

Marginal Cost Analysis in Process Decision Making

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Abstract

The marginal cost of an extra tonne of production is generally significantly different from the average cost of production. It can be much higher or lower depending on circumstances and, importantly, more production may not always mean more profit. Analysis of marginal costs can highlight and address what are effectively economic decision points in plant operations, capital projects and plant design. This analysis is also important with regard to equipment condition and availability and can be used to guide the allocation of maintenance resources to minimize production costs. Marginal cost analysis is based on a comprehensive process model of the plant combined with a cost model. A well-tuned plant model will give accurate information about the relative differences in raw materials and energy (RME) between different designs, plant configurations or operating targets. The model RME utilizations for different scenarios are input to the cost model that includes detailed fixed and variable operating costs. Comparison of scenarios with a base case will give accurate marginal cost information. Scenarios may be very detailed and include variations in individual equipment performance and availability that can be used to build an equipment sensitivity matrix for big data analytics. This method of analysis gives a true measure of the bottom line impact of any process change which can be used to select appropriate technologies and allocate limited resources to minimize production cost within real constraints. It is a proven tool in refinery management and in mature plants has allowed for continual production increase and cost reduction without major capital expenditure.

Keywords: Process simulation, cost analysis, resource allocation, marginal cost.

1. Introduction

An alumina refinery is in many ways a marvel of process engineering - a typical plant has hundreds of unit operations, thousands of pieces of equipment, many kilometers of pipes, wiring and conveyers, a big plant may have in the order of a billion liters of Bayer liquor and over 50,000 tonnes of solids inventory in the active circuit. These large complex operations produce a single product which has to meet a strict range of chemical and physical specifications and they need to be competitive in a global commodities market.

There is a strong incentive to reduce manufacturing costs and increase output. This can be done with large capital projects and brownfield expansions. It can also be done with small projects and the optimization of operating parameters, controls strategies and allocation of maintenance resources. Making smart choices of how to allocate limited resources over time will reduce costs and increase output with little or no capital expenditure. Many small improvements can add up to give significant benefits and there are a number of plants around the world that have done just that. The trick is to identify, justify and prioritize the improvements.

The effects of small process changes are generally not directly measurable in a plant and require a detailed process model in order to evaluate their effects. The relative benefits also need to be evaluated using *marginal* production costs rather than average production costs. Marginal cost

is often very different than the average cost and this can strongly skew the potential benefits when comparing projects.

The Wikipedia definition for marginal cost is: “*In economics, marginal cost is the change in the opportunity cost that arises when the quantity produced is incremented by one unit, that is, it is the cost of producing one more unit of a good.*” In a refinery the question is “how much did the extra tonnes cost to make” and in general, the answer is “a lot different than the average cost”. The focus of this paper is on quantifying the effect of small changes on a process and using that information to optimize plant operations.

2. Measurement

One characteristic of the Bayer process is that the liquor continuously circulates around the process. There are also many inner recirculation loops within the big outer loop. The consequence of this is that any change anywhere in the process will affect everything else everywhere in the process. The changes may be in ways that are not necessarily expected and may be counter intuitive. This means that in order to evaluate the effect of some change on a process you need to be able to look at what happens across the entire process and for small changes, you need to be able to accurately model small differences in RME use across the plant and add them up to get the real net change.

The measurement of process changes in a plant is not straight forward. A refinery is a harsh environment for instrumentation and the size of the equipment and the magnitude of the flows make accurate measurements difficult. It is easy to accurately measure a 1 kL/h flow – it is hard to accurately measure a 1,000 kL/h flow. In addition, it is not always clear when to measure. There are a range of time scales between cause and effect. It takes something in the order of days for the liquor to complete a full trip around the circuit, it takes about a day for solids to pass through precipitation and most are recirculated again, charge pump to blow off tank is an hour or two, transit through the heat interchange department (HID) is in the order of minutes and recirculating flows can transfer the effects of process changes to other parts of the process more quickly. For example, if there was a sudden change in test tank caustic, digestion would see this change much sooner than the back end of precipitation. Aside from any intentional process changes, there are continual changes in a process due to equipment condition, variation in bauxite quality, lag times in normal plant controls, equipment swaps, etc.

