Precise Ratio Control Structure for Nonlinear Blend Processes

Ciprian Lupu, Dumitru Popescu, Robert Gyorodi, and Andreea Udrea

Abstract — **This paper proposes a precise structure solution for the ratio control of nonlinear processes blend stations. The main feature of the proposed structure is to give a good ratio performance for both set-point changes and load disturbances that might occur in any one of the loops. For each loops (two or more) a nonlinear structure based on a multiple model and nonlinearity compensator is proposed. The main advantages consist in a classic tuning procedure for each of the controllers and a very simple procedure for nonlinearity determination, suitable for industrial applications. All components of control structure ware implemented on real time control application.**

I. INTRODUCTION

EFFICIENT control for real systems and processes was and continues to be, since the last decades, one of the targets of automatic control. A direction that gained major importance is the study and compensation of the nonlinear phenomena that characterizes the surrounding real world.

In industrial practice, it is common to find some specific control structures for nonlinear processes in order to lead, if possible, the design techniques towards classic control approaches [1], [2]. There are a lot of solutions for this problem; some are based on robust design, partial linearization or segment linearization (multiple model - MM) etc. [4] - [6].

Coming back to ratio control (RC), which consists in keeping a constant ratio between two (or more) process variables, irrespective of possible set-point changes, and load disturbances that might occur on the plant, is of concern in a variety of industrial applications such as chemical dosing, water treatment, chlorination, mixing vessels [2] and waste incinerators. For example, in combustion systems it is necessary to control accurately the air-to-fuel ratio in order to obtain a high efficiency, and in blending processes a selected ratio of different flows has to be maintained to keep a constant product composition.

In industrial mixing case, the existence of a class of processes with nonlinearities generated by natural forces like gravity must be observed, (filling-emptying with self stabilization) process [8], increased by storage recipients and installations specificity which lead to nonlinear approach for each contained loop.

For increasing performances (disturbance rejection and set point tracking) on larger domains, the classical real-time applicable solutions that exist in literature [6], [11] are based on adaptive structures, robust or multi-model structures. These, in general, are difficult to generalize and pretentious from the hardware and software point of view.

In the following chapters, we propose a precise control solution for ratio control where each controller has a multiple model or nonlinearity compensator structure.

Combinations of these three structures (mixing, multiple model and nonlinearity compensator) are not so simple, and some specific problems like controllers switching bump effect – common for multiple model structures - as possible disturbance for mixing structure, require special attention.

The solution's applicability is demonstrated by means of real time laboratory experimental installation.

II. PROPOSED CONTROL STRUCTURE

A. Ratio control blend processes

Starting from Visioli's proposed ratio control structures [2] and other efficient variant [7], in this paper a generalized ratio control structure that involves two blend stations is used (Fig. 2):

Here, formally denote by ***a*** the desired ratio to be kept between the values of two process variables y_1 and y_2. For this purpose, the control schemes shown in Figure 2 and Figure 1 (also termed *series metered control - for F_1 "=1" and F_2 "=0"*) can be implemented. Each variable is controlled by two separate controllers C_1 and C_2 (typically of PI, PID type) and the output y_1 of the first process is multiplied by a and adopted as the set-point signal of the closed-loop control system of the second process, *i.e.*, it is $r_2(t) = ay_1(t)$.

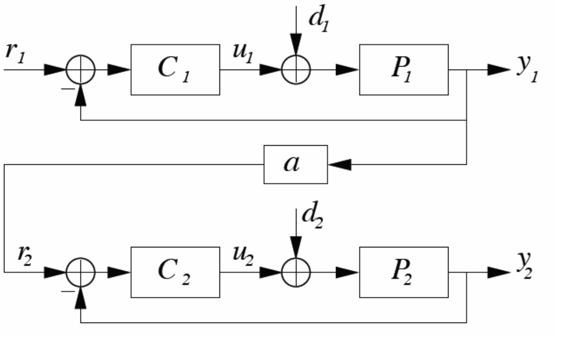

Fig. 1. The typical ratio control scheme (series metered control) [2]

