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Viva Energy's Geelong refinery reduces FCCU turnaround risk

Change involves risk. Many fluidized catalytic cracking units (FCCUs) can be operated more profitably, but changes to achieve more efficient operations can be risky. Gains realized by modifying a stable but suboptimal process can be quickly erased if unforeseen, adverse effects are experienced. Such was the case for Viva Energy's Geelong refinery.

The Geelong refinery (**FIG. 1**), formerly part of the Royal Dutch Shell group, is one of four refineries in Australia and employs more than 700 people. The refinery processes approximately 120,000 bpd of crude and supplies more than 50% of Victoria's—and 10% of Australia's—fuel demand. The refinery also provides feedstocks for the neighboring LyondellBasell polypropylene plant.

The Geelong refinery's 40,000-bpd residue catalytic cracker unit (RCCU), built in 1992, underwent multiple hardware changes during a turnaround in 2011. Following startup, afterburn (defined here as the temperature difference between the regenerator dense phase and the flue gas line) became prevalent. The full-burn unit was operated very close to the flue gas temperature constraint, but the operators were most concerned about frequent, dynamic flue gas temperature spikes; the temperatures would suddenly and unexpectedly rise, requiring immediate and frequent operator intervention. These incidents caused the refiner to ultimately reduce rates by almost 10% for certain feedstocks, impacting overall refinery economics by tens of thousands of dollars per day. The dynamic afterburn events showed no correlation with any measured process conditions, even after the utilization of advanced analytic techniques by refinery engineers.

In preparation for its 2016 turnaround, engineers began to ask questions during planning discussions. What really happened in 2011? Multiple hardware changes were made, but process conditions also changed over time. If the hardware changes were the primary cause, was it one change or the convolution of all changes that combined to decrease robustness and stability? Furthermore, if additional changes were planned, would they help, or would they perhaps make things worse?

These questions prompted refinery staff to utilize simulation to:

- Aid in understanding the root cause of underperformance
- Perform virtual testing of intended hardware changes
- Provide confidence in the likelihood of success
- Identify opportunities for further optimization
- Minimize the risk of unintended, adverse effects of the planned modifications.

In short, simulation was used to reduce risk. Historical, current and then-future configurations were tested via 3D, transient simulation of the gas catalyst hydrodynamics, thermal behavior and coke combustion kinetics.[a]

FIG. 1. The Geelong refinery in Australia.

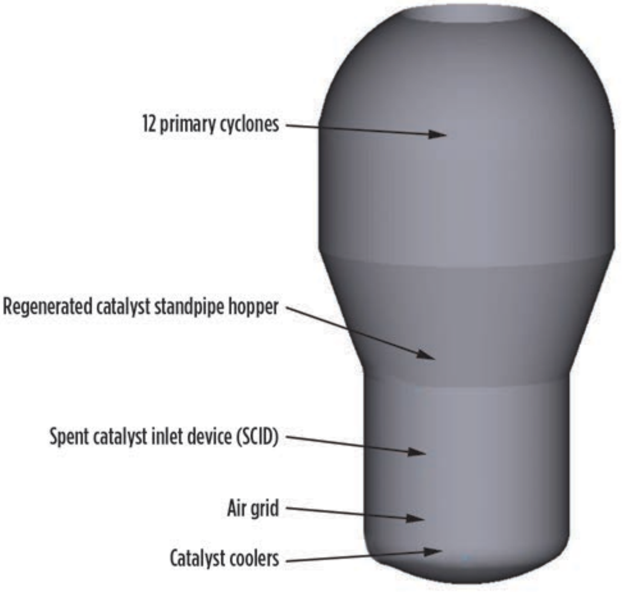

FIG. 2. Regenerator model domain.

