



# Balancing the Modern Challenge of Operating Aluminium Smelters—Minimizing Energy Consumption, Minimizing Greenhouse Gas Emissions, and Maximizing the Productivity of Assets

Barry Welch, Jie Bao, Sergey Akhmetov, Pablo Navarro, Gudrun Saevarsdottir, and Halvor Kvande

## Abstract

By applying fundamental data and developing simplified correlations while assuming uniform conditions in the cells, the aluminium industry has been able to design better cells and steadily improve the environmental performance, energy, and faradaic efficiencies. However, with the high capital and operating costs of the technology's retrofits aimed at increasing the productivity of installed capacity, the retrofitted cells have not always been able to sustain their previous environmental standards and production efficiencies. Hence challenges have emerged, with it becoming evident that the changed "physics" such as those contributing to magneto-hydrodynamic stability, alumina dissolution, electrolyte mixing, spatial heat generation and its transfer can all contribute to the deterioration in some of the key performance indicators and operating cost factors. Advancing the technology further is likely to be reliant on developing the design operating and control practices that solve these challenges.

## Keywords

Energy minimization • Cell productivity • Spatial GHG emissions

## Introduction

The development of electrical energy generation technology in the early part of the nineteenth century opened the path for electrochemical and electrothermal extraction of the more reactive metals by the use of electrically conducting molten salts and slags. Aluminium was produced electrochemically in 1854 from a molten mixture of NaCl–AlCl<sub>3</sub>—a time that preceded the development of the understanding of the electrochemical science that was subsequently developed by Gibbs and other researchers in Europe. By the start of the twentieth century, the economic virtues of the Hall–Héroult process, combined with the special properties of aluminium heralded the start of a continuing growth in aluminium extraction, and today it is second only to iron of the metals produced. Throughout the first half of its history, the development was dominated by cell design changes to accommodate the semi-batch operations of reactive materials introduction that was coupled with the continuous metal production. In the early period, there was limited transmission grid infrastructure, so smelters developed in regions where hydro-generation could be easily installed. Thus numerous electrothermal and aluminium smelting extraction processes commenced operation in the fjords of Norway, and just 30 years after the invention of the process there were six different smelters operating in Norway producing a combined total of 4000 tonnes of Al per month with cells of varying designs. While five of these smelters have closed, following World War II with six new and significantly larger smelter sites Norway's capacity now exceeds 115,000 tonnes of Al per month—using green energy.

B. Welch (✉)

University of Auckland, Auckland, New Zealand  
e-mail: [barry@barry.co.nz](mailto:barry@barry.co.nz)

J. Bao

School of Chemical Engineering, University of NSW,  
Sydney, Australia

S. Akhmetov

Emirates Global Aluminium, Dubai, United Arab Emirates

P. Navarro

Aluar, Puerto Madryn, Argentina

G. Saevarsdottir

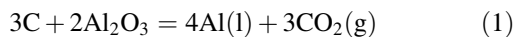
Department of Engineering,  
Reykjavik University, Reykjavik, Iceland

H. Kvande

Norwegian University of Science and Technology,  
Trondheim, Norway

Accompanying the Norwegian post World War II smelting expansion, NTNU had established one of the important centres for advancing the scientific understanding of metal extraction from molten salts and slags, especially for electrowinning aluminium [1, 2]. However, appreciation of the importance of science and physics (such as thermal excursions, electrolyte flow and mixing, alumina dissolution kinetics) continued to lag behind changes in cell design. Grjotheim [2] and Bredig [3] demonstrated that co-deposited Na dissolved in the electrolytes containing an equivalent concentration of NaF. Subsequently, other studies showed that reoxidation of the dissolved metal by CO<sub>2</sub> of the anode gas was a major contributor to the loss in faradaic efficiency [4], the rate being controlled by the mass transfer of the dissolved metal from the metal pad to the anode gas [5]. Other early studies indicated that the current efficiency (CE) underwent an exponential decay rate at short anode-cathode distances [6].

