

DYNAMIC MODELING OF AN INDUSTRIAL COPPER SOLVENT EXTRACTION PROCESS

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Abstract: The aim of this study is to examine the ability of first principles models to describe the behavior of an industrial copper solvent extraction plant. First, the process is introduced and the dynamical models are formulated. Then, the testing environment is described and comparisons of the simulated and real data are presented and discussed.

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Keywords: Copper solvent extraction, dynamic simulation, first principles modeling, mixer-settler.

1. INTRODUCTION

During the last two decades the copper leaching, solvent-extraction and electrowinning (LX/SX/EW) process has become an important option in the production of copper from low-grade oxidized ore. LX/SX/EW copper production has grown rapidly during the last decade and further growth can be expected. (Jenkins et al., 1999; Kordosky, 2002)

As the copper LX/SX/EW process is relatively young and intrinsically rather stable, development has so far mainly focused on the process equipment. This has led to a low level of automation compared to other mineral and metal processing operations. Improvements in plant control would make it possible to keep process variables close to their optimal values, thus increasing the amount of copper produced, reducing the amount of chemicals and energy consumed, and improving cathode copper quality. These improvements would lead to considerable economic benefits. (Jämsä-Jounela, 2001; Bergh et al., 2001; Hughes and Saloheimo, 2000)

The state of the art in the control of hydrometallurgical copper processes has been reviewed by Bergh et al. (2001). According to the review, no studies or application of multivariable regulatory control, adaptive, non-linear and robust control theory to the copper LX/SX/EW processes have been reported in the literature. Process control has mainly relied on basic control loops.

The motivation of the research is to develop a control system for an industrial copper solvent extraction process. Since the copper solvent extraction process has similar characteristics to those processes in which model predictive control has been successfully

applied, the first aim is to develop dynamical process models.

A phenomenological steady state simulation model for a copper SX/EW pilot plant has been developed by Aminian et al. (2000). The simulation model uses variables that are not measured online in industrial process operations.

Therefore, the modeling approach utilized here is based on dynamic first principles models for liquid-liquid extraction utilizing plant specific McCabe-Thiele diagrams and available online plant measurements. The first principles modeling adopted here, concerns only the measured and the most relevant phenomenon of the solvent extraction process, mass transfer.

2. PROCESS DESCRIPTION

The aim of copper solvent extraction is to concentrate and purify aqueous copper solutions. The process consists of extraction and stripping processes, both of which may contain parallel and series subprocesses. The process has two input flows, pregnant leach solution (PLS) and lean electrolyte (LE), and one recycling flow, the organic solution. In the extraction stage, the copper is extracted from PLS to barren organic (BO) solution. In the stripping stage, copper is stripped from the loaded organic (LO) to lean electrolyte (LE) solution. The result of stripping, i.e. rich electrolyte (RE), is blended and fed to the electrowinning process, where 99,99% pure copper cathodes are produced. A flow diagram of the process is shown in Figure 1.

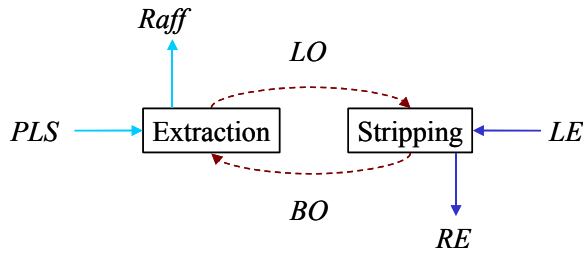


Fig. 1. General flow diagram of the copper solvent extraction process.

The copper transfer between the aqueous and organic phases takes place in mixers and the phases are separated in settlers. The equipment used for the solvent extraction process varies considerably. For example, in Outokumpu VSF technology the organic and aqueous flows are combined in a dispersion pumping unit followed by two mixers and a large, shallow settler. (Nyman et. al. 2003)

Typical process instrumentation includes flow rate, temperature, and level measurements. There may also be online measurements of conductivity in the mixers, pH, and copper and impurity,. The copper and impurity assays, and phase ratio, breaking time and other diagnostic measurements are made by the process operators and the laboratory.

3. DYNAMIC MODELING

The dynamic modeling is based on ideal mixing and plug flow assumptions, and requires flow rate and online copper concentration measurements, as well as offline settler and process tank level measurements. The plant isotherms, required to calculate the copper transfer in the mixing, are assumed to be constant during the operation.