Plant data tends to be noisy enough so that is difficult to see the effect of minor process changes.

In a plant making 5 000 t/d of alumina, a difference of 25 t/d cannot be readily measured. This might be discernible in long term data, but not on a daily basis. If you measured 25 t/d difference in production (which in this case is 0.5 %) you cannot know whether that is measurement error in the weight belt, material coming out of precipitation tank inventory, a change in charge pump flow from 5 days ago, a slight drift in charge control instrumentation, a small change in evaporation rate 3 days ago or the optimization of some control parameters to make the plant run more efficiently. An important point here is that 25 t/d extra production – for whatever reason – is worth in the order of \$1 M USD per year in bottom line profit.

Maintenance resources are always limited and there is a constant need to prioritize those resources for both planned and unplanned maintenance. Safety and plant integrity should always be at the top of the list, but beyond that some maintenance jobs have a larger return than others in terms of production rates or cost. Some of these changes might be noticed in an operating plant if they were long term but not in the short term. For example, permanently removing a precipitation tank from service would have a measurable impact, but if a tank comes out of

service for a week, the impact may not be easily observable above the normal plant background noise. However, there is an opportunity cost for not having that equipment in service.

Small changes in process operating set points, equipment condition and control strategies can make small, yet significant differences in plant performance. Many of these changes will be too small to measure directly or attribute to specific actions - yet they are real, they do make a difference to the bottom line profitability of the plant and they are an important aspect of plant management.

3. Evaluation and Modeling

The majority of the production cost of alumina is RME. There are people, maintenance and administration costs as well, but for small changes, these are going to be more or less constant and the modeling here is focused on RME. Evaluating RME changes must include the overall effect on consumption across the plant. An improvement in one area of the plant may decrease the performance in another area enough to offset any potential benefits.

3.1. Modeling

The evaluation of the benefits of process changes must be based on the *incremental* cost of the extra production rather than the average cost of production. To do that requires an accurate measure of the effect of process changes and the only realistic way to get that data is with a comprehensive plant model that is able to do accurate mass and energy balancing. The analysis calls for a steady state full plant model that includes the full circuit as well as power or steam plants. Steam and electricity generation are an integral part of the process and need to be explicitly included in the plant model.

The model needs to be tuned to match plant performance data. This is a difficult task in itself since refineries are never actually at steady state – changes in physical condition of equipment, changing bauxite composition, equipment or instrument failures, changing control parameters, etc. all introduce variation into the process. The base case set of data for tuning a model should ideally be for a relatively long and stable period of operation. Process simulation models can be used to make good estimates of plant performance. However, it is worth noting that model predictions of production for a given snapshot of plant conditions are usually plus or minus a few percent. This is because plants are never really quite at steady state and because process data has so much noise that you don't know how accurate the snapshot is to start with.

What models are very good at is evaluating the difference between two scenarios. While overall accuracy of a process model is reasonably good, the accuracy of the difference between two scenarios is generally very good and this is what is important for evaluating the incremental effects of process changes and determining marginal costs.

It is important that the model is set up so that it is free to respond to changes in inputs. The model needs to include:

- fundamental chemical reactions with extents based on chemical equilibria or known valid correlations.
- detailed impurities chemistry and the effect on solubilities, boiling point elevations (BPE) and precipitation rates.
- the effects of scaling on heat transfer coefficients – heat recovery has a large impact on energy costs.
- environmental heat loss throughout the process.
- side streams that are accurately modelled – carrier water for flocculent, SRT, caustic ingress, hose water, etc all have a significant impact.

- impurities balances around the circuit and across tears.
- a precipitation rate equation that covers the full range of process conditions.

