Valve Positioning Control for Process Through-put Maximization

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Abstract

The application of a valve positioning controller (VPC) on top of a basic regulatory plant-wide control structure for maximizing the process through-put to increase plant profitability is demonstrated. The input to the VPC is a measurement of the bottleneck constraint and its output is the set-point of the through-put manipulator. The HDA process is used as an example case study with the feed hydrogen compressor considered as bottleneck constraint that limits production. Results show that the automatic adjustment of the through-put manipulator allows the process to be operated close to (or at) the bottleneck constraint. De-rating of the plant through-put due to the possibility of disturbances is then avoided resulting in enhanced through-puts and plant operating profit. The increase in the through-put and plant profit over the de-rated operation is quantified with respect to variation in the magnitude of the principal disturbance that causes the bottleneck constraint to be hit. Results show that significant enhancement of more than 20% increase operating profit can be achieved. VPC is thus demonstrated to be a simple and effective means for increasing plant profitability by maximizing through-put.

Key-words: Plant-wide Control, Valve Positioning Control, Through-put Maximization

Introduction

Research in plant-wide control has received much attention since the early nineties, aided to a large extent by the ready availability of commercial packages such as HYSYS and Aspen, for rigorous plant-wide dynamic simulations. Much of the research in plant-wide control has focused on methodologies for synthesizing effective decentralized control structures. Luyben and coworkers have done seminal work on the design of effective plant-wide regulatory control structures for complex chemical processes culminating in a nine step heuristic procedure for (Luyben and Luyben, 1997). The application of the procedure for effective plant-wide regulatory control has been illustrated on a variety of chemical processes such as the hydro-dealkylation of toluene, the Tennessee Eastman challenge process, a vinyl acetate process, an isomerization process etc (Luyben et. al., 1999).

Given a basic regulatory structure, the next step is to adjust the key process set-points to maximize the plant operating profit. Surprisingly, there are very few articles in the open literature that explicitly address the issue of plant-wide control for maximum operating profit. This is even as modern chemical plants are known to routinely employ constrained NLP optimization of the key set-points to optimize plant profit within the framework of model predictive control or real time optimization. For continuous chemical processes producing bulk chemicals, maximizing plant profitability many-a-times boils down to maximizing the process through-put with the increased production rates translating directly to higher profits due to the raw material – product price differential. Constraints such as column flooding, maximum available flow rates (material, heating or cooling) etc typically act as bottlenecks limiting the achievable production rate. Maximizing the production rate then requires controlling the plant in a manner such that it operates as close as possible to the bottleneck constraint. The application of

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constrained NLP optimization approaches for the purpose is usually tedious requiring much effort in the proper formulation of the optimization problem as well as developing a high fidelity plant-wide process model. Simpler approaches to automate through-put maximization, are most desirable for the purpose.

In this work, the application of a valve positioning control (VPC) type approach (Shinskey, 1996) for maximizing plant through-put is demonstrated on the hydro de-alkylation (HDA) process converting toluene to benzene (Douglas, 1988). Rigorous dynamic simulation results for two bottleneck constraints, namely, the fresh hydrogen feed compressor duty and the furnace duty show that the plant profitability can be increased by more than 20% over a basic regulatory plant-wide control structure.

Valve Positioning Control for Through-Put Maximization

In a chemical plant, the set-point of the through-put manipulator is adjusted to change the production rate. Variables that substantially affect the reaction rate, e.g. the reactor temperature or limiting reactant flow rate into the reactor, are usually used as through-put manipulators. In order to maximize production, an operator would typically increase the set-point of the through-put manipulator until the bottleneck constraint is approached or hit. Hitting of the constraint generally implies the loss of a control degree-of-freedom. For example, the duty of a heater may hit its maximum value in order to heat the process stream to its temperature set-point as the plant through-put is increased. This corresponds to losing a control degree-of-freedom with the temperature control loop becoming ineffective. In order to prevent such a scenario, the operators must provide a trim so that the bottleneck constraint is approached closely but never becomes active. The magnitude of this trim is determined by the worst-case disturbance anticipated so that even in the presence of such a disturbance, the bottleneck constraint is not hit and the loss of a control degree-of-freedom is avoided. Clearly, the need for this trim represents a loss in the production rate when there is no disturbance into the process as the through-put manipulator set-point has been de-rated to ensure the rejection of the worst-case disturbance without the bottleneck constraint becoming active.