With the growth in smelter size, harmful environmental issues were becoming evident. The fluoride content of emissions being released from the cells and entrapped in the fjords adversely harmed surrounding vegetation. This highlighted the limitations of simplifications in the interpretation of cell reactions (which still continues), such as representing the electrochemical reaction for metal production solely by the equation:



Such an approach ignores the science that results in electrochemical reactions contributing to the varying proportions of CO(g), gaseous fluorides (HF, CF<sub>4</sub>, C<sub>2</sub>F<sub>6</sub>), and other gases coevolved during the range of cell operating conditions [7].

In the case of aluminium smelting and other electrowinning processes, electrochemistry is used to enable reactions that will not otherwise occur spontaneously under the conditions. However, electrochemical enabling requires an ionically conducting electrolyte, and electronically conducting electrode materials that can have their interfacial electrode energy level modified by either depleting or increasing the electron concentration sufficiently to shift the materials energy level to the point where either a single or multiple charge transfer reactions (oxidation at one electrode and reduction at the other) can occur! However, electron transfer does not necessarily form the final product, and classical chemical reaction steps involving structural rearrangements, energy, or materials mass transfer can be involved. These additional steps can also be aided by extra interfacial energy by increasing the potential gradient or current density. Depending on the interfacial potential gradient other “enabling” electron transfer interactions can also occur. Furthermore, the entropic energy requirement to

complete the formation of each product in the phase can independently exist and must also be satisfied in some way. Any required energy to form the final product phase can be supplied either by an increased interfacial potential accompanying the current flow, heat transfer, or a combination of both. Thus, parallel reactions do occur at various rates depending on cell conditions, and by ignoring them gives errors in interpretation and consequential mitigation actions. The science must be respected!

---

## 1960 Prebaked Anode Cell Design and Operation

The size of cells had increased to ones typically sustaining 60,000 A, operating in the anode current density band 1–1.2 A cm<sup>-2</sup>. With the demand for the metal increasing through the evolving commercial aircraft industry and other applications further capacity growth was needed. Because electrochemical reactions are inherently slow, the ~5 m<sup>2</sup> reacting cathode surfaces produced less than 450 kg of aluminium per day and there was a need for even bigger cells to overcome the energy inefficiencies arising from the high heat loss for small cells. But challenges had already emerged for further increase in size, mainly to be able to control the cell dynamics, and minimise the impact the intense magnetic field generated by the current-carrying busbars has on the metal pad stability and consequential performance. The semi-batch addition of consumable materials operation required the introduction of the equivalent of ~250 kg cold carbon anode and ~860 kg of smelter grade alumina at room temperature each day. These consumables were added in batch-wise manners, bringing about variation in the range of process conditions. The frequency of alumina replenishment varied between once every 2–8 h with the concentration of dissolved alumina swinging between the saturated value (typically about 9–12 wt.%) and that leading to an anode effect (~2 wt.% depending on operating conditions). The alumina feeding provided a surge of thermal energy demand that was greater than 40 kWh and this had to be extracted from a limited fraction of the total liquid electrolyte mass quickly to prevent sludge formation. As we know now, it would have to come in contact with more than 25% of the electrolyte in the cell that was at a superheat greater than 10 °C. Consequently, the electrolyte had to contain less than 3 wt.% excess aluminium fluoride in order to achieve high alumina saturation solubility and the operating temperature had to be too high to form any side freeze. The absence of side freeze led to erosion of the rammed cathode sidewall introduced another cell failure mechanism and cell lives were typically less than 900 days.

## Cell Performance Indicators and Control at ~60 kA

Typical key performance indicators of the more modern prebaked cells operating in the early 1960s were as follows:

- Operating Temperature (°C):  $980 \pm 10$
- Excess  $\text{AlF}_3$  Concentration (wt.%): 2–4
- Current Efficiency (CE% theoretical Al):  $84 \pm 2$
- Cell-to-Cell Voltage (V):  $4.6 \pm 0.1$
- Energy Efficiency (DC kWh/kg Al):  $16 \pm 1$
- Average Cell Life (days):  $900 \pm 50$
- Anode Effects (AEs)/cell day: 2–4
- Electrolyte Mass (kg/kA): >60
- Gross Carbon (kg C/kg Al):  $0.55 \pm 0.03$
- Net Carbon (kg C/kg Al):  $0.44 \pm 0.03$ .