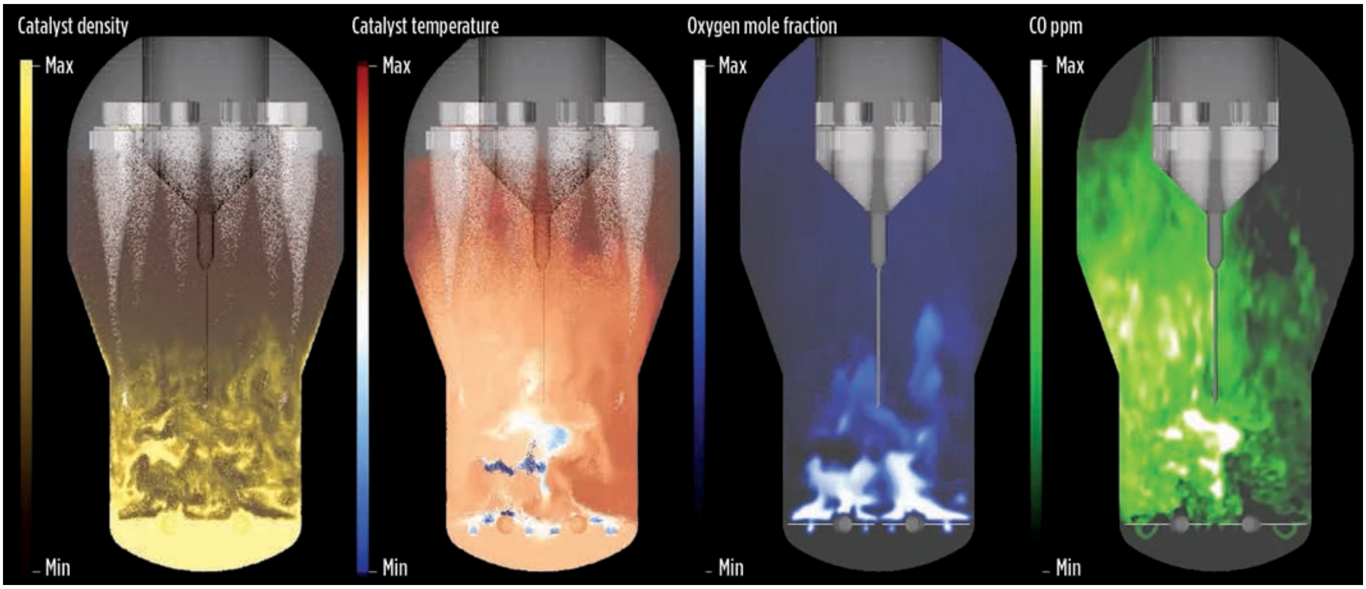

FIG. 3. Baseline simulation results.

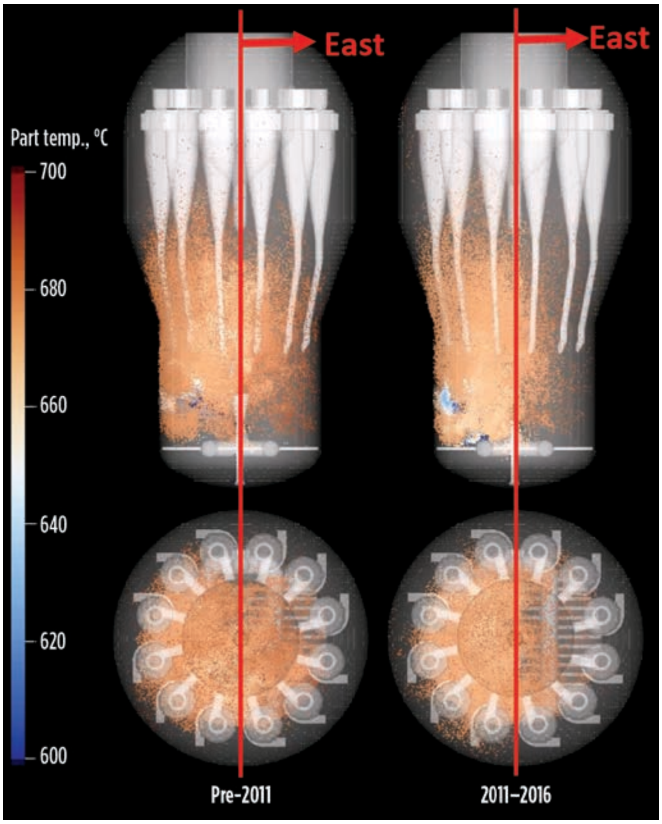

FIG. 4. Maldistribution of spent catalyst (pre-2011 and post-2011).

The computational model domain for the Geelong refinery's RCCU regenerator is shown in **FIG. 2.** While details of internals cannot be shown, the major components are listed. Twelve primary cyclones are distributed around the circumference of the upper vessel, with diplegs returning catalyst to the regenerator bed. A single regenerated catalyst standpipe hopper is used to withdraw regenerated catalyst near the top of the bed. Spent catalyst from the reactor stripper enters the regenerator in the middle of the bed via a spent catalyst inlet device (SCID). The air grid is present, with model boundary conditions defined for each air grid nozzle. Finally, two catalyst coolers withdraw catalyst from the dense bed, remove heat and return catalyst to the bottom of the vessel.

