Is it Time for a New Approach to PID Tuners

Mark L. Darby, CMiD Solutions

Michael Harmse, IPCOSAptitude Ltd.

Abstract: As typically practiced today, PID tuning is performed on a single-loop basis, via trial and error, or based on the results of an open-loop step test. For interacting loops, this is a time-consuming task that often results in compromised performance, or at least leaves open the questions: "How much performance is being left behind?" and "How robust is the tuning?" This contribution presents a multivariable approach to PID tuning, which takes advantage of modern identification methods and incorporates a control-relevant optimization to determine optimal PID tuning parameters. A key feature of the approach is that robustness margins to dead time and gain errors are explicitly addressed.

1 Introduction

Most applications of PID controllers today involve an interactive, multivariable process. The interaction is due to the inherent nature of typical chemical processes as well as complexities due to heat integration, recycle streams and intermediate buffers. As a result, tuning multiple interacting PID controllers for such situations can be a challenging task, further complicated by the sheer number of PID loops in the typical plant.

PID controller tuning is typically performed on a single-loop basis using a trial and error approach, or based on the identification of a low-order model from an open-loop step test and corresponding tuning rules linking process model parameters to PID tuning parameters. With the first approach, decisions are typically made as to which loops are the most important and therefore tuned tighter and the others tuned slower. If significant interactions are observed, then heuristic detuning is performed, and the least important loops are made even slower. A challenge with the second approach is that a low order model may not accurately represent the true process dynamics, due to the afore-mentioned complexities; in addition, such tuning rules are typically limited to the single input single output case.

A key concern once tuning is performed is how well will the loops perform in the future. Process plants are, to at least some degree, nonlinear. Tuning that performed adequately for a given production rate or a particular feed type may perform poorly later when operation changes. It is obviously desirable to ensure stable and adequate performance of the overall system when plant responses change.

This paper presents a new approach to tuning interacting PID loops, taking advantage of modern identification methods, advances in optimization theory, and the availability of fast multi-core processors. The procedure for developing PID tuning for multivariable systems consists of the following steps:

- Plant test of the multivariable process.
- Model identification.
- Model reduction to a state-space model with explicit dead time.
- Constrained optimization to identify optimal PID tuning parameters.
- Simulation of the optimized process.

The last three steps have been incorporated into a software package called AptiTune^{TM}. It allows the user to interactively determine the best tuning for his/her application based on *process safety* and *operability* considerations and *robustness* specifications, tailored to the *specific DCS*. The goal is to shift iterations from the process to the software package, and to achieve plant performance that closely matches the optimized results. This is referred to as "one shot tuning".

In the following, the workflow and optimal tuning approach are further described. Tuning results are presented for an application of AptiTune^{TM} involving selected loops in a Desulphurizer Unit.

2 Motivating Example

Here an example, shown in Figure 1, is considered to motivate the proposed approach. From the control configuration shown in the figure, process interactions can be expected to exist in the following:

- Tray temperature (TC1) and overhead pressure control (PC1).
- Feed preheat via LC1.OP and its effect on tray temperature (TC1.PV) and overhead pressure (PC1.PV).
- Interaction from reflux drum LC2, if drum level is controlled by reflux flow (FC2).

Here, PV refers to the process value (feedback variable) and OP refers to the output of the controller (the manipulated variable). As a result of these interactions, we see that we have at least a 2x2 interacting problem, possibly as high as a 5x5 problem. For illustrative purposes, assume that the reflux drum level is controlled with distillate flow. When interaction is small between certain subsystems, the overall problem can be split up. For this example, we could treat the distillation tuning problem as a 4x4 system or as multiple subsystems: a 2x2 (TC1 and PC1), a 1x1 (bottoms level, LC1) and a 1x1 (reflux drum level). For a given subsystem, we would need to develop a model between all the controller outputs (or manipulated setpoints if the PID output is connected to a cascaded controller) and all the controlled variables in the subsystem.