The mixer is modeled as an ideal mixer, followed by the settler. The settler model consists of parallel plug flow and ideal mixer models separate for both phases, as shown in Figure 2. The process tanks are modeled as plug flow due to the laminar flow conditions.

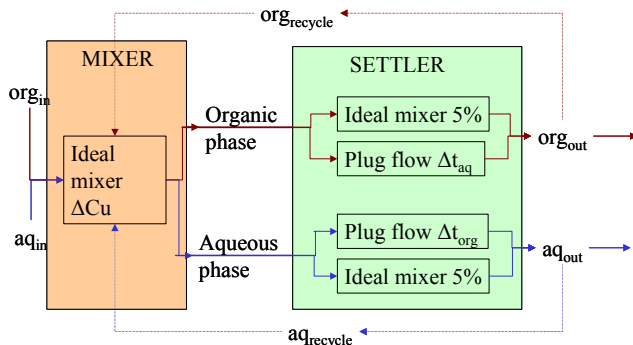


Fig. 2. Model for one mixer-settler pair

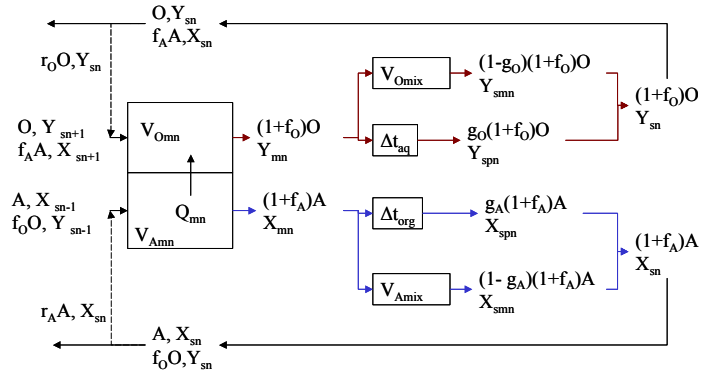


Fig.3. Detailed model of the one mixer-settler unit.

3.1. Mixer model

Since the mixer is assumed to be an ideal mixer, the concentration reaches equilibrium instantly and is constant in all parts of the mixer. The concentration X_{mn} of the aqueous volume $V_{A_{mn}}$ in the mixer changes over time according to the incoming copper X_{sn-1} from the aqueous flow A_{sn-1} , aqueous entrainment $f_A A_{sn+1}$ in the organic, and recycled aqueous phase $r_{A_{sn}} A_{sn}$, outgoing copper X_{mn} in aqueous flow A_{mn} , and copper transfer Q_{mn} to the organic phase. Entrainment in the recycled flow is neglected. The copper mass transfer in the aqueous phase of the mixer is defined by:

$$V_{A_{mn}} \frac{dX_{mn}}{dt} = (1 - f_A - r_{A_{sn-1}}) A_{sn-1} \cdot X_{sn-1} + f_A A_{sn+1} \cdot X_{sn+1} + r_{A_{sn}} A_{sn} \cdot X_{sn} - A_{mn} \cdot X_{mn} - Q_{mn} \quad (1)$$

$$Q_{mn} = K(X_{mn} - X_{mn}^*) V_{A_{mn}} = K(X_{mn} - f(X_{in}, Y_{in}, \alpha)) V_{A_{mn}} \quad (2)$$

$$A_{mn} = A_{sn-1} + f_A A_{sn+1} + r_A A_{sn} \quad (3)$$

The detailed flow sheet of the one mixer-settler unit is shown in Figure 3.

3.2. The settler model

The settler model consists of plug flow and mixer parts. Most of the liquid flow in the settlers is modeled as plug flow, in which only the time delay caused by flow is considered. A minor part of the liquid flow $(1 - g_A, 1 - g_O)$ is modeled with the ideal mixer, representing the back-mixing and other mixing phenomena in the settlers. The volumes of the mixing sections are $(1 - g_{VA}, 1 - g_{OA})$ times the total volume of the settler.