For example, pump off A/C should not be a fixed input value, rather it should be calculated as a function of precipitation conditions, process parameters and liquor chemistry.

3.1.1. The Model

We used SysCAD to model an imaginary full alumina refinery with a powerhouse generating steam and power for the plant. The plant we modeled is not a real plant (we are very conscious of client IP) but it is modeled in the same detail and manner that we would typically use for existing plants. The demonstration plant was a high temperature plant with sweetening, oxalate co-precipitation and a nominal capacity of about 4 000 t/d. It is a mix of well-known technologies to make it a relevant example. All of the data, correlations, flows, RME prices, etc. that were used in the model were taken from the public domain – there is no proprietary information in the model. The table below gives an overview of the plant properties and Figure 1 shows a screen shot.

This paper is not about modeling packages – any comprehensive mass and energy balancing software could be used. However, it is important that the model includes the entire process circuit, can fully balance all flows in and out of the process and across any recycles and has predictive capability based on correct and accurate physical models. It is also worth noting that this paper is not about plant design - the demonstration plant design is neither modern nor particularly good – it is simply used for illustration.

Table 1. Plant overview.

Plant Parameter	Typical Value
Bauxite	46.2 % TAA, 43.2 % THA RxSi 2.8 %, 1 % Qtz, 0.7% CaCO ₃
Sweetening Bauxite	49 % TAA, 41% THA, RxSi 4%, 1% Qtz, 0.8 % CaCO ₃
Digestion	10 Flash tanks, 9 regen htrs, 1 inj htr, 1 cond htr
Digestion	1800 kL/h SPL flow, 245 °C Tdig, 4.4 MPa steam to digester
Charge	25 pt dig margin, 0.770 Last FT A/C
Trim and Dilution	225 gpL BOT C, 0.720 BOT A/C, 19 t/h BOT steam
Export	66 t/h FT7 evap, 69 t/h FT8 evap and desilication
Settler	500 gpL UF solids, 5 points reversion
Washers	7 washers, 500 – 540 gpL UF solids, 85% Scandrett eff
HID	Plate heat exchanger
Precipitation	1 row, 4 000 kL Tanks, 4 agglom., 17 growth, 1 pump off 8 coolers, 47 MW duty, 18 MW env heat loss, 64.8°C PO T 210 gpL 1 st agglom solids, 450 gpL PO solids 63.7 gpL yield, 2.6% scale loss
Classification	Gravity primary thickener, seed cyclone
Seed Prep	FS deliq., FW wash, CS deliq. drum filters
Product	Deliq. plus 2 stage wash drum filters
Evaporation	3 units, 7 regen. flash, 2 baro stages, 2 export htrs 1 live steam htr, 130 °C, 60 °C Tbaro
Powerhouse	4 x 250 t/h boilers, 10.25 MPa, 2 topping turbines, 2 extraction turbines, 40 kPa condenser, BFW, AC and FDID fan turbines, 4.4 MPa, 650 kPa and 275 kPa steam to plant