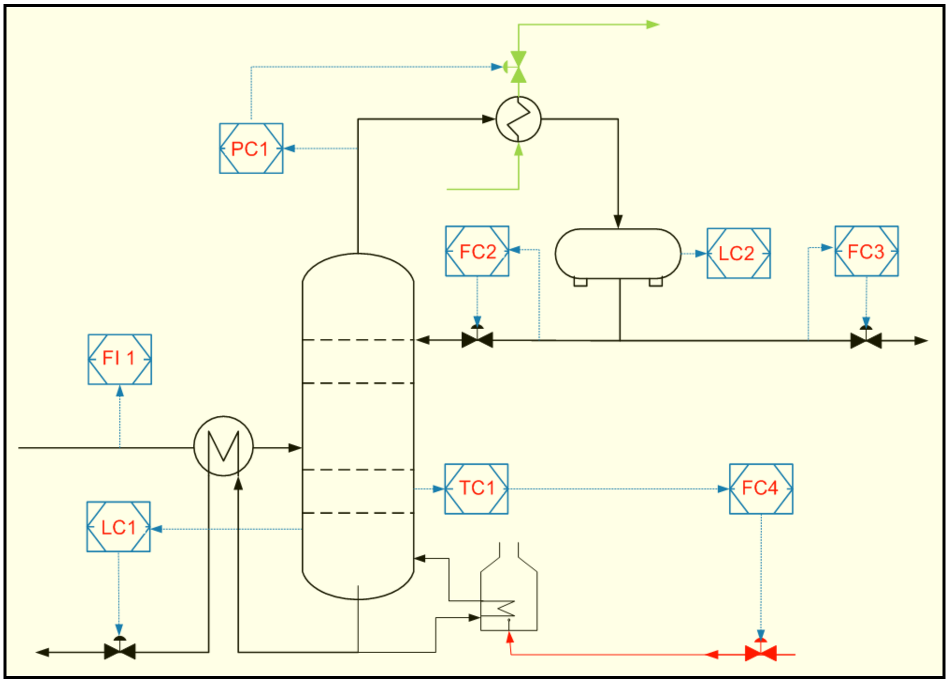


Figure1. Example Distillation Column.

To tune this overall system, we would need to account for interactions. In addition, we would want to ensure the following:

- Avoid excessive kick from tray temperature output (TC1.OP) to reboiler fuel gas (FC4) to prevent flame out on the low side or tube damage on the high side
- Adequate performance for disturbance rejection and setpoint changes
- Limit the amount of PV overshoot in TC1 and PC1.
- Take advantage of buffering capacity in the bottoms sump to smooth out bottoms flow fluctuations and minimize interactions between feed preheat and the other loops.
- Take advantage of buffering capacity in the reflux drum.

3 Work Flow

The work flow parallels that used for model predictive control (MPC): Test » Model » Tune.

A plant test is performed on the various subsystems where PID tuning is desired. The goal is to identify a dynamic model that accurately captures the input-output behavior of the plant. With an accurate model, we have high confidence that performance of the optimal tuning on the actual plant will match the simulated responses. The plant test can be performed open-loop or closed-loop.

> Open-Loop: Controllers are placed in manual mode and steps made to the controller outputs. Four to six steps of varying duration is typically sufficient, as long as the steps are of sufficient amplitude.
>
> Closed-Loop: Controllers are kept in automatic and steps are made to the controller setpoints.

In both cases, it is important to vary the step durations in order to capture the relevant dynamics of the process. It is recommended that steps be made of various durations, ranging from 10% to 100% of the estimated open-loop response time Methods based on PRBS (pseudo-random binary sequence) or GBN (generalized binary noise) can be used to automate the testing. Testing time is often on the order of 1/2 day to 1 day for problems of size 5x5 or smaller, with settling times in the order of 1 hour or less. It is important to make moves larger than the amount of valve stiction. Note that some DCS systems have standard features for enforcing a minimum move size in the controller outputs for the closed-loop case.

The next step is to analyze the raw data to select the data ranges for model identification. Modern techniques based on FIR (finite impulse response), ARX (auto regressive moving average) and Subspace methods are used. AptiTune™ supports the import of models created by identification tools from MPC packages. It also allows the user to specify a transfer function matrix.

Once a model has been imported into AptiTune™, the next step is to smooth the model (if necessary) and covert the model to a state-space model using a singular value decomposition (SVD) reduction technique (Maciejowski, 1989). State space models exhibit much more process realism than FIR models identified from short noisy step test data sequences. State space models are also much more compact (less parameters), and therefore improves simulation speed. Dead time can also be set explicitly during this step.