If the bottleneck constraint is known *aprior*i, the adjustment of the throughput manipulator can be automated using a VPC type approach. Shinskey (1996) first proposed the concept of VPC for reducing energy consumption in distillation columns. For plant through-put maximization, the input to the VPC controller is a measurement of the bottleneck constraint and the output is the through-put manipulator set-point in a cascade arrangement as illustrated in Figure 1.

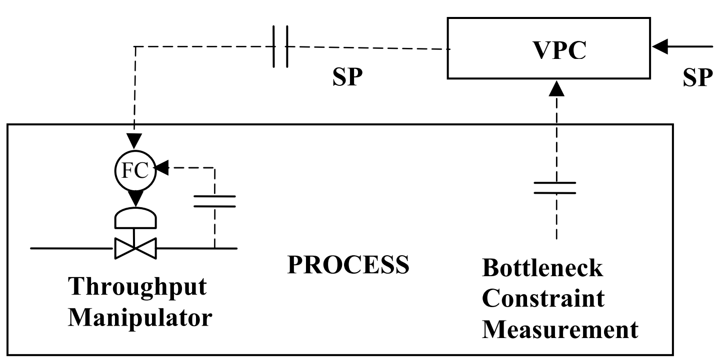


Figure 1 VPC for through-put maximization

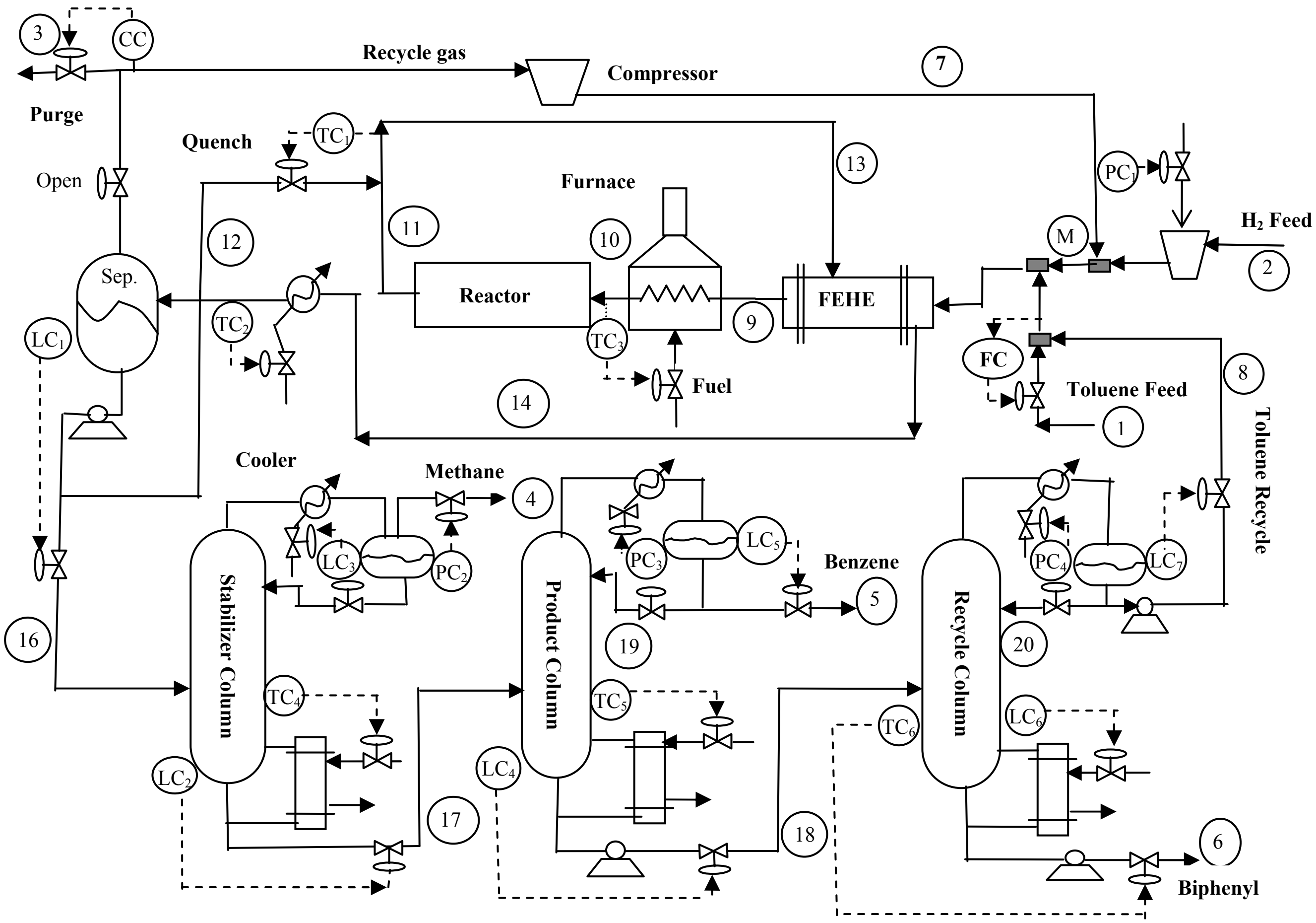
Referring back to the heater example, the input to the VPC is the heater duty (the bottleneck constraint) and its output is the through-put manipulator set-point which is equivalent to adjusting the flow of the process stream through the heater. Notice that with the VPC in place, even with a heater duty valve being saturated at 100% during transients, i.e. transient operation at the bottleneck constraint, temperature control of the process stream is not lost as the VPC reduces process stream flow through the heater such that the heater duty remains at its near maximum set-point (say 95%). The VPC scheme then emulates temperature control of the process stream using the feed flow as the manipulated variable with the heater duty close to maximum. Thus, even as a control degree-of-freedom is lost, temperature control of the process stream is not lost and the need for de-rating is mitigated resulting in an increase in the achievable production rate. A VPC type controller on top of a basic plant-wide regulatory control structure thus provides a simple and effective means for maximizing the plant through-put.