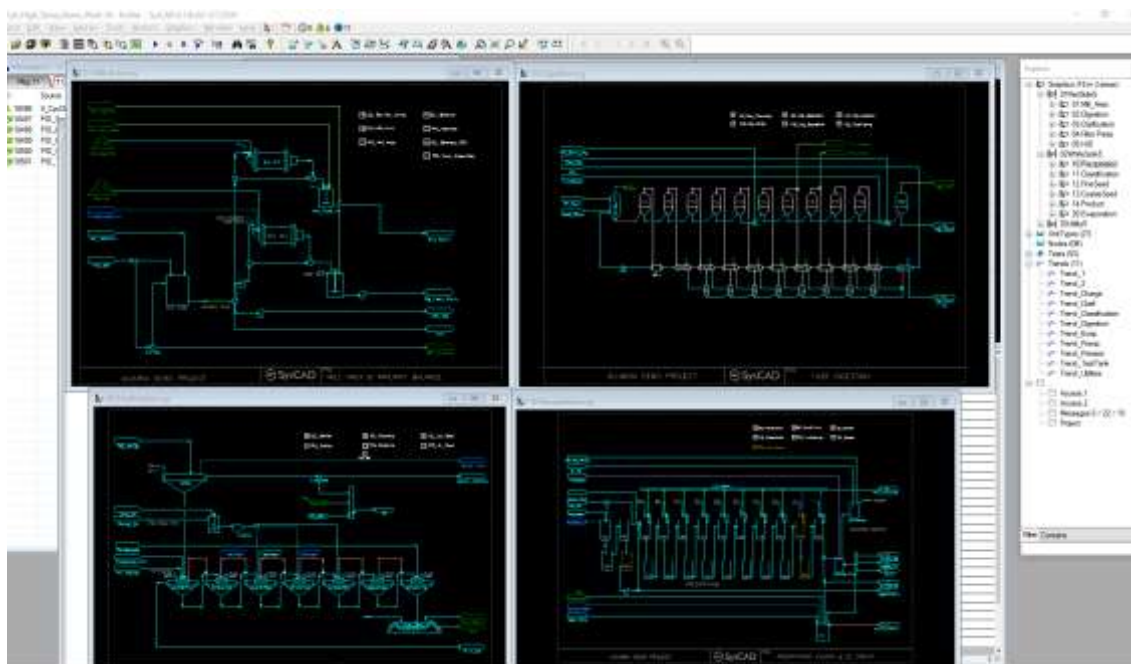


Figure 1. Screen shot of SysCAD model.

Control of liquor chemistry is important in evaluating process changes. The concentrations of some solutes in the liquor change very slowly while others change quickly. It usually takes many months for TOC to change, while caustic, A/C and oxalate can change within a day. The levels of chloride, sulfate and carbonate tend to stay at more or less constant levels with consistent bauxite and caustic composition. In this study we were looking at relatively small process changes over shorter time periods so we selectively held some solutes constant. The model was set up with the main process tear (recycle) at the exit of the test tank and had the option to selectively close the loop on individual test tank properties. This allowed some solutes to be held constant while others were adjusted continuously. Test tank temperature, caustic, A/C, silica and oxalate were always closed loop. Solute that were held constant were adjusted to maintain fixed ratios of solute to caustic – the amount of water in a circuit may change and thus the caustic concentration, but caustic and solute masses in the circuit are relatively constant, so the ratios do not change. The Cl/C, SO₄/C and C/S ratios were controlled by adjusting the relative incorporation of these impurities into DSP. TOC was held to a fixed level by adjusting the adsorption of NOOC onto mud. Once the model was tuned, there were only very small adjustments required to maintain the target levels for different scenarios. The relative difference in the total mass flow and the mass flows of individual species were all within 0.001 % across the recycle tear. This method of chemistry control is appropriate for this application – different techniques may be required to look at greenfield design or major process or bauxite source changes.

Table 2. Control of liquor chemistry.

Parameter	Adjustment	Control Method	Typical Test Tank Value
Caustic	Continuous closed loop	Process parameters	260 gpL
A/C	Continuous closed loop	Process parameters	0.400 – 0.770
C/S	Fixed ratio	Carbonate to DSP	0.89
Cl/C	Fixed ratio	Chloride to DSP	~ 1.4 gpL NaCl
SO ₄ /C	Fixed ratio	Sulfate to DSP	~ 1 gpL Na ₂ SO ₄
Si/C	Continuous closed loop	Silica to DSP	~ 1.6 gpL Na ₂ SiO ₃
TOC/C	Fixed ratio	NOOC on mud	20 gpL C
Oxalate/C	Continuous closed loop	Process parameters	~ 1.6 gpL Na ₂ C ₂ O ₄

Caustic was primarily added via cleaning caustic makeup. Caustic addition to the plant was modeled by adding caustic to balance the total amount of elemental sodium into and out of the process. A scaling rate was calculated for the base case and this rate was used for all the other cases. Process parameters like digester charge margin were assumed to be at practical limits for the process and variations were only taken in the direction of safer operation. When making a change in a parameter, all the other process set points and equipment conditions were held constant.