Before simulating potential changes, the model was first used to understand present and historical operations. The primary purpose of the baseline model was to identify the root cause of the afterburn, thereby enabling targeted changes and a benchmark against which future improvements could be quantified in a virtual environment. Both pre-2011 and post-2011 operations were simulated before decisions were finalized regarding the 2016 turnaround.

TABLE 1 shows a summary of the changes made in 2011. Design changes included the installation of a new SCID, as well as the removal of the hoppers at the top of the catalyst coolers. Around the same time, operating conditions changed significantly. Following the turnaround, the catalyst circulation was increased by 4.4%, and the air rate was increased by nearly 15%. The SCID alterations necessitated major changes to SCID aeration, while minor changes to cat cooler operations were also made.

With so many changes, the question remained: Which change, if any, would prove to be the root cause? The regenerator experienced both design and operational changes in the periods before and after the 2011 turnaround. Which of these caused the increased afterburn and dynamic temperature spikes resulting in frequent panel operator intervention, conservatism and, ultimately, reduced throughput?

Sample results from the baseline model are shown in **FIG. 3.** The figure is a snapshot in time from the transient simulation results. The view on the far left in **FIG. 3** shows catalyst colored by density. Several distinct fluidization zones are observed. Poor fluidization is observed below the air manifold, as expected. A well-fluidized dense bed is visible from the air grid up to an elevation around the area change. Above this elevation, the bubbles and turbulent mixing behavior observed within the dense phase give way to a splash zone and subsequent dilute phase, which persists up to the cyclone inlet elevations. All gas exits the regenerator at the cyclone inlets, and virtually no catalyst is present in the upper dome.

The second view from the left in **FIG. 3** shows the same catalyst particles colored by temperature. Several observations can be easily made. The particles are relatively cool at the air grid near the bottom and on the left, or west, side corresponding to the spent catalyst inlet location. The catalyst quickly heats to the regenerator operating temperature as it is mixed and as the coke combusts. The bed temperatures vary in both the radial and axial directions, with the highest particle temperatures present in the upper dilute phase, where the density is low. If combustion occurs with less particle mass present, then the resulting temperature rise is greater.

The second view from the right in **FIG. 3** shows oxygen (O_2) on a centerline slice through the model. O_2 is high where injected, and it decreases with elevation. The O_2 is observed to break through the dense bed in bubbles, and excess O_2 is apparent in the dilute phase through the top of the model.

The right-most frame in **FIG. 3** shows carbon monoxide (CO) on the same centerline slice. This view is particularly telling; much more CO is present on the west side of the unit, and a significant amount of CO enters the cyclones on the west. In practical terms, this full-burn regenerator operates somewhat like a partial-burn unit on the west side.

The CO imbalance provided Viva Energy engineers with their first clue to the potential root cause of the afterburn problem. While it may be obvious that temperature spikes in the flue gas line are related to localized combustion of CO with O_2, the mechanism for the presence of CO, despite high concentrations of excess O_2, was becoming clearer. While a significant volume of O_2 is present, it is not evenly mixed with the CO in the unit, resulting in downstream combustion. However, the question remained: Why was the CO higher on the west than on the east?

To answer this question, the distribution of spent catalyst was investigated. **FIG. 4** shows only the spent catalyst with a residence time of less than 10 sec, colored by temperature. An elevation view is shown on the top, and a plan view is shown on the bottom. The pre-2011 and post-2011 configurations are compared on the left and right of the figure, respectively.

TABLE 1. Summary of 2011 design and operational changes

	2011 changes*
Design	
SCID	New design
Cat cooler	Hoppers removed
Operational	
Catalytic circulation	+4.4%
Air rate	+14.7%
SCID aeration	Major changes
Cat cooler	Minor changes

*Changes relative to pre-2011 baseline configuration and operating conditions

Visually, **FIG. 4** suggested that the pre-2011 configuration did a better job of rapidly distributing the spent catalyst across the regenerator. However, with simulation results, a quantitative analysis for each computational particle is possible and was undertaken. The quantitative results were more telling. While only 21% of the pre-2011 particles with less than 10 sec of residence time were found on the east side of the unit, the post-2011 case was much worse; only 11% of the post-2011 particles crossed the unit centerline in less than 10 sec.

Based on these results, it was concluded that the 2011 changes introduced more maldistribution of air and catalyst in the regenerator. Put another way, the post-2011 case performed about half as well at distributing the spent catalyst all the way across the unit.