HDA Process Case Study

The HDA process flowsheet as synthesized by Douglas (1988) is shown in Figure 2 along with the basic regulatory control structure. The plant design and operating conditions have been taken from Luyben (2002). Table 1 reports the controller tuning parameters for the different controllers in the plant-wide control structure along with the sensor and controlled variable spans.

Table 1: Controller parameters for HDA process plant-wide control structure

ID	*Type*	*Action*	*Set-point*	K_c	τ_I *(min)*	*Sensor Span*	*Valve Span*
PC_1	PI	reverse	3240 kPa	3.78	5	2413-4068 kPa	0-1E+06 kJ/hr
FC	PI	reverse	Varies	0.5	0.5	0-300 kmol/hr	
TC_3	PI	reverse	621 °C	1.5	10	482-760 ^oC	0-1E+08 kJ/hr
TC_2	PI	direct	45 °C	1	10	26-63 ^oC	0-1.5E+08 kJ/hr
TC_1	PI	direct	612 °C	1	10	482-743 ^oC	
CC	PI	direct	59.23 mol% CH_4	0.226	14	0-100%	
PC_2	PI	direct	1000 kPa	20	2	500-1500 kPa	
TC_4	PI	reverse	135 °C	1	10	100-200 ^oC	0-1.5E+07 kJ/hr
PC_3	PI	direct	200 kPa	1	10	131-269 kPa	0-3.9E+07 kJ/hr
TC_5	PI	reverse	131.6 °C	1	10	63-200 ^oC	0-3.4E+7 kJ/hr
PC_4	PI	direct	331 kPa	10	3	207-510 kPa	0-5E+06 kJ/hr
LC_7	PI	direct	50%	1	10	0-100 %	0-5E+06 kJ/hr
TC_6	PI	direct	186.6 °C	1	30	100-272 ^oC	
LC_{1-6}	P	direct	50%	2		0-100 %	



1. Toluene fresh feed
2. Hydrogen feed
3. Purge gas
4. Methane (stabilizer gas)
5. Benzene product
6. Biphenyl product
7. Gas recycle
8. Toluene recycle
9. Furnace inlet
10. Reactor inlet
11. Reactor effluent
12. Quench
13. FEHE hot in
14. FEHE hot out
15. Separator gas out
16. Stabilizer feed
17. Stabilizer bottoms
18. Product bottoms
19. Product column reflux
20. Recycle column reflux

Figure 2 The HDA Process flowsheet and its plant-wide control structure

The total toluene flow set-point or the reactor inlet temperature set-point are the two possible through-put manipulators that substantially effect the reaction rate and by extension the benzene production rate. For an equal increase in the production rate of benzene, the decrease in the product selectivity turns out to be lower when the total toluene is used as the through-put manipulator. Accordingly, the total toluene flow set-point is used as the production rate handle that is adjusted for through-put maximization. A rigorous dynamic model of the HDA process with the control structure in Figure 2 is developed in Hysys 3.2 and used to investigate the enhancement in the plant through-put using the VPC based approach described earlier.