Essentially, the anode effect was used for alumina feed control activation. The anode effect had also been considered to be useful for cleaning the cathode surface from undissolved alumina sludge. To quote the diary of a trainee operator in the 1940s [8].

*I understand that under ordinary operations we plan to try to have "lights" 5 to 6 hrs apart.*

During an anode effect the thermal heat generation typically increased threefold, and over a 24-h period for the normal 3 AEs/day an extra ~110 kWh of thermal energy would be introduced in surges. AEs also brought about significant changes in the electrochemical anodic reactions products.  $\text{CO}(\text{g})$  then becomes the dominant product, and the proportion of  $\text{CF}_4(\text{g})$  evolved tended to match the  $\text{CO}_2(\text{g})$  [9].

## Emerging Issues from 1960s Design

Several weaknesses were evident, especially in view of the rapidly growing demand for the metal.

- Cell operation was labour intensive, and work functions were at the limit of manual capability.
- Infrastructure was expensive and needed to be made more cost-effective with larger cells.
- The health and environmental impact of the emissions from the ignored reactions on the surrounding environment was bad, so the continued use of a buffer land zone around the smelter complex was no longer a solution.

- The faradaic and electrical energy efficiencies had considerable room for improvement, as was hinted from some of the fundamental studies, and with the emergence of solid-state electronics greater and more scientific automation of the cell control was needed.

Bigger cells incorporating new more automated features were needed as an aid to overcome many of the listed weaknesses. But future designs needed to firstly resolve the problems created by the combination of magnetic field, metal pad stability, and the dynamics arising from the work practices.

Furthermore, when one looks at the cell energy balance relationship:

$$\begin{aligned} \text{Cell-to-Cell Energy Input (kW)} \\ &= \text{Energy heat and convert consumables to all the products} \\ &+ \text{Cell heat loss} + \text{Interconnecting busbar cell energy wastage} \\ &= V_{\text{cellto cell}} \cdot I_{\text{line}} \end{aligned} \quad (2)$$

There is also an obvious need to reduce the cell's ohmic resistance, or substantially increase the current efficiency so that heat balance is maintained, since the productivity increase for larger cells is not matched by a proportionate increase in heat loss.

While normal cell operation targets constant line current, because of voltage limits in the potline power supply, whenever multiple anode effects occur simultaneously the potline current decreases, magnifying the dynamics of the cell. Limited cell temperature control was simply being a judgemental voltage change. Superheats were too high for side freeze formation, and cell failures were high due to a combination of erosion of the rammed carbon sidewall or the consequences of high cathode heave caused by the sodium uptake of cathode blocks. The impact of the harmful fluoride emissions was contained by smelters owning a substantial buffer zone of land around and downwind of the smelter. Faradaic efficiencies were in the mid to low 80%, and the energy consumption typically exceeded 16 kWh/kg Al.

Attempts to go above the typical amperage were thwarted by the metal pad instability generated by the magneto-hydrodynamic forces arising from the busbar configuration. Furthermore, the individual anode block size was starting to exceed 150 kg, making block changing difficult and signalling the need for mechanization.

In 1960, it was an industry devoid of detailed scientific understanding and facing magnifying challenges with the need to have a larger production capacity of cells.

## Knowledge Advances (1950–1980) and Consequential Design Changes Enabling Cell Size Growth

In parallel with the design operating and control advances, more fundamental knowledge was being generated by both industry, R&D departments, and selected universities. The information formed the basis for the next generation of designs evolving in the late 1970s. The following are some of the changes achieved.

**Magnetic compensation.** Magnetic field measurements and associated model developments for different cell-to-cell busbar arrangements led to innovative redesigns of the way the current could be introduced to the anodes from adjacent cells.

**Superstructure redesign.** New cell superstructure designs incorporated:

- Multiple volumetric alumina additions ports (usually in the centre channel) with an automated crust breaker to enable programmable controlled alumina feeding.
- Anode change used machinery to both break the crust and position the new anode, thus overcoming constraints on size.
- Simultaneously the superstructure was changed to enable capture of the cell gases so that they could be subsequently treated to remove most of the harmful emissions.