4. Cost Model

The cost model in this example looked only at differences in RME costs between different scenarios. This implicitly assumed people, maintenance and other costs did not change with minor differences in operation. More detailed cost models might also include people and maintenance costs broken down into fixed and variable costs, maintenance might include cost detail for specific equipment, operation might include minor consumables such as grinding media and flocculent. We also assumed that RME costs were constant, however these costs may be variable depending on the contract. For example, the cost for fuel may reduce beyond a certain annual usage or imported power cost may increase above a certain load. These variations can be important factors in determining incremental costs and need to be included in a cost model. Cost models need to be reviewed with accounting and finance personnel to make sure they capture the relevant information and in some plants accounting “owns” the cost model.

The construction of a cost model was done in Excel. The SysCAD model was used to calculate production and RME usages and the data exported to the Excel spreadsheet. The Excel spreadsheet had entries for RME prices. The table below gives the nominal prices for RME into the plant. Obviously, these costs can vary substantially based on plant location, current market conditions, tariffs and where the plant is located relative to the mine site. We used natural gas as the base case fuel for all calculations.

Table 3. Raw materials and energy costs.

Item	Nominal Price (USD)
Process water	\$0.18 / t
Demineralized Water	\$0.40 / t
Burnt Lime (as CaO)	\$140.00 / t
Caustic (as NaOH)	\$500.00 / t
Bauxite (dry)	\$80.00 / t
Sweetening Bauxite (dry)	\$85.00 / t
Fuel Energy, Coal (GJ)	\$2.00 / GJ
Fuel Energy, Natural Gas (GJ)	\$3.00 / GJ
Fuel Energy, Oil (GJ)	\$10.00 / GJ

The calculations in the cost model are straight forward. A SysCAD plant model was tuned to a period of stable plant operation and this data was used as a base case. The model was then run for different scenarios and these data compared against the base case. The average RME cost is simply the sum of the raw materials costs divided by the production. This tells some of the story but looking at the incremental costs adds another level of detail. The calculation of the marginal cost is simply the incremental cost divided by the incremental production. The total incremental cost or the incremental cost of any specific RME item can be found as;

$$\text{Marginal Cost} = \frac{\text{Incremental Cost}}{\text{Incremental Production}} \quad (1)$$

More detail about what is happening comes out when the costs are broken down by item. While in this exercise we are looking only at RME, the incremental cost can include people, administration, maintenance or other relevant costs. For example, they could include a penalty cost for not meeting supply contracts.

5. Examples Scenarios – Variations to Base Case

For an example calculation we added another precipitation growth tank and compared the results against the base case. Table 4 below shows the production and usage rates, the RME cost per hour, the difference between the two scenarios and the average production cost. With an extra precipitation tank and everything else held constant the yield increases by 0.9 gpl due to the increased residence time and the production rate increases by 2.6 t/h (62 t/d). Although the usages increase, the average production cost decreases by \$0.46 /t.

Table 4. Production, usage and cost.

Rate	Base Case Usages	Additional Precip Tank Usages	Difference
Production (t/h)	170.87	173.45	2.57
Process water(t/h)	373.23	373.88	0.52
Demin Water(t/h)	209.86	213.31	3.44
Lime(t/h)	4.07	4.09	0.01
Caustic(t/h)	18.03	18.34	0.31
Bauxite(t/h)	275.84	280.72	4.87
Sweetening Bauxite(t/h)	131.68	132.58	0.90
Fuel energy (MW)	497.59	499.51	1.91
Cost	Base Case Costs	Additional Precip Tank Costs	Difference
Process water (\$/h)	67.18	67.30	0.12
Demin Water (\$/h)	83.94	85.32	1.38
Lime (\$/h)	569.11	572.62	3.51
Caustic (\$/h)	9013.05	9169.37	156.32
Bauxite (\$/h)	22067.33	22457.63	390.30
Sweetening Bauxite (\$/h)	11192.61	11269.33	76.73
Fuel energy (\$/h)	5374.00	5394.68	20.68
Total cost (\$/h)	48367.23	49016.26	649.03
Avg Prod cost (\$/t)	283.06	282.60	-0.46