The VPC based through-put maximization strategy is demonstrated for the fresh hydrogen feed compressor duty as the boyttleneck. For a quantitative evaluation of the economic benefits of increasing production rate, the plant operating profit is calculated using the cost data in Table 2. The corresponding VPC scheme is shown in Figure 3 where the VPC controller manipulates the total toluene flow set-point, the through-put manipulator to maintain the feed compressor duty at its near fully open set-point of 99%..

Table 2 Cost data for calculation of plant operating profit

Utilities	
Electricity	\$11E-06 per KJ
Heating (Steam)	\$8E-6 per KJ
Cooling (water)	\$0.7E-06 per KJ
Fuel	\$4E-06 per KJ
Raw Materials and Products	
Hydrogen feed	\$2.5 per kmol
Toluene feed	\$14 per kmol
Benzene product	\$19 per kmol
Hydrogen purge	\$1.08 per kmol
Methane purge	\$3.37 per kmol
Biphenyl product	\$11.84 per kmol

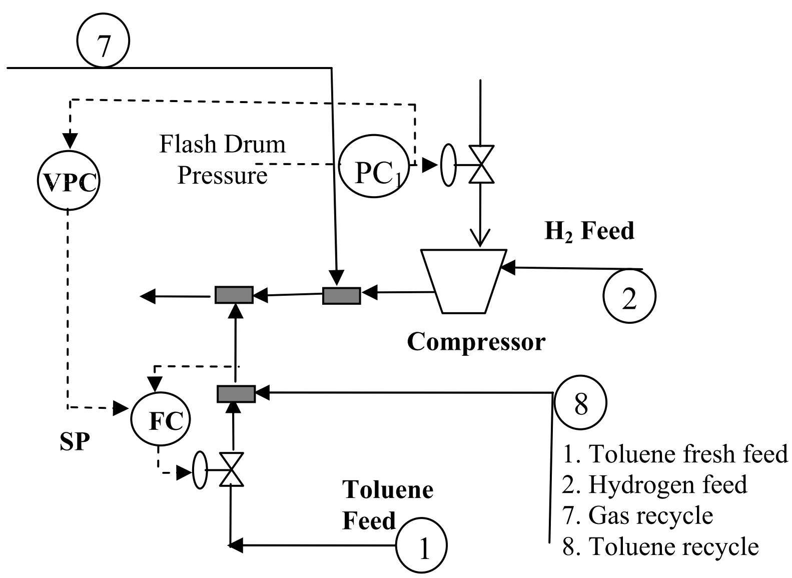


Figure 3 VPC for hydrogen feed compressor bottleneck

The hydrogen feed compressor duty is manipulated to maintain the pressure in the gas recycle loop. For higher production rates, as the total toluene feed (the through-put manipulator) to the reactor is increased, the gas pressure reduces due to greater hydrogen consumption in the reactor. The pressure controller then increases the feed compressor duty to push in more hydrogen for maintaining the gas pressure. For a marginally over-designed feed compressor, the compressor duty would reach its maximum and thus become the bottleneck constraint limiting production. With no disturbances such as changes in the composition of the fresh feeds to the process, the maximum production rate possible is obtained by changing the total toluene flow set-point in small increments until the feed compressor duty approaches 100% (say 99%). A benzene production rate of 170.8 kmol/hr is achieved. The total toluene feed rate, the through-put manipulator, is 262.8 kmol/h. The steady-state hourly operating profit is calculated to be $305.62 per hour.

The worst-case disturbance considered for the feed compressor bottleneck is a step decrease in the fresh hydrogen feed composition from 95 to 90 mol%. The reduced hydrogen composition causes the recycle gas pressure to go down as more gas gets consumed in the reaction than is being fed. The pressure controller then must increase the feed compressor duty to maintain the pressure. If the compressor duty is already touching 100%, the recycle gas pressure can no longer be controlled due to the loss in a control degree-of-freedom. The possibility of a disturbance thus necessitates de-rating the total toluene flow set-point such that even for a 5 mol% step decrease in the fresh hydrogen feed purity, the pressure controller is able to maintain recycle gas pressure with the compressor duty approaching 100% (say 99%).