**Alumina consumption monitoring.** By combining detailed anode overpotential measurements as a function of a range of cell variables [10] it was shown [11] that for any solvent electrolyte composition, temperature, and operating inter-electrode distance a minimum cell voltage occurred at a cell-specific alumina concentration for each combination of operating variables. The rate and magnitude of change of voltage around that minimum also followed a distinct pattern, so tracking voltage changes following controlled alumina feeds offered a path for triggering alumina feeds before the onset of an AE. However, this required minimizing sludge formation and allowance for the dynamics of the process conditions arising from work practices. A barrier to simple control logic, however, was the unwritten law that cells should have several anode effects each day! Potline voltage reduction and consequently line current and cell-to-cell voltage would lower when multiple anode effects occurred simultaneously on a potline. Without compensation, this prevented effective voltage tracking. An effective but technically erroneous compensation method was used to overcome this giving birth to what is commonly referred to as “resistance control”, which automatically reduced the component of the voltage that generated the Joule heating.

**Current efficiency gains.** The fundamental knowledge associated with metal solubility showed that by increasing

the excess aluminium fluoride content of the electrolyte and lowering the temperature a current efficiency gain would be achieved if all other conditions were maintained. Since the anode–cathode distance (ACD) also impacts the current efficiencies, some of the materials used to construct cells were replaced by less resistive components enabling an increase in ACD without altering heat balance.

**Dry scrubbing development.** Initially, the anode gas emissions were treated by wet scrubbing, which tended to produce a by-product that was not readily recyclable, and presented other environmental problems. Thus, during the 1960s it was found that the smelter grade alumina chemisorbed gaseous hydrogen fluoride that was coevolved from the cell, as well as collecting vapours predominantly ( $\text{NaAlF}_4$ ) emitted from the high concentration aluminium fluoride electrolytes. These gases tended to be almost quantitatively captured, and when combined with the precise thermodynamic and vapour pressure data of electrolytes developed by Kvande [12] optimization of the electrolyte composition for minimum chemical loss was enabled, while also increasing current efficiency.

One of the first dry scrubbing systems in the world was installed in Årdal in 1967.

**Point feeder development.** The most important challenge was simultaneously controlling the cell heat balance and alumina dissolution to prevent undissolved alumina accumulating below the metal pad. Small and frequent additions through “point feeders” helped to minimise this problem when used in conjunction with a control strategy aimed at keeping the average concentration near the minimum of the alumina–voltage relationship.

**Lowering rate of Joule heat generation.** Since larger cells have less heat loss per unit metal production, in order to optimise productivity a reduction in either the current density or resistance of all current-carrying components used in cells was needed. Both approaches were used with the resistance being reduced through better material to material jointing and using better quality carbon for cathode construction. This development was occurring during the first global “energy crisis”, and therefore operations shifted to lower current densities, which resulted in an increase in anode–cathode distance and hence an increase in the ratio of kg electrolyte/kA. The latter was beneficial for providing more time for the alumina to dissolve and mix around in the cell and minimise concentration gradients.

**Extending cell life.** With alumina feeding and dissolution becoming less of a problem through the small mass additions, and the electrolyte having sufficient thermal energy capacity to minimize freezing on the cold alumina added, cells were able to operate at lower superheats. Coupling this with the lower energy input to the electrolyte, and relatively thin sidewalls resulted in a frozen layer being formed on the

inner side of the rammed cathode lining. This then presented a barrier to the erosion of the sidewall, which was a major contributor to early cell failure. Subsequently by focusing on controlling the superheat—typically below 10 °C, and minimising thermal excursions of the cell (which typically contributed to initiating cathode block cracking) the cathode lives were extended to in excess of 3000 days for some technologies!

Coupling the combination of the new generation of designs with better control and operating practices underpinned by basic science, the new technologies introduced in the 1980s were hovering around the 200 kA cell size, and more important performing under quality benchmark conditions that smelters would still be proud of today.

The following tabulation gives the band for the typical key performance indicators for technologies introduced in the late 1970s and early 1980s.