The copper concentration through the plug flow region is the same as that in the outflow from the mixer. The plug flow is defined for the aqueous phase:

$$X_{spn} = X_{mn}(t - \Delta t_{aq}) \quad (4)$$

, where the time delay is computed by the settler area, aqueous depth, plug flow fraction and aqueous flow:

$$\Delta t_{aq} = \frac{a_{settler} \cdot h_{settler,aq} \cdot g_{VA}}{A} \quad (5)$$

The mixing parts in the settlers are separate for both organic and aqueous flows. The change in the copper concentration is modeled as the difference between the incoming and outgoing concentrations. The model equations of the mixed parts of the aqueous and organic flows are defined:

$$V_{Asmix} \frac{dX_{smn}}{dt} = A_{mn}(1 - g_A)(X_{mn} - X_{smn}) \quad (6)$$

, where

$$V_{Asmix} = a_{settler} \cdot h_{settler,aq} \cdot (1 - g_{VA}) \quad (7)$$

3.3. Copper transfer model

In the ideal mixing model the copper transfer rate Q is calculated on the basis of the plant isotherm and is defined by:

$$Q = K(X - X^*)V \quad (8)$$

where V is the mixing volume and K is the mass transfer (yield) coefficient for the aqueous phase. Ingham et. al. (1994) suggested $K = 25$ m/s. The equilibrium concentration X^* is defined by:

$$X^* = f_{eq,oper}(Y, \alpha) \quad (9)$$

, where $f_{eq,oper}(Y, \alpha)$ is the optimal extraction rate, and α is an efficiency coefficient of the extraction.

The optimal extraction / stripping rate is calculated using the McCabe-Thiele step diagram presented in Figure 4. The isotherm curves for extraction and stripping are assumed to be constant, whereas the operating lines are determined by the volume ratio between the organic and aqueous phases in the mixers. The number of steps is determined by the number of mixer-settler pairs. For example, in Figure 4. there are two mixer-settler pairs in extraction and one in stripping.

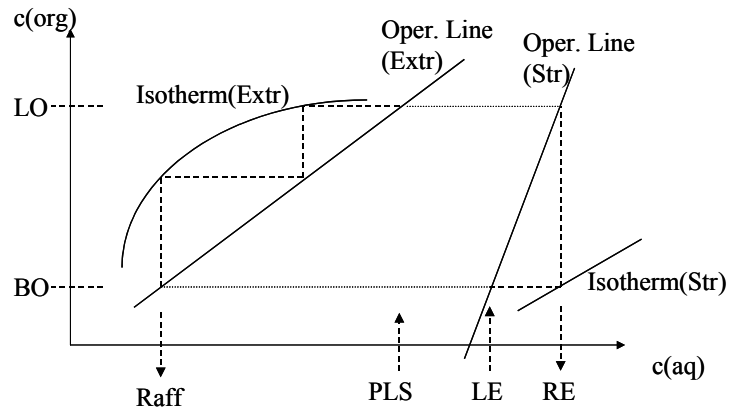


Fig. 4. The McCabe-Thiele step diagram.

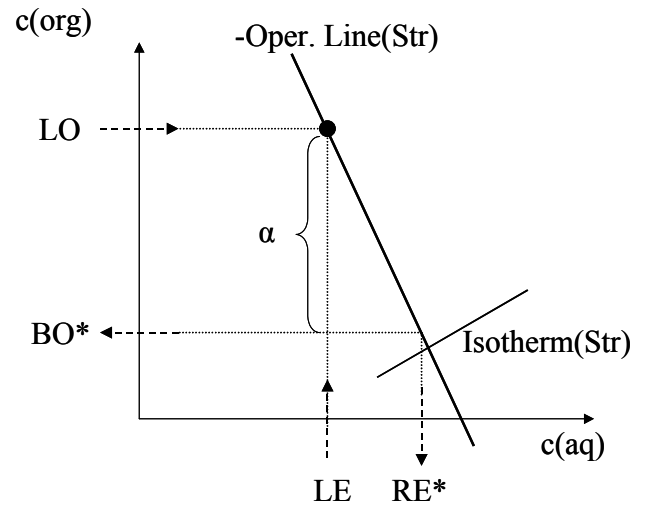


Fig. 5. Calculation of output concentrations (RE, BO) from the step diagram with efficiency coefficient α .

The output copper concentrations of one stripping unit are estimated in Figure 5 as follows: The concentrations of rich electrolyte (RE) and barren organic (BO) are determined by drawing an inverted operating line from the point (Lean electrolyte, loaded organic) towards the stripping isotherm. The point where this line and isotherm intersect is the optimal value for rich electrolyte and barren organic. The rich electrolyte concentration is calculated from the horizontal axis and the barren organic from the vertical axis.