This maldistribution clearly affected the elevated CO concentrations observed on the west side, nearest to the SCID, but it also affected O_2 much higher in the regenerator. **FIG. 5** shows the O_2 mole fraction on a cut plane at the cyclone elevations. The increase in O_2 maldistribution in the post-2011 simulation is apparent, further enabling the dynamic temperature spikes.

While no data is available for gas compositions within the regenerator itself, the flue gas compositions were measured further downstream and compared with simulation results exiting the regenerator, as shown in **FIG. 6** for both the pre-2011 and post-2011 configurations. Both the magnitude and trends agree between the simulations and measured data. The slightly higher excess O_2 concentrations predicted by the simulation are expected; the simulation composition is measured at the cyclone inlets, whereas the refinery measurements are downstream of the third-stage separator. Additional combustion in the flue gas line will lower the simulation O_2 concen-

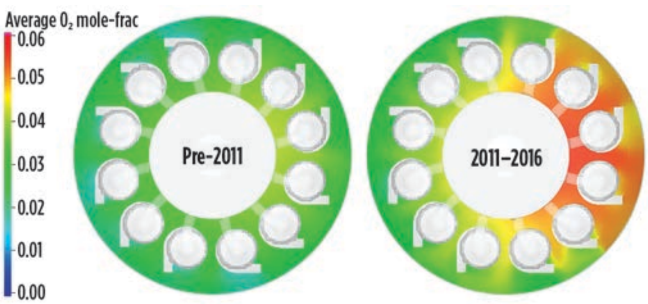

FIG. 5. Maldistribution of excess O_2 (pre-2011 and post-2011).

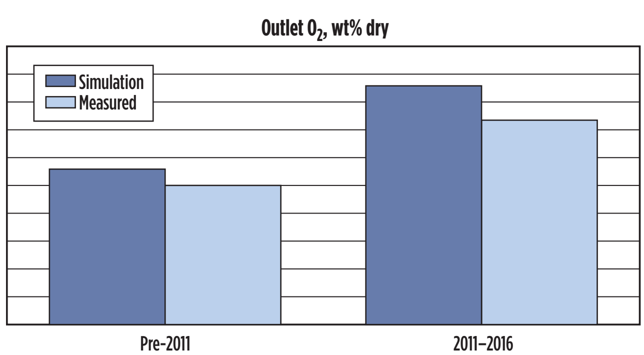

FIG. 6. Excess O_2 validation.

trations between these two locations, as evidenced by the observed afterburn.

The excess O_2 validation showed that the simulation results reflected reality. Excess O_2 was mixing with CO in the plenum, resulting in the reduced performance, and the root cause was much lower in the vessel, likely the SCID. The ability of the model to match historical observations gave refinery engineers confidence that the same model could predict the effects of future changes.

The model was then applied for simulations of changes proposed for the 2016 turnaround. **TABLE 2** shows a summary of changes planned for 2016. Design changes included reverting the SCID to the pre-2011 configuration and some optimization of air grid nozzles (but no change to the catalyst cooler inlet hoppers). Operationally, it was expected that the higher catalyst circulation would be maintained, but the air rate could be decreased slightly. No changes to catalyst cooler operations were planned, but reverting the SCID design necessitated changes to SCID aeration.

TABLE 2. Summary of 2016 design and operational changes

	2011 changes*	2016 changes*
Design		
SCID	New design	Same as pre-2011
Cat cooler	Hoppers removed	Hoppers removed
Air grid	Unchanged	Nozzle optimization
Operational		
Catalyst circulation	+4.4%	+4.4%
Air rate	+14.7%	+13.7%
SCID aeration	Major changes	Similar to pre-2011
Cat cooler	Small changes	Same as 2011–2016