Operating amperage (kA)	175–220
Operating temperature (°C)	960 ± 5
Excess AlF <sub>3</sub> concentration (wt.%)	10.5 ± 1
Current efficiency (CE% theoretical Al)	93 ± 2
Cell-to-cell voltage (V)	4.1 ± 0.05
Energy efficiency (DC kWh/kg Al)	12.8 ± 0.3
Achievable cell life (days)	>2500
Anode effects (AEs)/cell day	0.3 ± 0.15
Electrolyte mass (kg/kA)	~55
Gross Carbon (kg C/kg Al)	0.52 ± 0.02
Net Carbon (kg C/kg Al)	0.42 ± 0.02

The improved alumina feeding strategy, coupled with better signal sensing for cell voltage tracking (although most of the industry refers to it as resistance tracking) reduced the tendency for cells to form sludge. With that change, the anode effect frequency steadily decreased. However, the ingrained importance attached to an anode effect for “cleaning up cells” took a long time to dissipate. Even though the industry became aware of the environmental harm associated with the PFCs coevolved during the anode effect, and started addressing options for mitigation, almost all control strategies existing in the late 1990s still have a scheduled anode effect available in their control logic—typically to achieve a minimum of one AE per week.

It should also be noted that during this, and also the more recent periods, the quality and capacity of the rectifying transformers used to supply the energy to potlines has continued to change. Whereas in the 1960s potlines were limited to approximately 90 cells and ~415 V DC, today potlines exceeding 400 cells at more than 2000 V DC exist. This has improved the energy efficiency by having less of the shared interconnecting busbars wastage between pot rooms and the power source.

## Studies of the “Physics” of Cell Operation

Each high amperage potline brings on a significant increment in metal production and involves a high capital expenditure. Consequently, capacity creep by increasing potline amperage with or without modifications to some of the adjustable design features, has always been a favoured way of increasing productivity. When retrofit changes are made one needs to question or adjust to the impact they may have on the rates of the various processes (the “physics”), which could bring about adverse spatial variation in cell conditions. Previously, largely through the high electrolyte mass per kA it was assumed that all conditions in the cell were constant. However, as shown by Martin et al. [13] for some technologies there has been a threefold decrease in the electrolyte mass per kA and a similar increase in the amperage being supported by each point feeder in some technologies. This increases the probability of spatial variations in cell conditions occurring. Hence the rates of all interfacial heat transfer, alumina dissolution kinetics, the flow and mixing of electrolyte, the metal pad stability through changed magneto-hydrodynamic fields, and the relative proportion of the various cell reactions occurring through changed current densities all suddenly become important but hitherto have been largely ignored. So what do we know about the physics of the cells?

**Electrolyte flow and mixing.** By adding tracers to the electrolyte at various locations in a cell with anode and electrolyte volume design features compatible with the leading 1980s technology, Purdie et al. [14] found that it could take up to 1 h for the electrolyte to homogenize in a cell operating with a high electrolyte volume and low productivity. Furthermore, it has clearly been demonstrated that the freeze formed on the undersurface of newly set anodes severely impacts the flow pattern of the electrolyte from the point feeders resulting in the zones of sufficiently low concentration of alumina that PFCs are being coevolved [15].

**Driving force for electrolyte flow and mixing.** Fortin et al. [16] used physical modelling of anode gas bubble behaviour to demonstrate how they coalesced and then contributed to the electrolyte flow pattern as they were released from the undersurface. This understanding was reinforced and extended by Chen et al. [17], who demonstrated a major influence generated by the pumping action of gas released in the backchannel on that heat transfer rate at the metal–electrolyte interface, and electrolyte mixing. These generated a weak point in the protection of the frozen ledge at the sidewall. The greater travel path for gas release with enlarged anodes also increased the bubble layer electrical resistance. Somewhat surprisingly, the models typically applied for electrolyte flow had not taken into account the gas release pattern and forces in determining its impact on electrolyte mixing.