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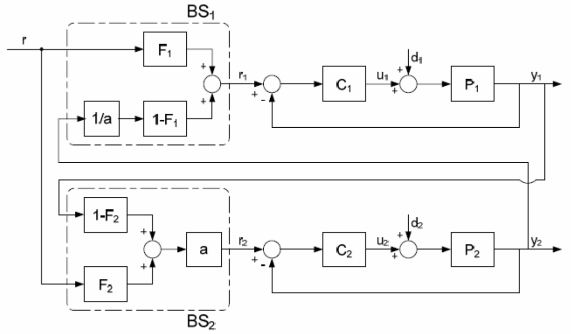


Fig. 2. The proposed generalized ratio control structures involving two blend station [7]

In Fig. 2 BS1, BS2 are “Blend Station” (BS) [3], F_1 and F_2 - coupling factors are dynamic systems $F_1(s)$ and $F_2(s)$, associated here to filtering components.

For this approach F_1 and F_2 are considered constant [7] included in (1.0 – 0.0) range. These two factors represent dependence of each system’ set point to neighborhood’s output. A lower value means important coupling (dependence). In the same time, for particular values $F_1 = 1$ and $F_2 = 0$ system presented in Figure 2 becomes a corresponding system from Figure 1 (classic series connection).

The main advantage of this structure is to keep the ratio in a predefined value, but the other objectives of the control are to assure a good set-point following and disturbance rejection performance. Changes, even small, as the output of one of the systems affected by a perturbation leads to corresponding modification of the reference system so that other report to be maintained.

Another interesting observation is related to "synchronize" the dynamics of evolution regarding the references of the two systems, as systems "expect each other". Following this "synchronization" performance in references tracking is affected, having been a slowdown in dynamics. This can be countered with a redesign of tracking performance of the real time control loop.

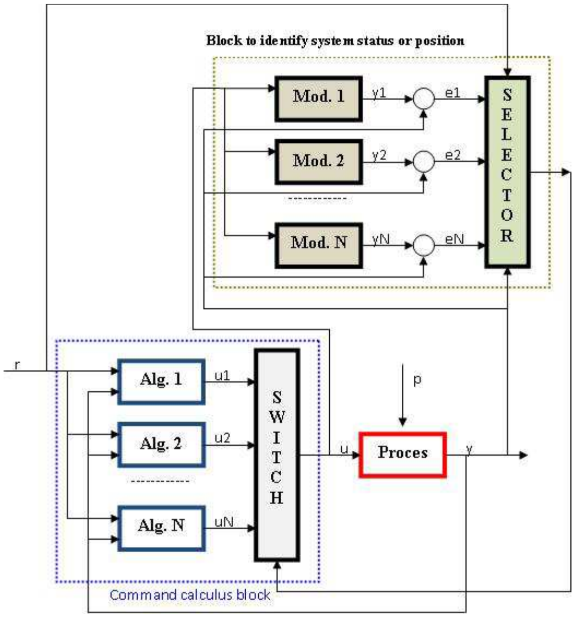


Fig. 3. Classic multiple model control structure

B. Multiple model

The applicability of the multiple model (MM) solutions is well known and demonstrated on real time nonlinear industrial installation [4], [5].

The proposed structure is based on the classical MM control strategy (Fig. 3.). The main idea of the multi model construction structure is based on dividing the process functioning region in *n* small disjoint and adjacent zones, for which the models are simpler and the *n* corresponding control algorithms have low complexity.

$$M = \{\mathcal{M}_1, \mathcal{M}_2, \mathcal{M}_3 \ldots \mathcal{M}_n\}, \quad C = \{C_1, C_2, C_3 \ldots C_n\}$$

In Fig. 3 the blocks and signals are:

- Process – physical system to be controlled;
- Command calculus – unit that computes the process control law;
- System’s state or position identification– component that provide information about the model–control algorithm “best” matching for the actual system’s state;
- Mod. 1, Mod. 2, …, Mod. N – previously identified models of different regimes or operating points;
- Alg. 1, Alg. 2, …, Alg. N – control algorithms designed for the N models mentioned above;
- SWITCH – mixing or switching between the control laws;
- SELECTOR – based on adequate criteria evaluations, provides information about the most appropriate model for the system’s current state;
- y and y1, y2, …, yN – output of the process and outputs of the models, respectively;
- output generated by Command calculus block;
- u and u1, u2, uN – output of the Command calculus block and outputs of the N control algorithms, respectively;
- r – system’s set point or reference trajectory;
- p – disturbances of physical process.-

Some specific problems of the MM control structures are: select the most adequate model/controller pair and bumpless algorithms switching. Usually, the selector block (System’s state or position identification block) contains the process’ models.