The efficiency coefficient, which has values between 0 and 1, is taken into account by drawing the line to the length determined by the coefficient, as illustrated in Figure 5. For several steps the output values are solved iteratively.

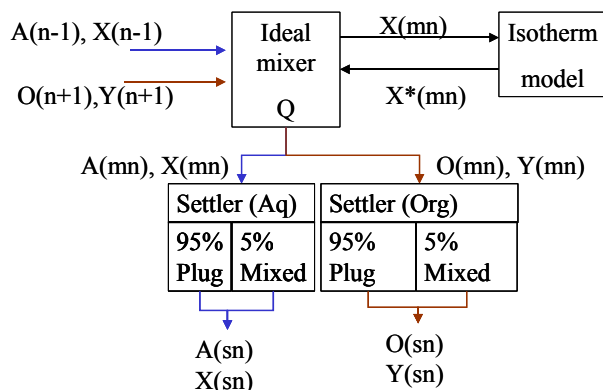


Fig. 6. Implementation of the one mixer-settler pair.

4. SIMULATION MODEL

A dynamical simulator of the process was constructed on the basis of first principles models. The aim of the simulator is to enable study of the dynamical behaviour of the solvent extraction process and, in the future, to provide a test bench for the control system. The unit operation models were implemented in Matlab Simulink according to the plant configuration. The Simulink model for a mixer-settler pair consists of three different types of sub-block, mixer model, isotherm model and settler model, as shown in Figure 6. The measurements used for the simulation were flow rates, copper concentrations, organic phase depths in the settlers, and heights of the process tanks. The constant parameters, including the isotherm coefficients, efficiencies, backmixing, entrainment and recycle rates, were chosen according to plant information.

5. TEST RESULTS

The simulator was tested for one month. The simulated outputs, the primary outputs: copper concentrations of barren and lean organic, and rich electrolyte, and the secondary outputs: the raffinates were compared to real process measurements. The mean and standard deviation of the residuals, standard deviation of the residual divided by the mean value of the variable, and the standard prediction error of the whole test period, are presented in Table 1.

Table 1: Residuals for the process outputs

	Mean	Variance	Res. %
Barren Organic	0.05	0.28	6.23 %
Loaded Organic	0.30	0.30	3.06 %
Rich Electrolyte	0.80	0.67	1.50 %
RaffinateP	-0.05	0.06	15.81 %
RaffinateS	-0.02	0.07	16.28 %

The mean of the residuals is close to zero for all the variables and the standard deviation is well below 10% of the signal for the primary variables, the organic and electrolyte measurements. The real raffinate measurements are less exact and this concludes in larger variation between the simulated and measured raffinate concentrations.

The input copper concentrations, PLS and lean electrolyte, are shown in Figure 6, and the simulated and measured output data are compared in Figures 8 – 9. The simulated data follows the process trends especially good for rich electrolyte and the raffinates.

The analyzer disturbances cause two peaks to the results; An analyzer failure at 1050, and a calibration change at 2000 caused peaks for all the measured variables.

The reason for deviation between the simulated and measured variables is the assumption of constant efficiency in extraction and stripping units. In real plant the efficiencies are changing due to changes in reagent concentration of the organic solution, pH of the aqueous solutions and major temperature changes. For example during 500 – 1000 the simulated raffinate parallel level is higher and the simulated loaded organic level lower than the measurements, which indicates that the extraction efficiency for the parallel unit should have been changing during this period. During 1000 – 2000 the efficiencies between series and parallel have changed, which resulted in the simulated values being lower than measured for raffinate series and higher for the raffinate parallel.

The model with constant efficiencies is able to capture major process trends, however the modelling accuracy would be improved by introducing varying efficiency parameter for each of the extraction and stripping units.

6. CONCLUSIONS

A dynamic process simulator utilizing mass transfer models based on McCabe-Thiele diagram and constant extraction and stripping efficiencies, was constructed. The simulator was successfully tested with industrial data. The simulated data followed the real process measurement trends during the one month testing period.

In the future the modeling accuracy will be improved by introducing varying efficiencies for each of the extraction and stripping unit models. The goal of the modeling is to develop an accurate enough dynamic simulator that can be used in the future control studies of the process.