**Retrofit of larger anodes.** Increasing the cross-sectional area of the anodes potentially enables higher amperage in the cell without increasing the anode potential and hence the range of reactions enabled. Unfortunately, this results in a substantial reduction in electrolyte volume per kA due to displacement of electrolyte by the cross-sectional area and the consequential necessary reduction in anode–cathode distance in order to maintain heat balance. Compounding these problems is the increased rate of depletion of the alumina concentration between feed shots. Hence, it introduces a greater need for both good electrolyte flow, and fast alumina dissolution.

**Slotted anodes.** Introduction of slots in anodes reduces the travel path for the gas, hence lowering the resident gas volume and the consequential resistance it generates [18]. This enabled an increase in anode–cathode distance and therefore current efficiency. While the electrolyte flow pattern was modified by concentrating the zone for gas release, the driving force for electrolyte mixing was still retained providing the slots were fully immersed in the electrolyte. This constraint led to compromises between the value of slots for reducing resistance and the importance of anode gas release and electrolyte mixing.

**Lowering cathodes total resistance.** Early attempts saw a shift to more graphitized cathode blocks. Without a change in collector bar and associated contact resistance, this resulted in a very high current density through the cathode block to the metal pad near its extremities. Combined with thermal conduction of adjacent materials, the resulting high heat generation in the outer zone made retention of ledge protection below the metal pad much more difficult, and there has been accelerated cathode wear and lowering average cathode life. More recently attention has been directed towards the current collector bar materials and design—lowering the combined ratio of the resistance of collector bar plus contact to carbon. The accelerated wear in the outer zone of the cathode block can be minimised. With better control of cell temperature, copper has re-emerged as the obvious substitute, and a range of designs have emerged. However, cell autopsies indicate that contact resistance is often higher than models predict, and there is a need for better models as well as giving greater attention to the role of collector bar fixing its shape and design.

**Anode current distribution.** The physics of the cell changes each time an anode is replaced, due to the need to re-melt the freeze formed on the introduction of the cold anode block. With the reduction in driving forces for electrolyte flow and mixing, and the higher thermal demand through the enlarged anode, this has magnified dynamics introduced by current distribution. For newly set anodes in cells with liquid electrolyte mass greater than 50 kg/kA, newly set anode would be drawing near full current in about 12 h. Recent measurements on high productivity cells

indicated that this time extends out to closer to 50 h, and the re-melt process of the freeze is less uniformly driven. Not only does this extend the period of cell imbalance, but also it tends to bring in concentration gradients across the width of the cell, with a consequential coevolution of PFCs that are not readily detected in the zone by the individual anode current measurement.

**PFC coevolution.** PFC coevolution initiates without the cell transgressing to conditions leading to those that result in an anode effect through the combination of reduced electrolyte mixing forces, reduced liquid electrolyte volume in the electrochemical active zone between anode and cathode, and a higher rate of depletion of the oxide through the higher anode current density. If the cross-cell flow from the point feeder to the sidewall results in a higher concentration of alumina at one end of the anode than the other, the interfacial current density will decrease in the low current zone if all other conditions are the same because of the increase in anode potential required to maintain the same rate for the low concentration. This results in the rate of consumption of the anode and hence anode–cathode distance in that zone decreasing. The current density slowly readjusts towards that on the rest of the anode to ultimately match the carbon consumption rate in all parts of the anode, since the electrolyte voltage drop through the resistance reduction accompanying the changed localised ACD enables the localised anode potential to increase. PFC coevolution, however, automatically initiates once the interfacial potential exceeds the critical value. While individual anode current measurements can give an indication of the onset of this problem, the rate and magnitude of changed total anode current are usually only within the noise level of measurements. When this does happen—and it is usually towards the extremity of the anode—the electrolyte becomes depleted in aluminium fluoride, and the electrolyte can transgress to the sodium-fluoride-rich side of the phase diagram. This then adds the complication that the oxide has an extremely low solubility and some can precipitate out. This combination then leads to the formation of the blue colored mineral diaoyudaoite ( $\text{Na}_2\text{O}\cdot 11\text{Al}_2\text{O}_3$ ) either on the anode, or the adjacent ledge. This is confirmed by a combination of photographic evidence (see Fig. 1) anode gas analysis and chemical and mineralogical analysis.