The smooth state space model is then used to determine optimal PID tuning parameters for the closed-loop control system. The user is required to enter instrument ranges and to select the desired PID equation associated with a particular DCS. The user can select from popular DCS systems, such as

Honeywell TDC and Experion, Emerson DeltaV, Foxboro I/A, ABB Symphony, Yokogawa and others. Optimal controller parameters can also can be calculated for P-only, I only, and PI options.

Specification and constraints are specified by the user and an optimization is performed. Design cases are logged and the user can analyze the different scenarios in terms of setpoint tracking, disturbance rejection, and noise attenuation. The integrity of the system to one or more controllers being placed in manual can also be analyzed. Finally the robustness against plant-model mismatch is evaluated and displayed in an easy to understand graphical way.

4 Optimization Formulation

The PID controller parameters, controller gains, $k_{c,i}$, integral times, $\tau_{I,i}$ and derivative times $\tau_{I,i}$, are determined by solving the nonlinear, constrained, optimization problem

$$\begin{aligned} &\min_{k_{c,i},\tau_{I,i},\tau_{D,i}} J \\ &st \\ &c_j(k_{c,i},\tau_{I,i},\tau_{D,i}) \le 0, \quad i=1,\dots,n,\ j=1,\dots,m \end{aligned} , \qquad (1)$$

where J denotes the objective function and the c_j denote the constraints. The objective function J is a weighted sum of three terms $J = J_1 + \alpha J_2 + \beta J_3$, each of which contributes different aspect of closed-loop performance. The first term $J_1 = \int_0^{t_f} \left| y^r(t) - y(t) \right| dt$ is based on an integrated absolute error (IAE) criterion. Here y^r is an optional user-defined first order trajectory. The second term $J_2 = \int_0^{t_f} \left| y^{sp}(t) - y(t) \right| dt$ denotes the IAE for a load disturbance occurring at the input of the process. The third term $J_3 = \int_0^{t_f} \left| \Delta u(t) \right| dt$ is a criterion used to affect control effort. The weighting coefficients α and β are user specified to reflect the relative importance of the three objectives. For each control loop, the user can specify one or more of the following inequality constraints:

- Maximum OP change following a setpoint change.
- Maximum PV overshoot following a setpoint change.
- Minimum damping ratio.
- Maximum noise amplification.
- Maximum process-model gain mismatch.
- Maximum process-model dead time mismatch.

For buffering applications such as with liquid level, the following may be specified:

- Maximum setpoint deviation and minimum return time following a preset level disturbance.

By specifying appropriate constraints, the user can tailor the tuning to process-specific requirements.

Initialization values for the optimization may be selected as the actual DCS values or calculated based on the Cohen-Coon SISO tuning rule. The degree of difficulty of the optimization depends on the number of controllers, the order of the state-space model and the number of constraints. The problem is non-convex and local minima can occur. Therefore, several optimization methods have been implemented, including a brute force global search, a genetic algorithm and a generalized gradient algorithm (Vlachos et al., 2000).

Robustness plots are provided for each PID control loop. The robustness plots are similar to those proposed and developed and by Garry and Hansen (1987). A robustness plot is shown in Figure 2. The plot shows the stability line (in purple) as a function of gain ratio and dead time deviation, relative to the nominal model. The minimum stability region is shown in red. The robustness spec is shown in light blue. A gap between the robustness spec (the blue circle) and the stability line indicates that the robustness specification is not an active constraint in the optimization. The user can mouse click on any point in the plot to observe both the open loop step response relative to the nominal model (shown in the same frame as the robustness plot) and closed-loop PV and OP setpoint responses in the plots to the right.

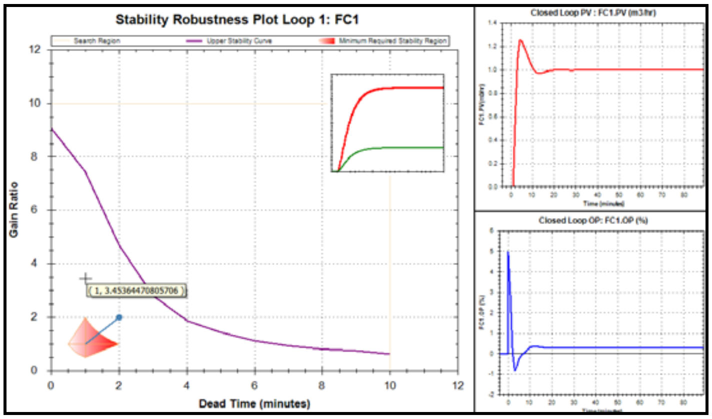


Figure 2. Robustness Plot

5 Industrial Example

The method and software package have been applied to loops in a Desulphurizer Unit. The client had deemed the performance of the existing loops as adequate, but was interested in determining if performance could be improved. Note that no use was made of the existing tuning parameters.