In the case of tank level, the section is based on level measure – process output - or set point value. This corresponds to a specific, prefixed interval. Number of intervals corresponds to models/controllers pair’s numbers.

Switching solution is solved based on [10] proposed solution. The main idea is that, during the current functioning of MM control systems with N model-algorithm pairs, it is supposed that just one single algorithm to be maintained active, the good one, and all the other N-1 algorithms rest inactive. The active and inactive states represent automatic, respectively manual, regimes of a law control. The output value of the active algorithm corresponds to the manual control for all the other N-1 inactive algorithms. In the switching situation, when a "better" *Aj* algorithm is found, the actual *Ai* active algorithm is commuted in an inactive state (manual), and *Aj* in active state (automatic), respectively.

For a bumpless commutation, the manual–automatic transfer problems must be solved. Used solution [10] proposes the computation of that set point value that determines, according to the algorithm history and process output, a control equal to the manual command applied by the operator (or active algorithm).

Even if there are bumps, due to algorithms switching, they have no effect on the *a* ratio as coupling factor loops leads to a corresponding correction of the set point to the other system.

C. *Nonlinearity compensator*

Nonlinearity compensation can be realized by using a nonlinearity compensator structure in series or parallel with the classical control algorithm. For better performances the parallel structure is used [9], [11], leading to an architecture based on a classical feedforward – feedback structure presented in figure 4.

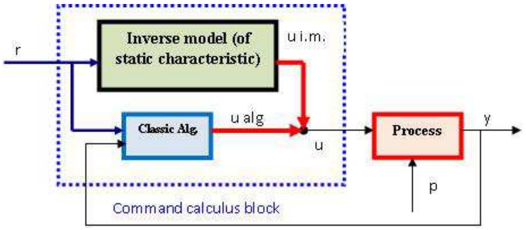


Fig. 4 Hybrid feedforward - feedback control structure

The feedforward component represents the static geometric inverted model, or, when possible, the dynamic one.

The feedback component has the role of correcting the identification differences of the static characteristics and the dynamic comportment differences.

The determination of the geometrical characteristics of a nonlinear process is based on several experiments of discrete step increasing and decreasing of the command applied to the process - $u(k)$ and measuring the corresponding stabilized process output - $y(k)$. The command $u(k)$ covers the entire admissible domain 0 - 100% (in percentage representation).

Obviously, the characteristics are not identical because the process is disturbed by noises. The final static characteristic is obtained by meaning the correspondent positions of the corresponding experiments.

Figure 5 presents the way a characteristic can be modeled. The graphic between two "mean" points is obtained by using an extrapolation procedure.

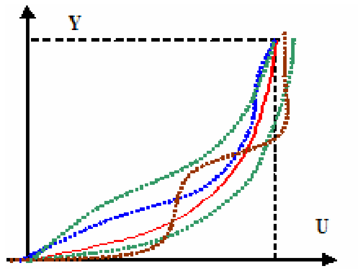


Fig. 5 The static characteristic for a branch; the continuous line represents the geometric characteristic

The maximal "dispersion" of process trajectory expresses the influence of the superposing noise that action onto process, the process's nonlinearity and modeling incertitude. From its measure we obtain the information that we need in order to design a robust control algorithm.

This way of approaching the problem is indeed very simple, but gives all the information needed for the compensator design and the classical robust control algorithm.

The classical way of constructing the compensator is based on inverting the nonlinearity model. This inversion can cause a lot of problems under different conditions and is sometimes impossible. That is why we propose a method that does not effectively inverse the model.

This step can be interpreted as a geometric „transposition" operation (inversion relatively to the first bisector) of the process's geometric characteristic.

Figure 7 presents the effective construction of the bijective characteristic. According to this, $u(k)$-the inverted model output depends on the value of the set point $r(k)$. This dependency is stored in a table [12].

This approach can be easily applied for bijective characteristics but, for the more complicated models, there will be more than one output value for each given input. That means that we need a selector of a higher degree to choose between the multiple outputs and select the specific pair that represents the compensator output for the input at that specific moment.