*Changes relative to pre-2011 baseline configuration and operating conditions

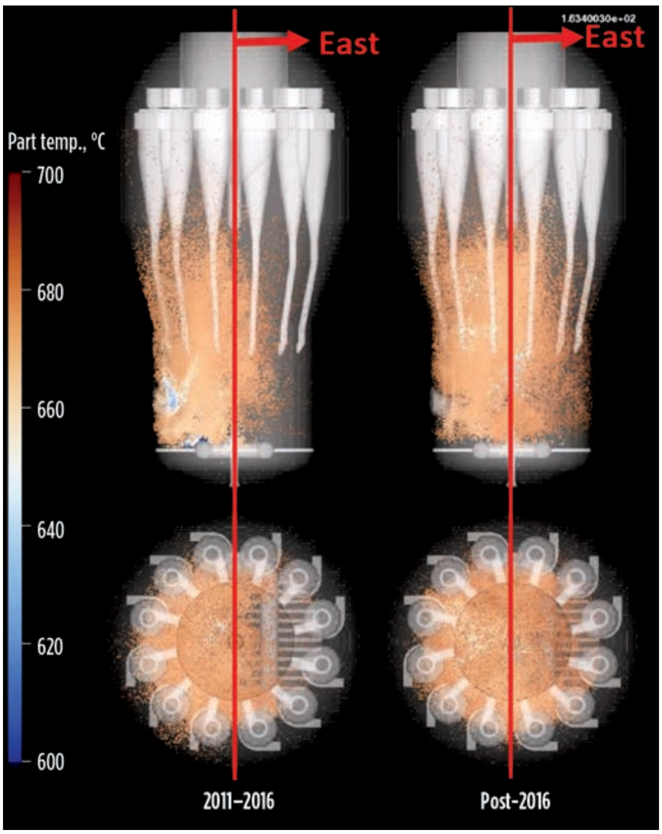

FIG. 7. Improved spent catalyst distribution.

FIG. 7 shows the effect of these changes on spent catalyst distribution. Similar to **FIG. 4, FIG. 7** shows only the spent catalyst with a residence time of less than 10 sec, colored by temperature. The 2011–2016 configuration is shown on the left, with the post-2016 case on the right. Visually, an improvement was expected. However, a quantitative analysis revealed that the percentage of spent catalyst that would reach the east side of the unit within 10 sec would increase from 11% to nearly 25%—similar to the pre-2011 performance. Subsequent analysis showed that similar improvements were expected to reduce the gas maldistribution entering the cyclones.

In short, simulation showed that the changes were expected to address the major root cause of the maldistribution that was constraining the system operation, and gave the Geelong refinery staff confidence in the planned changes in advance of the 2016 turnaround. All simulated changes were implemented in 2016.

Success was confirmed via post-turnaround operational data. The data from the 12-month periods immediately before and after the 2016 turnaround revealed that, following startup, the average afterburn was 5°C lower, the maximum daily throughput was 4% higher and the number of panel operator interventions was reduced by 75%. Afterburn is no longer a regular topic of operational meetings at the Geelong refinery.

Viva Energy's Geelong refinery reduced turnaround risk not just by using simulation, but by using it well. Lessons learned and illustrated by the Geelong refinery case study, which can be applied by any operator of FCCUs, include:

- **Ask good questions.** The Geelong refinery team identified specific questions to ask during simulation and created a model suitable for answering those questions.
- **Use the right simulation tool.** By correctly identifying the RCCU regenerator design as the culprit, the team then selected a simulation package suitable for capturing the relevant gas catalyst hydrodynamics, thermal behavior and coke combustion kinetics at industrial scale within pragmatic time constraints.
- **Seek to understand root cause before jumping to conclusions.** Geelong refinery engineers first sought to understand the root cause of underperformance. Understanding why something happens often helps determine how to address it.
- **Simulate current or historical operations before virtual testing of potential changes.** Baseline modeling is an important first step. Without the baseline simulation, what did the 25% mixing number mean? Only by comparing it with the 11% mixing was it clear that the post-2016 design would result in a meaningful improvement. Baseline models enable direct comparisons of changes and aid in the interpretation of results. Additionally, they are useful for model validation, calibration and refinement, if warranted.
- **Test proposed modifications.** In this case study, the planned changes were successful. However, in many

cases, some changes are helpful and others are not. Simulation can decouple the effects of multiple changes to design or process conditions individually. Early simulation results often provide insights into additional potential for improvement, which should also be simulated.

- **Ensure that all stakeholders are part of the process.** Geelong refinery engineers combined outside simulation expertise with their own knowledge of refinery operations to ensure success. In many cases, even more stakeholders are involved, such as a central engineering group, technology licensors, hardware vendors, catalyst suppliers, consultants or outside research/testing companies. Working together, whenever possible, often leads to better outcomes than working in isolation.
- **Plan early.** Geelong refinery staff allowed sufficient time to act on simulation results. Additional optimization is under consideration now, well in advance of the next turnaround.
- **Be proactive.** Turnaround planning is a natural way to start with simulation for many refiners, as was the case for Viva Energy. However, it is not necessary to wait for a turnaround to start planning. Having a baseline model ready enables rapid response to surprise behavior at any time in the FCCU operational cycle, and is a vital component to every refinery's digitalization strategy. HP

NOTES

[a] All simulations were performed using Barracuda Virtual Reactor from CPFD LLC.

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