A diagram of the process is shown in Figure 3. PC001 sets the backpressure on the unit, as the other PCs are kept in manual with the valves 100% open. The PC manipulates fuel gas import, which affects the heating value of the gas, and thus interacts with loops TC001 (reactor temperature) and TC002 (column temperature). In addition, changes to fuel gas flow via TC001 and TC002 affects the pressure loop PC001. As a result, the 3x3 system is interactive. The normal cascade for the reflux level controller is the distillate flow, and the step test was conducted in this mode. An interesting aspect of this problem is that the three loops are associated with different equipment in the unit. As a result of this distribution and the interaction, the settling time is fairly slow, on the order of 90 minutes.

A closed loop multivariable test was performed with the following loops.

Loop	**Controller variable**	**Manipulated Variable**
TC001	Reactor inlet temperature	Fuel gas flow setpoint
TC002	Reboiler outlet temperature	Fuel gas flow setpoint
PC001	Unit back pressure	Fuel gas make up valve

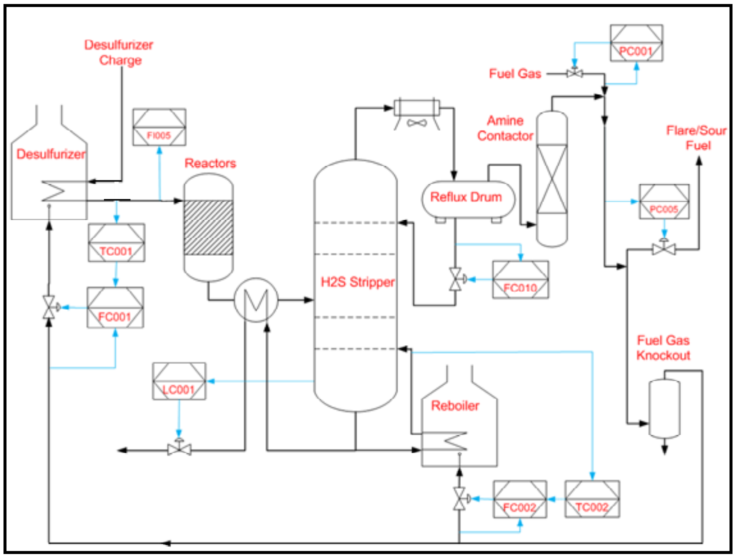


Figure 3. Desulphurizer Unit

Standard TDC3000 capabilities were used to conduct the step test with operator supervision only. OPMNCHG parameters were used to ensure valve changes greater than stiction limits. Setpoint changes of varying step size were made to all loops. Since the settling time was fairly slow (90 minutes), a total of 23 hours of test data was required for this particular application. The test signals are shown in Figure 4.

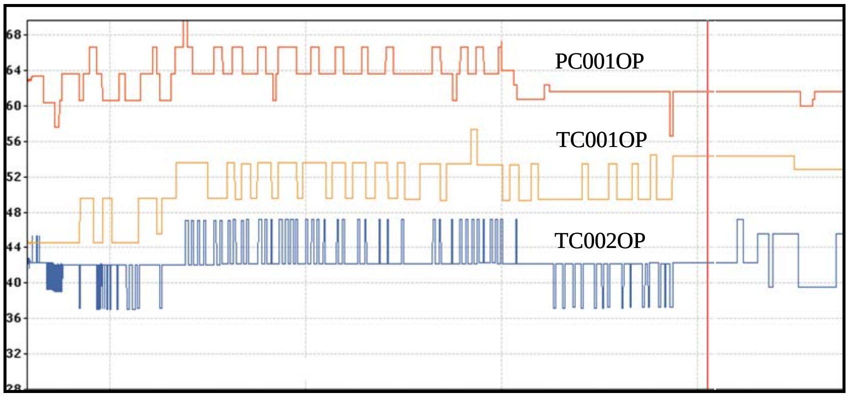


Figure 4. Step Test Trends

Intermediate identification results were made to check progress of the test in terms of model accuracy. These results were used to modify the step width durations (frequency content) of the step signals.