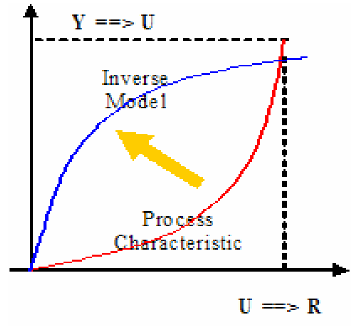


Fig. 7 Transposition operation for a bijective characteristic

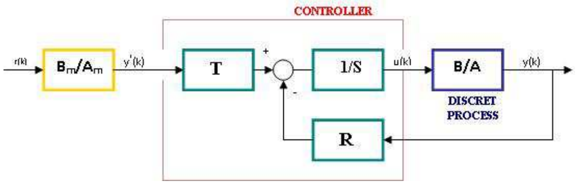


Fig. 8 RST control algorithm structure

A priori loading the models' values in tables, in an inverted manner and using selectors - is easy to implement, it is less computational costly than inverting other complicated models during run time and do not alter the control algorithm performances because this will be design with sufficient robustness (function of dissipation).

D. Control algorithms design

Classical control algorithms are included in the multiple model and feedback component of the nonlinearity compensator proposed structures. For this study, we decided to use a RST algorithm (Fig. 8). This is designed using pole placement procedure [6]. Here the "RST" controller's output is:

$$S(q^{-1})u(k) + R(q^{-1})y(k) = T(q^{-1})y^*(k) \qquad (1)$$

The R, S, T polynomials are:

$$\begin{aligned} R\left(q^{-1}\right) &= r_0 + r_1 q^{-1} + \ldots + r_{nr} q^{-nr} \\ S\left(q^{-1}\right) &= s_0 + s_1 q^{-1} + \ldots + s_{ns} q^{-ns} \\ T\left(q^{-1}\right) &= t_0 + t_1 q^{-1} + \ldots + t_{nt} q^{-nt} \end{aligned} \qquad (2)$$

Here, in some specific conditions ($T = R$) RST controller becomes PID numeric case [6].

The pole placement procedure is based on the identified process's model:

$$y(k) = \frac{q^{-d} B(q^{-1})}{A(q^{-1})} u(k) \qquad (3)$$

where:

$$\begin{aligned} B\left(q^{-1}\right) &= b_1 q^{-1} + b_2 q^{-2} + \ldots + b_{nb} q^{-nb} \\ A\left(q^{-1}\right) &= 1 + a_1 q^{-1} + \ldots + a_{na} q^{-na} \end{aligned} \qquad (4)$$

The identification is made in a specific process operating point and can use recursive least square algorithm [6].

This approach allows the users to verify and, if necessary, to calibrate the algorithm's robustness [6]. The "disturbance-output" sensibility function is given by the next expression:

$$\begin{aligned} & S_{vy}(e^{j\omega}) \overset{def}{=} H_{vy}(e^{j\omega}) = \\ & = \frac{A(e^{j\omega})S(e^{j\omega})}{A(e^{j\omega})S(e^{j\omega}) + B(e^{j\omega})R(e^{j\omega})}, \quad \forall \omega \in R \end{aligned} \qquad (5)$$

At the same time, the negative maximum value of the sensibility function represents the module margin.

$$\Delta M\big|_{dB} = -\max_{\omega \in R} \left| S_{vy}(e^{j\omega}) \right| \Big\|_{dB} \qquad (6)$$

E. Composed ratio control structure

Thus, the ratio control structure may contain classical algorithms, multiple model or nonlinearity compensation for C1 and C2 positions.

In industrial installations where different functioning points impose different models, wits different dynamics MM control structure should be recommended. In other situations, with especially, process gain varies; nonlinearity compensator control can be implemented.

F. Advantages and disadvantages of proposed control structures

The main advantage of the proposed control solution is the fact that it solves a difficult problem using the generalization of a well known control structure. The design procedure for the controller associated to a process regime is classical (pole placement) [2], [6].

The proposed structure is capable of a good ratio performance for both set-point changes and load disturbances that might occur in any of the two closed-loops. In the same time processes nonlinearity are treated correctly, using performing control structures.

Multiple model algorithms switching "sensitivities" are attenuated by "synchronizing" the dynamics of loops evolution.

The main disadvantage is the necessity of the model identification for RST control structure. Entire combined control structure needs supplementary computational resources.

