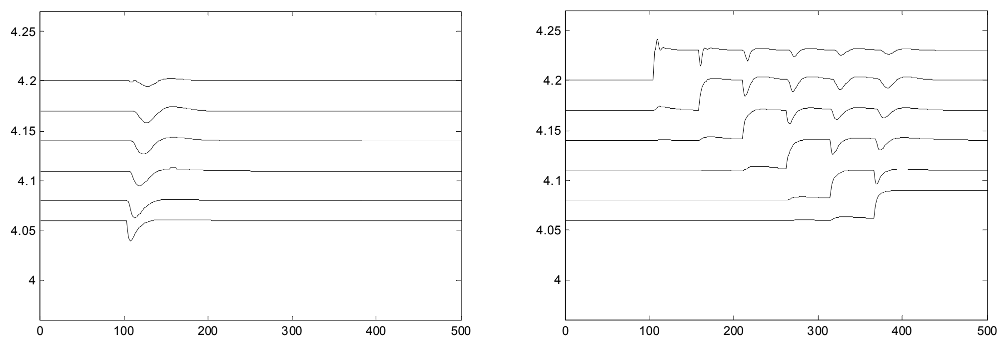

Fig. 6. Feed-forward controller. On the left the response to a -20% change in pulp feed, and on the right the response to set point changes.

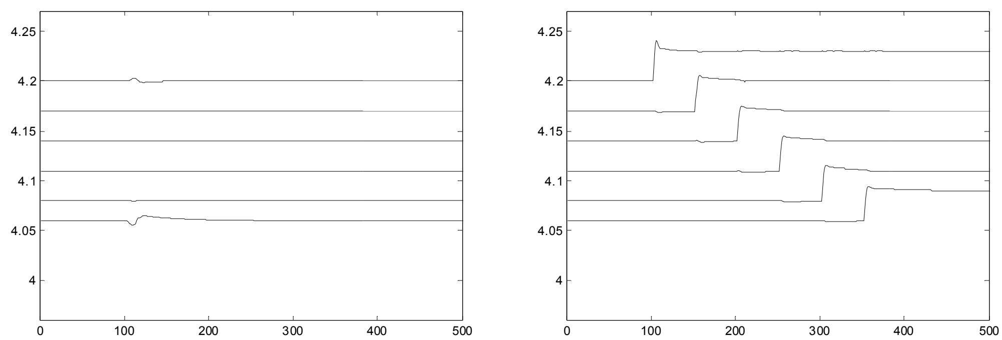

Fig. 7. Decoupling controller. On the left the response to a -20% change in pulp feed, and on the right the response to set point changes.

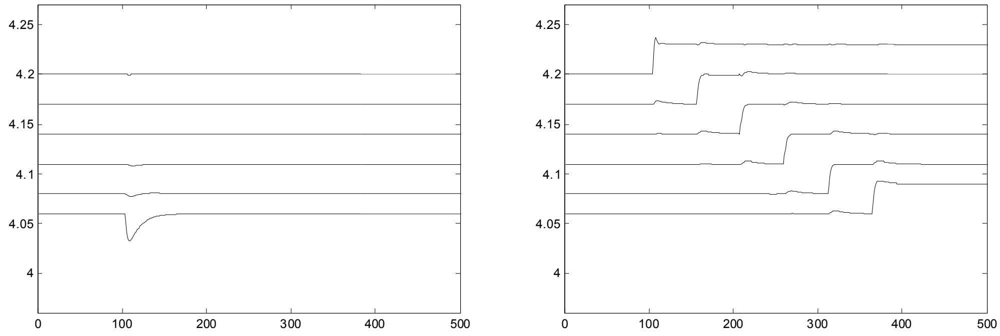

Fig. 8. Multivariable controller similar to Floatstar™. On the left the response to a -20% change in pulp feed, and on the right the response to set point changes.