The identification modeling and PID tuning was completed in 2 hours. A 3x3 state space model was identified using a subspace method. The model matrix is shown in Figure 5. Note that the model matrix is fully dense and that most of the responses exhibit high-order responses (which cannot be approximated as first order plus dead time models). The loop pairing (OP and PV of the same control loop) is determined by the diagonal curves so these model curves need to be strong on typical move scaling. The strength of the off-diagonal model curves provides a good indication of the degree of interaction between the three PID loops. This is the reason why all three loops should be optimized simultaneously, rather than one at a time.

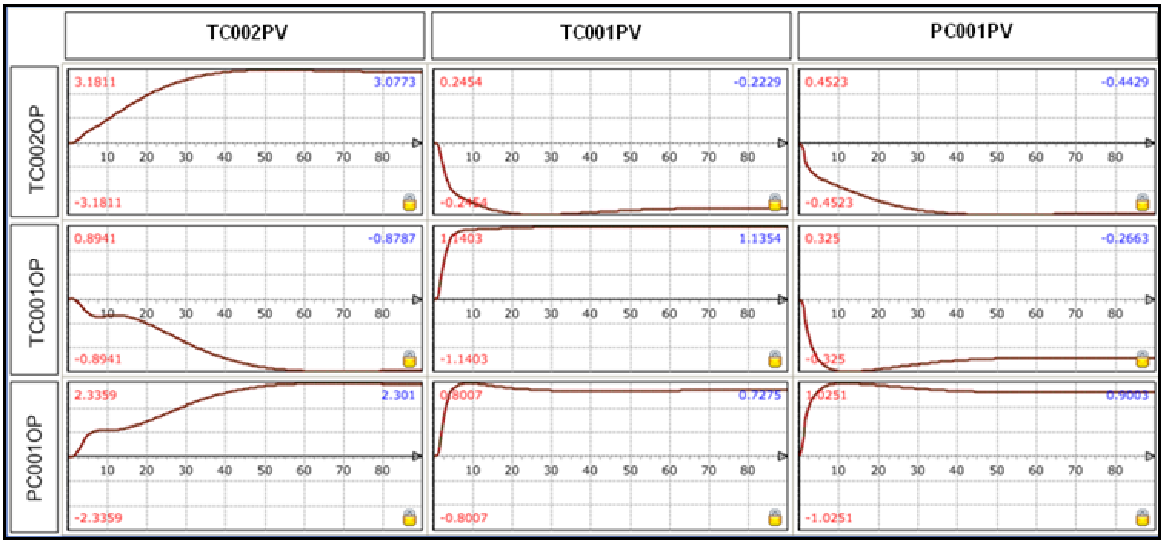


Figure 5. Step Response Model

The following criterion was used to tune the loops in AptiTune^{TM}

Column TC002	**Reactor TC001**	**Backpressure PC001**
•Remove cycle tendency •Improve dampening •Minimize overshoot	•Tuning more agressive •Good dampening •Low PV overshoot •Hard constraint on OP	•Remove cycle tendency •PV to SP in 3 min •10% max PV overshoot •Moderate kick in OP

The existing tuning showed that both TC002 and PC001 had a tendency to cycle. TC001 was tuned very sluggishly.

Robustness margins for gain and dead time were specified for all loops and were active for TC001 (gain ratio margin 2 and dead time margin of 1) and PC001 (gain ratio margin 1.25 and dead time margin of 1). Closed loop simulated setpoint responses are shown in Figure 6. Each row in the matrix corresponds to a setpoint change in the corresponding controller (diagonal entries). Interaction responses are shown in the non-diagonal entries. Stable responses, without excessive manipulated variable movement, are observed in the simulated responses of all loops.