---

## Modern Technology and Cell Operation

Using the same principles that enable the 200 kA barrier to be breached, coupled with variations in cell control logic, anode change work practices, alumina feeding strategies, innovative materials selection and design features to reduce the cells' resistance further, high productivity cells operating at line currents near and above 500 kA have been developed.



**Fig. 1** Samples of the mineral diaoyudaoite removed from a modern cell. Upper from ledge sample and lower on sidewall end of anode butt. (Color figure online)

The designs have followed one of two distinct paths—either extending the length of the cell but otherwise retaining the general design features of the high performing 1980s technologies or increasing the width as well as the length. Performances generally have approached but not exceeded the benchmark values previously achieved.

Problems of spatial variation in process operating conditions have become more evident. While good operations and control can minimise the process excursions, modern cells have a greater tendency to co-evolve PFCs in a manner that can only be detected by anode gas analysis. The environmental impact of this is probably in question because hitherto the industry's approach has been to reference the evolution based on the more easily measured anode effect frequency and duration.

## Addressing the Global Challenges Aluminium Smelting Faces

Aluminium is playing an important role in helping address the global issue of climate change, especially in helping making safe lighter weight electric vehicles. So the demand for aluminium will continue to grow.

While we are well aware of the environmental harm that results from PFC coevolution and the need to reduce the global GHG contributions from the industry [19], we have

good models to estimate the amount being released during anode effects, and more important leading control strategies keep this level very low [20]. In fact, they can be so low that their PFC emissions are below the variability cells have in emissions from net carbon consumption. The issue with PFCs is the undetectable low-level coevolution arising from the spatial variations in cell conditions leading to concentration gradients enabling different reactions in different zones of anodes without giving a measurable indicative signal.

Since the physics causing the variations is known, design and operating practice changes are needed to prevent the onset. More updated electrolyte flow models are needed. As are answers to the question—why should we feed at the same rate on the same pattern in each feeder location? Do we need distributed feeding [21] and control logic that adjusts to a more sophisticated analysis of the variation in cell physics following work practices? With developments of online distributed alumina concentration estimation techniques [22], distributed feed rate control approaches based on dynamic mass balance models can be useful [23].

Another simple and meaningful action would be to use a more realistic average electrochemical voltage when calculating resistances for corrections. The use of the value 1.65 V is still fairly widespread. This was appropriate at the time it was introduced because the average alumina concentration was higher and the anode current density less. Today the electrochemical potential is much higher and typically above 1.75 V and under conditions where the rate of change in voltage with alumina concentration or temperature is vastly different—especially in view of the reduced electrolyte volume.

It is still necessary to use a carbon anode, which is criticized in some quarters—despite it being one of the most energy-efficient uses of any carbonaceous materials. But as has been emphasized recently [19], the greater issue revolves around the limitations for further hydro-electricity generation expansion. The greatest environmental problem rests with the use of coal and high molecular weight hydrocarbons to drive the generators.

Solving all these issues requires a better understanding of the physics of the operation of cells.

## Conclusions

Aluminium smelting has advanced to becoming the second most important metal because of the role it plays in improving society's quality of life. The growth has been achieved by applying control and design changes that better match the underlying science. While aluminium has an ever-growing demand, its extraction will always be dependent on electrical energy—a resource that is also required by

many of the technological advances that improve our quality of life. Since global warming is a reality, it is important that not only the energy be used to its maximum efficiency, but its generation should not harm the environment. Hence, further improvements in energy efficiency are needed. The scientific fundamentals of the process are a constraint, so the emerging challenge is to modify control and design practices so that they minimise the adverse effects that arise through the hitherto ignored “physics” of the process.

**Acknowledgements** Part of the success of the design, control, and productivity advances has been by merging fundamental knowledge with practical and operating constraints. This has been aided by bridging the gap between the science, the design, and physics of the process such as through TMS and University promoted training courses led by Halvor Kvande, A uniquely qualified person because throughout his career he has been involved simultaneously in fundamental R&D and industrial cell operation, and therefore has an appreciation of both sides of the equation!