Figure 4 shows the dynamic response of the feed compressor duty to the disturbance for a de-rated total toluene feed rate of 248.5 kmol/h. The compressor duty increases to 99% for maintaining the gas pressure in the presence of the disturbance. In view of the possibility of this unmeasured disturbance, the plant must be operated at this de-rated operating condition all the time even when the feed gas purity is 95 mol% hydrogen i.e. no disturbance. The benzene production rate corresponding to a de-rated total toluene flow of 248.5 kmol/h for 95 mol% fresh hydrogen is 156.8 kmol/hr and the steady-state operating profit is calculated as $249.88 per hour. These values represent a significant decrease from the maximum production rate and operating profit values of 170.8 kmol/h and $305.62 per hour reported earlier.

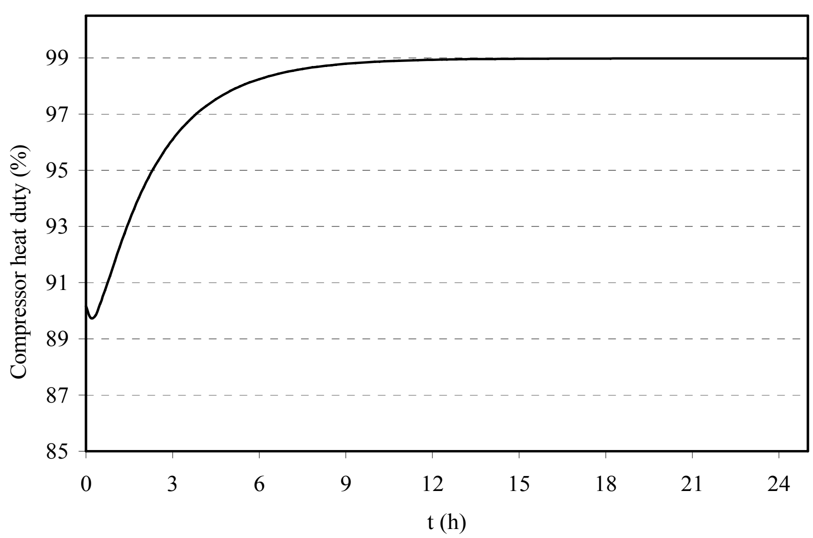


Figure 4 Dynamic response of feed compressor duty to 5 mol% step decrease in feed hydrogen purity with de-rated total toluene set-point.

As explained, the problem of de-rating can be largely overcome by using VPC. The advantage of the VPC arrangement is that now through-put de-rating for disturbances is automated so that in the absence of the disturbance, the VPC automatically adjusts the through-put manipulator to increase production. Without the VPC, this increase is not achieved since the through-put manipulator remains at its de-rated set-point.

The de-rating necessary depends on the magnitude of the worst-case disturbance. Figure 5 plots the steady state benzene production rate and the total toluene feed rate (through-put manipulator) as the fresh hydrogen feed composition varies from 90 mol% to 100 mol% in steps of 2.5 mol% with and without VPC. It is seen from the Figure that the production rate remains nearly constant at 156.8 kmol/h for de-rated plant operation without VPC as the feed hydrogen composition changes. This is because the total toluene feed rate remains fixed at its de-rated value of 248.5 kmol/h. Automated adjustment of the total toluene set-point using VPC causes the production rate to increase as the hydrogen composition increases due to an increase in the total toluene feed rate. At the base-case fresh hydrogen feed composition of 95%, this increase in benzene production rate is a significant 14.0 kmol/h over the de-rated benzene production rate. The total toluene feed rate increases to 262.8 kmol/h. Note that feed hydrogen composition is expected to be around its base-case 95 mol% value most of the time except during short duration process upsets on the upstream hydrogen feed side. The VPC approach thus ensures that higher through-puts are achieved during normal plant operation with no disturbances. Table 3 tabulates the steady state plant operating costs and profit data with and without VPC as the feed hydrogen composition is varied. The increase in profit over de-rated plant operation is also noted in the Table. As expected, the incremental profit using VPC increases from zero for 90% hydrogen (worst-case disturbance) to \$55.74 per hour for 95% pure hydrogen (no disturbance) to \$110.98 per hour for 100% pure hydrogen. At the base-case hydrogen composition, where the plant is expected to operate all the time barring process upsets, the operating profit is a significant 22.4 % more than the profit for de-rated plant operation.