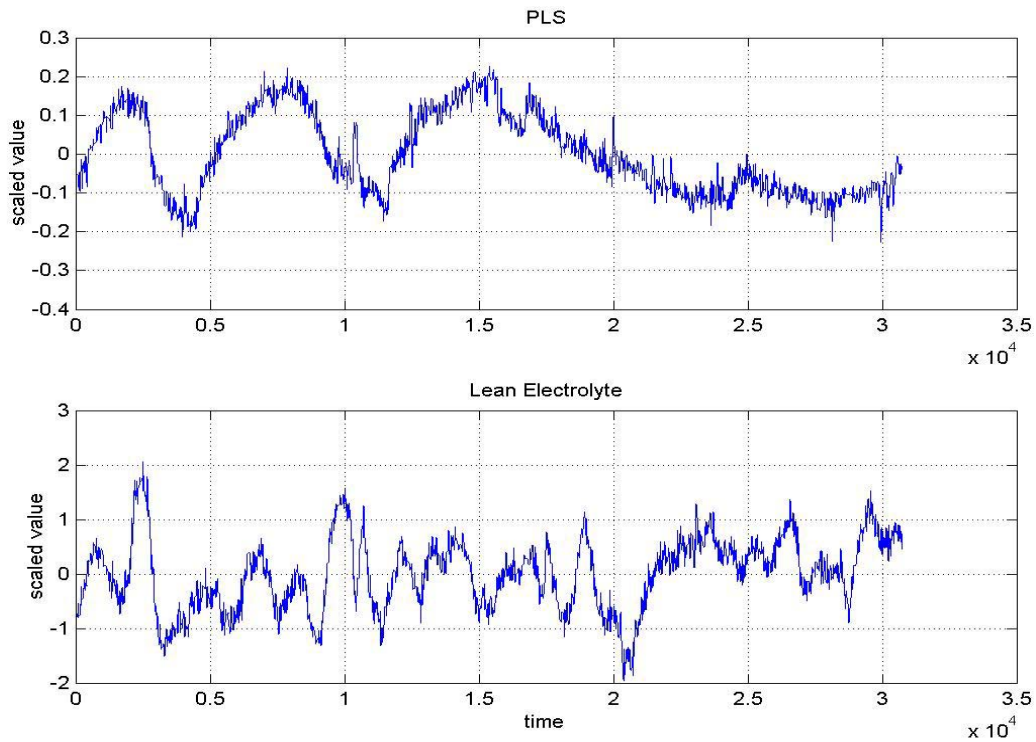


Fig. 7. PLS and lean electrolyte copper concentrations.