III. CASE STUDY

We have evaluated the performances of the combined structures (Fig. 2, 3 and 4) on a laboratory platform – a double filling-emptying process with self-stabilization, where the level/flow control of the upper tank is of interest (Festo^{TM}) – figure 9. In the same time, two equivalent software nonlinear level process control simulator ware designed and implemented (Fig.10 and 11). First one (simulator no. 1 – Fig. 10) require multiple model control structure and second one, (simulator no. 2 – Fig. 11) respectively, nonlinearity compensator solution.

These platforms (laboratory and simulators) allow the connection of control software for maintaining the desired level of the water in the upper tank and prescribed ratio. The control application (Fig. 12, 13 and 14) has three stand alone modules: real time implementation of nonlinearity compensator controller (for C1 or C2) in Fig. 12, multiple model controllers (for C1 or C2) in Fig. 13 and mixing module (BS1 and BS2 from Fig. 2) in Fig. 14. Their

Fig. 9. Festo laboratory platform

functionalities are: connection with the platform and software simulators, setting/applying the automat command, setting/applying the set point value, evaluation of RST algorithms output, setting the sampling period value, loading of process model and control algorithm parameters, displaying the real time evolution curves etc.

The control applications were developed by using National Instruments's LabWindows/CVI. A connection with Festo laboratory platform is possible by using a National Instruments cDAQ data acquisition module. The code for all control structures is directly portable for other more powerful hardware structure like PXI. For upper tanks the nonlinearities was accentuated by introducing different objects – similar to simulators figures.

For the considered process (laboratory platform/ simulators), the static characteristic was determined (Fig. 11) and three models were identified (Fig. 10). A convenient sampling period for 5 seconds was used.

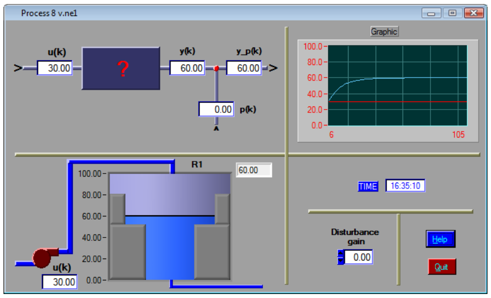


Fig. 10. Software simulator no. 1 for nonlinear level control (MM)

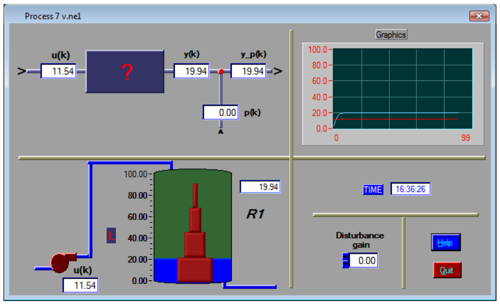


Fig. 11. Software simulator no. 2 for nonlinear level control (NC)

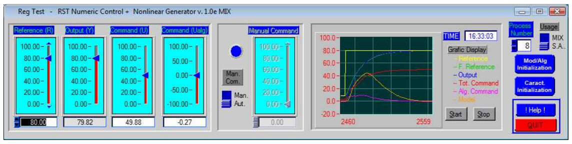


Fig. 12 Interface of real-time control software application with nonlinearity compensator – main window

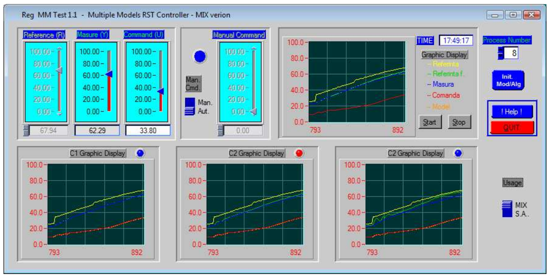


Fig. 13 Interface of real-time control software application multiple model – main window