The tuning parameters for the three DCS loops were updated with the AptiTune^{TM} results. The loops were then tested by making setpoint change and observing responses to normal plant disturbances. The results on the process were found to closely match those in simulation. For example, PV rise times, amount of overshoot, time to reach the peak OP value, time at which maximum PV overshoot occurs, etc, were all within 20% of actual.

Before and after trends, of length 48 hours, are shown in Figure 7. The data for the trends were selected to obtain similar values of feed rate and number of setpoint changes. In analyzing the before and after results, it is necessary to analyze both PV and OP variances, since tuning affects both. As indicated earlier, both column TC002 and backpressure PC001 had a tendency to cycle. This tendency was removed with the new tuning. In both cases, we see that the OP variance is decreased. We see that the cycle in column TC002 is removed. The decrease in the OP variance for pressure PC001 is accompanied with an increase in PV variance. This is consistent with the stated desire to increase robustness of the PC001. The tighter tuning in reactor TC001 is clearly seen.

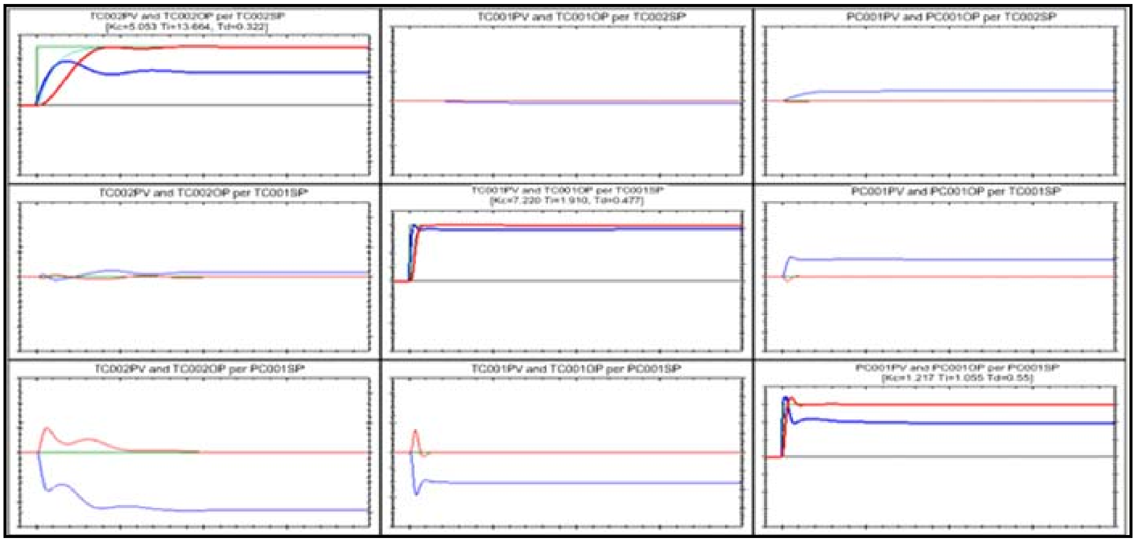
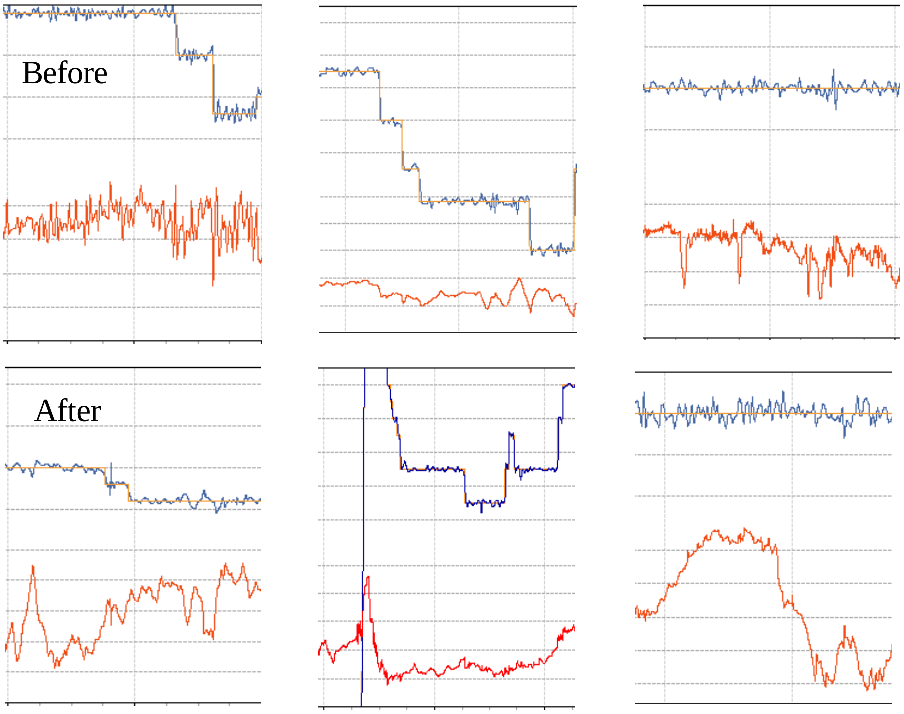