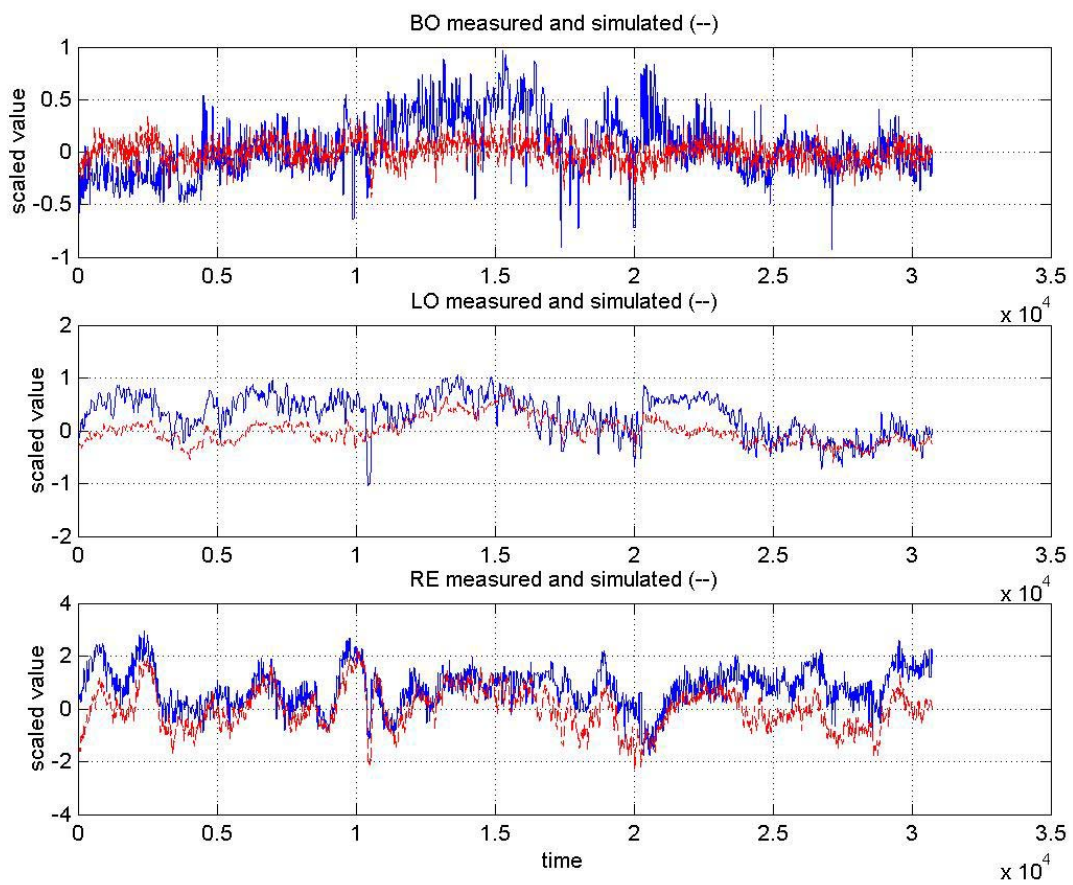


Fig. 8. Measured (blue) and simulated (red, dashed) copper concentrations of barren organic, loaded organic and rich electrolyte.

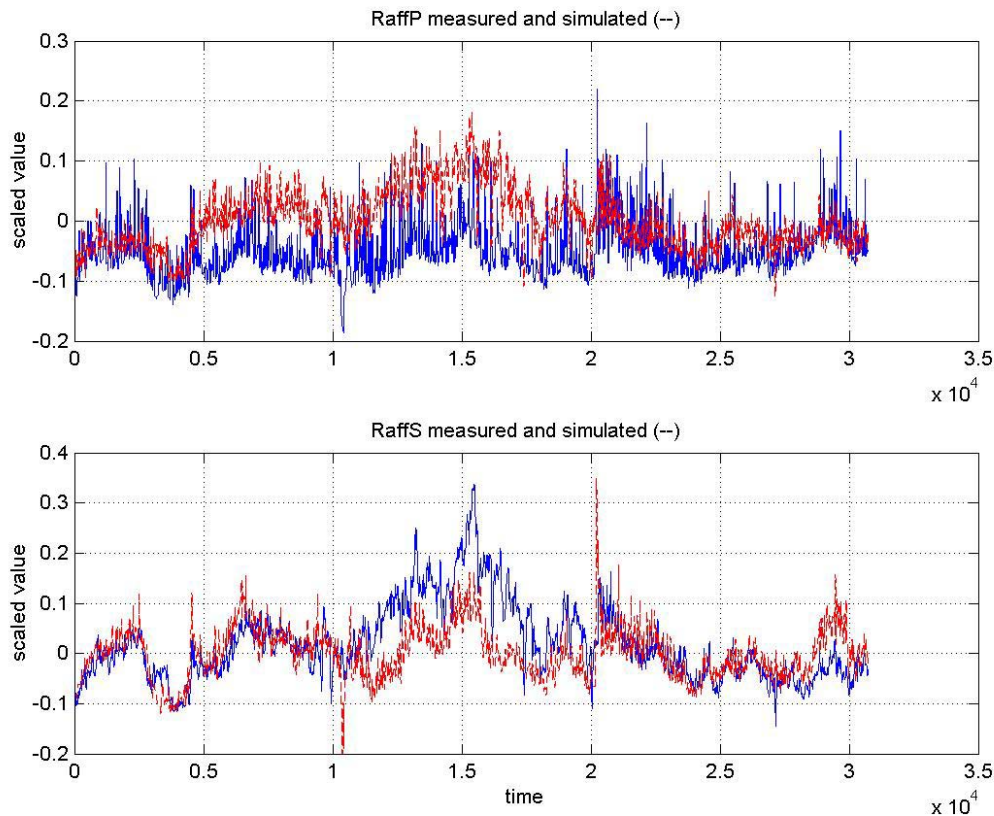


Fig. 9. Measured (blue) and simulated (red, dashed) copper concentrations for raffinate parallel and series.

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