Table 5 below shows the same information broken down to show the marginal costs which are the difference in RME costs per hour divided by the difference in production per hour. This shows that the extra tonnes have a very different cost than the average cost - the RME cost for the extra tonnes is \$30.81 less than the average cost for the base case (with \$3/GJ natural gas). There are a number of reasons for this, but the basic reason is that production increase brought about by a yield increase has a lower energy cost than the average cost - it's the same amount of liquor through digestion which consumes about the same amount of energy but gives more production. The table below shows that the energy consumption is 2.7 GJ/t compared with the base case average of 10.5 GJ/t. There are other changes in the incremental usages which reflect small changes across the plant. Caustic use goes up slightly with increased mud load, a lower test tank A/C increases bauxite to the digester for the same margin. Wash water demand does not change so there is almost no increase in process water.

Table 6 below shows the differences between the base case and the extra precipitation tank

scenario. There are only slight differences between the two cases and nothing that would be easily measurable, especially if the duration of the event was only a few days. There is slightly more yield with the increased residence time. There are slightly more solids in the last washer which increases the flow of liquor out of the last washer. This is offset somewhat by a little more water coming in with the increased bauxite flow, but the overall effect with a fixed evaporation rate is a slight increase in test tank caustic.

Table 5. Marginal cost.

Specific Usage	Base Case Usage tonne	Incremental Usage per tonne
Process water(t/t)	2.184	0.254
Demin Water(t/t)	1.228	1.341
Lime(t/t)	0.024	0.010
Caustic(t/t)	0.105	0.122
Bauxite(t/t)	1.614	1.896
Sweetening Bauxite(t/t)	0.771	0.351
Fuel energy (GJ/t)	10.483	2.679
Cost	Base Case Costs per tonne	Marginal Costs per tonne
Process water (\$/t)	0.39	0.05
Demin Water (\$/t)	0.49	0.54
Lime (\$/t)	3.33	1.37
Caustic (\$/t)	52.75	60.75
Bauxite (\$/t)	129.15	151.69
Sweetening Bauxite (\$/t)	65.50	29.82
Fuel energy (\$/t)	31.45	8.04
Production cost (\$/t)	283.06	252.25

Table 6. Scenario comparison.

	Base Case	Extra Precip Tk	Difference
Precip Residence time (h)	26.42	27.60	1.18
Precipitation yield (gpl)	63.68	64.57	0.89
Yield extra Precip Tk (gpl)	-	1.0	1.0
LWUF Caustic (gpl)	9.32	9.57	0.25
Water in LWUF (t/h)	273.54	276.91	3.37
Water in Bauxite (t/h)	66.34	67.28	0.94
Test Tank A/C	0.411	0.407	0.004
Test Tank Caustic (gpl)	262.50	262.71	0.21

We can generate table of costs for tanks out of service. We have calculated a profit difference by taking the difference between the average base case RME cost and the marginal cost for the incremental production. The actual profit would include the margin between the sale price and the average production cost and for this example we have assumed a profit of \$50 per tonne for the base case using natural gas at \$3/GJ. The table below shows the profit difference for different fuel types. With expensive fuel, process changes that increase yield are much more favored.

It is worth noting here that an extra precipitation tank on-line has a significant value and that value is realized as soon as the tank comes on-line. The value of a tank on-line depends on fuel costs, but even for cheap fuel it is worth in the order of \$5 000 per day. This information could be used to justify additional resources to accelerate the turn around on equipment.

Table 7. Precipitation tank scenarios.