Figure 6. Closed-loop Step Response (Red: PV; Blue: OP).



Column TC002	Reactor TC001	Backpressure PC001
•Reduced sp rise time •Tuning less aggressive •Cycle tendency removed	•Tuning more aggressive •Robustness still good	•Tuning more robust •Tuning less aggressive •Cycle tendency removed

Figure 7. Before and After Data, Duration: 48 Hours (Blue: PV; Red: OP).

The tuning determined by Aptitune^{TM} was placed in service in October 2008 and was kept unchanged until a review of the loops was help in January. Since stable control was observed, it was decided to tighten the tuning of backpressure PC001. This was accomplished in Aptitune^{TM} by reducing the gain ratio margin (see Figure 7). The resulting tuning, which was observed to perform well, remains in place today.

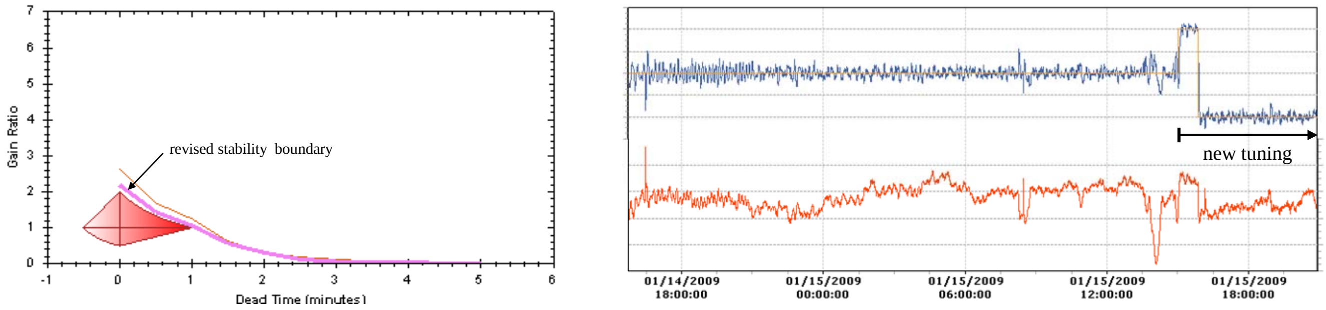


Figure 7. Updated PC0001 Tuning - Tighter Robustness Margin (left), Before and After Data (right).

6 Conclusions

A new approach to tuning multiple interacting PID controllers was presented. The methodology consists of a dedicated plant test, followed the use of modern identification technique to identify an accurate dynamic model. The resulting model is used in AptiTune^{TM} to develop optimal PID tuning parameters based on process safety/operability considerations and plant-model mismatch margins. The advantage of the approach is that if an accurate model is identified, the resulting tuning will yield process responses that closely match the optimized responses, thereby avoiding trial and error methods.

The approach was successfully applied to an interacting 3x3 Desulphurizer problem. The application was completed in 25 hours. Although not mentioned in the main body of this paper, the sump level was also tuned with Aptitune^{TM}. The SISO test and tuning for the level was performed in 6 hours

These results demonstrate that a model-based approach to multivariable PID loop tuning is viable. The majority of the effort involved step testing, which was automated in this application and did not burden the APC engineer. The responses calculated in Aptitune^{TM} performed similarly to the actual plant responses. There has been no need to manually modify the tuning that was calculated by the software. The goal of "one-shot tuning" was demonstrated, as well as the ability to take process operability considerations and safety constraints into account.

References

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