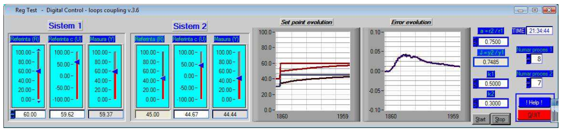


Fig. 14 Interface of real-time mixing component – main window

For the process where the control is based on static characteristic, the model (identified at 50% of the medium static characteristics), was obtained by using the recursive least squares procedure [6] and WinPIM software [6]:

$$M_1(q^{-1}) = (0.14572 + 0.08775q^{-1}) / (1. -0.64022q^{-1} -0.18434q^{-2})$$

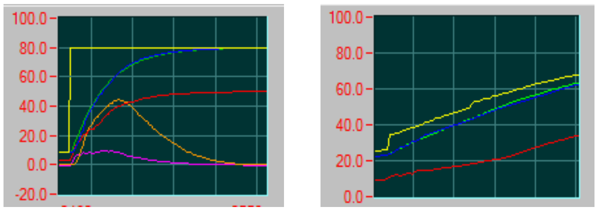
The corresponding RST controller determined using poles placement procedure and WinREG [6] software is:

$$R\ (q\text{-}1) = 3.373494\ \ -2.184162\ q^{-1}\ \ -0.102683q^{-2};$$
$$S\ (q\text{-}1) = 1.000000\ \ -0.951121\ q^{-1}\ \ -0.048879\ q^{-2}$$
$$T\ (q^{-1}) = 4.283206\ \ -4.710477\ q^{-1}\ \ 1.513920\ q^{-2}$$

The same procedure was applied for each of the three models/controllers of process controlled based on multiple model (Fig. 10).

To verify the proposed control structure with multiple model controller for C_1 and nonlinearity compensator controller for C_2 few tests were made: for example – the set point was increased from 20% to 80% (Fig. 15), in order to show the use of nonlinear compensator (Fig. 15, a)) and switching between controllers (Fig. 15, b)). Here, controller switching (from 1 to 2) is not visible.

In Fig. 15 the colors code is: yellow – set point; green – filtered set point; blue – process output; red – control structure output (command).



a) b)

Fig. 15 a) Evolution for nonlinearity compensator controller; b) Evolution for multiple model controller;

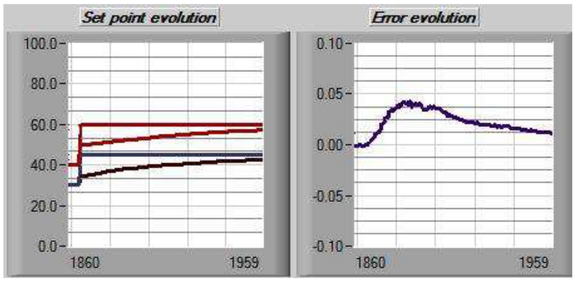


Fig. 16. Weak to medium coupling ratio factor (right) and set points evolution

Similar tests were made to verify ratio maintaining during disturbances effect. During these, the *a* ratio was imposed to 0.75, the coupling factors k_1 and k_2 set to 0.3 respectively to 0.25 (high coupling). As results, for "brutal" 20% to 80% set point switching test, *a* ratio factor has a maximal 0.01 (effective) deviations.

In other test, with the same *a* ratio imposed to 0.75, the coupling factors k_1 and k_2 ware set to 0.5 respectively to 0.3 (weak to medium coupling). Here, for 40% to 60% set point switching test the ratio factor has a maximal 0.045 deviations (Fig. 16). Here, on left side, red and blur represents first, respectively second system set point evolution.

In these tests the ratio factor evolution tends to 0. When smaller deviation is imposed corresponding high coupling is required (small values for k_1 and k_2).

For all tests both close loops system remain stable and track precisely the set point. These performances could not be reached with a classic close loop system; important set point changing means different functioning points and different performances.

The real time controller output's (command) evolution has small oscillations caused by ultrasound sensor output.

These shocks are attenuated and by RST algorithm that "corrects the jumps" between controllers.

V. CONCLUSIONS

In this paper a precise solution structure for ratio control of nonlinear processes blend stations is proposed. Comparative to classic solution (Fig. 1), [2], [13] important precision improvement is provided. The main feature of the proposed structure is to give a good ratio performance for both set-point changes and load disturbances that might occur in any one of the loops. For each loop (two or more) a nonlinear structure based on multiple model (MM) and nonlinearity compensator (NC) is proposed.

The main advantages consist in classic tuning procedure for each controllers and very simple procedure for nonlinearity determination, suitable for industrial applications. The considered modeling technique and design methodology combined with the powerful RST control algorithm proved in practice that this solution is feasible for the control of the presented classes of nonlinear processes.

The real-time implementation of the proposed structure is possible and easy to realize on the existing dedicated industrial control platforms.

ACKNOWLEDGEMENT

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