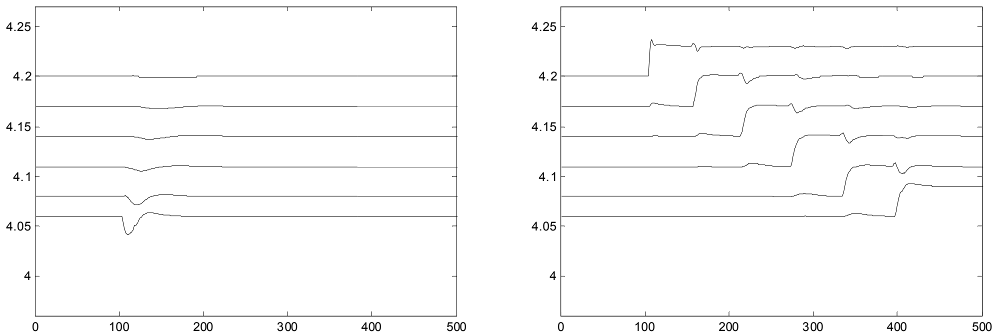

Fig. 9. Feed-forward multivariable controller. On the left the response to a -20% change in pulp feed, and on the right the response to set point changes.

Table 1. The performance indices for the feed-forward controller

Feed-forward controller	ISE(+20%)	IAE(+20%)	ISE(-20%)	IAE(-20%)	ISE(s.p.c.)	IAE(s.p.c.)
1.	3.1	348	3.7	325	2.7	298
2.	2.9	361	3.1	360	4.0	520
3.	2.7	371	3.3	391	5.0	712
4.	2.2	341	2.7	371	5.33	818
5.	2.0	322	2.4	361	6.5	970
6.	0.5	159	0.4	147	3.7	604

Table 2. The performance indices for the decoupling controller

Decoupling controller	ISE(+20%)	IAE(+20%)	ISE(-20%)	IAE(-20%)	ISE(s.p.c.)	IAE(s.p.c.)
1.	1.1	327	0.7	263	2.0	220
2.	0.0008	10.2	0.0003	8.0	2.2	242
3.	0.0000008	0.3	0.00000004	0.1	2.2	241
4.	0.00000005	0.05	0.000000001	0.01	2.2	241
5.	0.00003	1.2	0.000004	0.4	2.1	228
6.	0.04	29.5	0.019	19.5	2.1	179

Table 3. The performance indices for the multivariable controller similar to Floatstar™

MV controller	ISE(+20%)	IAE(+20%)	ISE(-20%)	IAE(-20%)	ISE(s.p.c.)	IAE(s.p.c.)
1.	7.81	499	9.02	527	2.2	202
2.	0.1	65.0	0.07	58.6	2.3	243
3.	0.03	25.2	0.03	26.5	2.3	248
4.	0.0001	2.2	0.00006	1.4	2.3	249
5.	0.0001	2.7	0.00008	1.5	2.3	245
6.	0.01	18.5	0.002	7.8	1.7	151

Table 4. The performance indices for the feed-forward multivariable controller

Feed-forward MV controller	ISE(+20%)	IAE(+20%)	ISE(-20%)	IAE(-20%)	ISE(s.p.c.)	IAE(s.p.c.)
1.	1.9	226	2.2	237	2.6	224
2.	0.6	159	0.8	173	2.8	313
3.	0.3	120	0.3	133	2.9	357
4.	0.2	105	0.2	119	2.9	389
5.	0.09	92	0.2	107	2.8	389
6.	0.03	55	0.03	53	1.9	241

6. CONCLUSIONS

All the simulated configurations were successfully tuned. It is noticeable that the classical SISO strategy with feed forward controller cannot even approach the performances of the MIMO controllers. This is due to high interactions between the control loops, which SISO systems cannot take into account.

The differences between different MIMO systems are somewhat smaller. All the controllers performed robustly to disturbances in pulp feed and to set point changes. The decoupling controller had the best IAE and IDE indices. However, the decoupling controller is sensitive to model uncertainties (Skogestad and Postelwaite, 1996). This also means that process changes can strongly degrade the control performance.

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