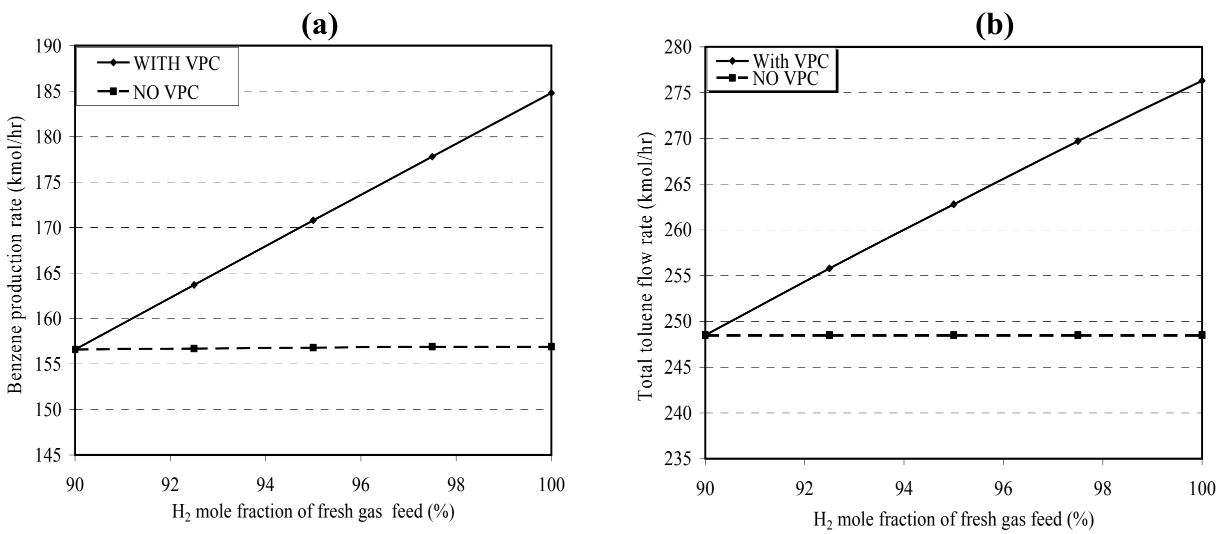


Figure 5 Variation in benzene production and total toluene rate with feed gas composition for VPC based and de-rated operation with feed compressor bottleneck
(a) Benzene production rate
(b) Total toluene rate

Table 3 Steady state plant operating cost and profit data for VPC based and de-rated operation as the mole fraction of H_2 feed is varied with the feed compressor as bottleneck

S No	With VPC				De-rated Operation (No VPC)			
	Fresh H_2 mol fraction (%)	Product Price – Raw Material Cost ($/hr)	Utility Cost ($/hr)	Operating Profit ($/hr)	Product Price – Raw Material Cost ($/hr)	Utility Cost ($/hr)	Operating Profit ($/hr)	Incremental Profit* ($/hr)
1	90	803.22	558.44	245.07	803.07	558	245.07	0
2	92.5	840.05	564.99	275.06	805.74	557.53	248.12	26.94
3	95	876.29	570.67	305.62	808.46	558.58	249.88	55.74
4	97.5	912.02	576.35	335.67	811.52	557.12	254.4	81.27
5	100	949.71	583.25	366.46	812.15	556.67	255.48	110.98

*: Incremental Profit = Operating Profit with VPC – Operating Profit for De-rated Operation

Conclusions

This work has demonstrated the application of a valve positioning control type approach for maximizing plant through-put for enhanced plant operating profitability. Results from the HDA process case study for the hydrogen feed compressor and the furnace bottleneck clearly demonstrate that *apriori* knowledge of the bottleneck constraint can be exploited to automate the manipulation of the through-put manipulator using a VPC to continuously operate the plant close to the bottleneck constraint. Quantitative results show that using the VPC approach, the plant operating profit can be increased by significantly by 10% or more. VPC is thus a simple and effective way for improving plant profitability by maximizing plant through-put.

Acknowledgements

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