	Minus 3 Tanks	Minus 2 Tanks	Minus 1 Tank	Base Case	Extra Tank
Production rate (t/h)	162.13	165.20	168.13	170.87	173.45
Average Cost (\$/t)	284.94	284.24	283.61	283.06	282.57
Marginal Cost (\$/t)	248.27	248.75	249.27	-	250.14
Production Difference (t/d)	-209.90	-136.15	-65.82	-	+61.80
Profit Difference, Coal (\$/d)	-16 146	-10 408	-4 998	-	4 508
Profit Difference, Nat Gas (\$/d)	-17 797	-11 477	-5 514	-	4 990
Profit Difference, Oil (\$/d)	-29 359	-18 965	-9 127	-	8 364

In practical terms, this allows us to assign a real cost to having a precipitation tank in service when the changes to the process are too small to measure. We can do this same exercise for a whole range of plant equipment and operating parameters and use this information to compare and prioritize maintenance work to get the most benefit out of the resources available. This approach is also very useful for evaluating proposed projects, including small capital cost projects with changes to equipment, the process, piping, etc. Project economics can be heavily influenced when marginal costs are significantly different than the average costs.

5.1. Further Examples

The calculations work the same whether the production is increased or decreased by some change. There is now a marginal cost for the decrease in production and this marginal cost is what should be used for evaluating the impact. If the tonnes that are not made have a lower production cost than the average cost, then the loss for not making those tonnes is greater than if they had the average cost. The calculations can evaluate what it is worth to get back to the base case from current state; as well as improving on the current state. In the precipitation example above adding a tank was worth about \$2 000 per day extra and removing a tank cost about \$2 200 per day which is the value realized for recovering a tank when down by one. The table below gives some examples of other process changes and the effect on plant performance. Precipitation cooling and evaporation make cheaper tonnes than average, pushing charge flow makes slightly more expensive tonnes and a washer offline makes almost no change in production, but costs a lot in caustic. There are no surprises here, but having an accurate cost can help manage the situation.

Table 8. Marginal costs – relative to \$283.06 /t RME and \$50 /t nominal profit.

	Production Difference (t/d)	Marginal Cost (\$/t)	Cost Δ to avg (\$/t)	Δ profit (\$/d)
1 Precip cooler off-line	-42	268.55	-14.51	-2 682
2 Precip coolers off-line	-87	267.08	-15.98	-5 768
Charge pump -100 kL/h	-140	285.30	2.24	-6 701
Charge pump +100 kL/h	+138	287.01	3.95	6 353
Evaporation Rate -5t/h	-7	172.78	-110.28	-1 127
Evaporation Rate +5t/h	+7	176.19	-106.87	1 100

There are some changes that do not make a difference to production rates, but still change significantly RME costs – sometimes in subtle ways. An example of this is cake solids in fine seed deliquoring. The cake moisture in the base case is 20% at a caustic of 80 gpL. This is 100 t/h of liquor going to the oxalate dissolver tank. This must be diluted with condensate addition to a target B-filtrate of 35 gpL. When the cake moisture is reduced, less condensate addition is required and there is less flow to oxalate destruction and ultimately less flow to the washer train. Wash water increases to balance this reduction in flow and the caustic and alumina losses in last

washer underflow decrease. The flows to digestion and precipitation are virtually unchanged so the production changes only very slightly (0.11 t/h). However, the reduced caustic and alumina loss from the last washer reduce RME cost by \$3 120 per day which is a little over \$1 M per annum. Here is a situation where something like vacuum pump performance or filter cloth condition which have virtually no impact on production and make an almost unnoticeable change in wash water flow can change the bottom line by a million dollars per year. The only way to see this and quantify it is by modeling. Some constant production results are shown below.

Table 9. Production cost differences at constant production.

	Production Difference (t/d)	Difference in RME cost (\$/d)	Difference in avg cost (\$/t)
Fine seed deliq moisture 20→ 19%	0	-3 120	-0.58
High Press FT off-line	0	+5 601	1.42
Low Press FT off-line	0	+1 505	0.36
Washer Off-line	0	+12 521	2.77
Dig HTC poor	0	+1 916	0.48
Dig HTC good	0	-2 076	-0.52
Tdig -2 (243) °C	0	4 247	0.91
Tdig +2 (247) °C	0	-4 077	-0.87
Digester Margin +10 pts	0	+6 656	1.38
Digester Margin +20 pts	0	+13 374	2.77

We can also use this approach to calculate steam costs. The incremental steam costs are found by adding small extra steam demands to the base load and calculating the change in RME. These costs can vary a lot depending on the mode the plant is operating in. The table below shows steam and power costs for various fuel types. The table also shows the value of reducing condenser pressure which might be accomplished by cleaning heat exchangers, improving ejector operation or cooling tower maintenance.

Table 10. Steam and power marginal costs.

Utility	Cost @ \$2/GJ	Cost @ \$3/GJ	Cost @ \$10/GJ
4.4 MPa steam (\$/t)	5.27	7.69	24.64
650 kPa steam (\$/t)	3.30	4.74	14.79
275 kPa steam (\$/t)	2.51	3.58	11.04
Electric Power (\$/MW-h)	34.62	51.86	172.55
Condenser Pressure 40 → 35 kPa	230 \$/d	345 \$/d	1 155 \$/d
Condenser Pressure 40 → 30 kPa	479 \$/d	719 \$/d	2 399 \$/d

This approach can also be used to find the cost of water ingress into the circuit. With detailed balances it is possible to calculate the difference that small ingress streams make to the process. The cost of ingress depends on where the water comes into the circuit. For example, 1 kL/h of water ingress to the settler backs out almost exactly 1 kL/h of wash water and LWUF caustic increases by 0.05 t/h. Production does not change, but the extra caustic loss translates into a production cost difference of \$44.53 per day which is about \$16 k per year in bottom line loss while the same ingress into the second washer would cost about \$63 k per year in lost profit.

6. Incorporation into Practice

The technique of calculating a production difference and marginal cost relative to a base case can be expanded to include a series of base cases that cover different plant operational modes –

for example, a boiler off and steam restrictions, an evaporation unit off, different quality bauxite, etc. The same process model and cost model can be used for all of these situations. The different base cases can be stored as scenarios and reloaded as needed. The calculations can be made as deviation from plan and this can be done on a weekly basis to provide support for short term planning.

The simulation model with its reports, calculations and collection of base cases should be maintained and improved on an ongoing basis. This requires adequate allocation of human resources, procedures and standards for the modeling program and regular use of the model. A key point is that an investment in building and maintaining a good quality plant model together with best practices and procedures for using the model, can provide a bottom line benefit for a range of applications in operating a refinery. With a commitment to the ongoing use of modeling at a refinery, there can be significant ROI with benefits for production, quality, reducing costs and understanding of process plant operational relationships. Golubev [1] and Franco [2] give some good examples of modeling use in process optimization.

7. Conclusions

Plant data in an operating refinery is noisy and instrumentation often lacks the accuracy necessary to sensibly analyze the effects of small changes to a process. This makes it difficult to use plant data for making decisions to improve production and/or reduce costs. However, with the “digital twin” of a good plant model, analysis is readily available to make and prioritize decisions that have the most effect on the bottom line for an operating refinery.

This paper focuses on marginal cost analysis for RME to improve operational profitability, but the same simulation model and technique can be used for justifying and prioritizing small capital cost projects for process equipment changes or improvements. A further example of use for this technique is in evaluating alternate or improved process control strategies and/or operator procedures.

These techniques can be taken further through automation of running a large number of scenarios for producing significantly large data sets and then using novel data analytics and reporting (visualization) software to gain insights into refinery operations and optimization. A possible future progression, now possible with emerging technology, is to use model simulation data together with plant data in a range of interesting AI applications.

8. References

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