## References

- Flood H, Førlund T, Grjotheim K (1954) Über den Zusammenhang zwischen Konzentrationen und Aktivitäten in geschmolzenen Salzmischungen (Activities in Reciprocal Molten Salt). *Z. Anorg. Chem.*, 276(5–6): 289–315.
- Grjotheim K (1956) Contribution to the Theory of the Aluminium Electrolysis, *Kgl. Norske Videnskabers Selskabs Skrifter*, Nr. 5, Trondheim, Norway, 90 pp.
- Bredig MA (1964) *Molten Salt Chemistry*, ed Blander M, Interscience Publishers, New York, pp. 367–425.
- Pearson TG, Waddington J (1947) Current Efficiency in Al Electrowinning. *Discussions of Faraday Soc.*, No. 1, p 307.
- Gjerstad S, Welch BJ (1964) The Oxidation of Aluminium by Carbon Dioxide in Presence of Cryolite Electrolyte, *J. Electrochemical Soc.*, 111: 976–980.
- Pearson TG (1955) *The Chemical Background of the Aluminium Industry*. Royal Inst. Chem. Lectures, London, No. 3, 103 pp.
- Wong D, Welch BJ (2018) PFCs & Anode Products – Myths, Minimisation and IPCC Method Updates to Quantify the Environmental Impact. Paper 4c1, Proc. 12th Australasian Aluminium Conference, Queenstown, New Zealand.
- Mills RJ (1941) *Personal Diary Notes for Cell Startup and Operation*.
- Tabereaux AT, Richards NE (1995) Composition of reduction cell gas during normal conditions and anode effects. In: Evans, JW (ed) *Light Metals 1995*, The Minerals, Metals & Materials Society, pp. 325–333.
- Richards NE, Welch BJ (1963) Anode Overvoltage in Electrolysis of Cryolite-Alumina Melts. *Extractive Metallurgy of Aluminium*, Vol. 2, Aluminium (Ed. G. Gerard) pp. 15–30 (Interscience Publishers, New York).
- Welch BJ (1965) Technology of Electrolyte Reduction of Alumina by Hall-Héroult Process: I. A Voltage Analysis Under Conditions of Varying Alumina Concentrations. *Proc. Aust. I.M. & M.*, No. 214, pp. 1–19.
- Grjotheim K, Kvande H, Motzfeldt K, Welch BJ (1972) The Formation and Composition of the Fluoride Emissions from Aluminium Cells”, *Canadian Metallurgical Quarterly*, 11, pp. 585–599.
- Martin O, Garipey R, Girault G (2017) APXe and AP60: The new References for Low Energy & High Productivity Cells. Proc. 11<sup>th</sup> Australasian Aluminium Smelting Conference.
- Purdie JM, Bilek M, Taylor MP, Zhang WD, Welch BJ, Chen JJJ (1993) Impact of Anode Gas Evolution on Electrolyte Flow and Mixing in Aluminium Electrowinning Cells. *AIME Light Metals 1993*, pp. 355–360.
- Jassim A, Akhmetov S, Welch BJ, Skyllas-Kazacos M, Jie B, Yuchen Y (2015) Studies on Background PFC Emission in Hall-Héroult Reduction Cells Using Online Anode Current Signals. In Hyland, M (ed) *TMS Light Metals 2015*, pp. 545–550.
- Fortin S, Gerhardt M, Gesing AJ (1985) Physical modelling of bubble behaviour & gas release from aluminium reduction cells. In McGeer, P (ed) *Light Metals 1985*, The Minerals, Metals & Materials Society, pp. 721–741.
- Chen JJJ, Chen ZD, Walker ML, Wei CC, Welch BJ (1995) A Review of Model Studies of the Hall-Héroult Cell for Aluminium Production. Proc. IPENZ Annual Conference 1995, Vol. 2, 1995, pp. 385–391.
- Aaberge RJ, Ranum V, Williamson K, Welch BJ (1997) The gas under anodes in smelting cells. Part II: Gas volume and bubble layer characteristics, *Light Metals 1997*, pp. 341–346.
- Saevarsdottir G, Kvande H, Welch BJ (2020) Reducing the Carbon Footprint: Aluminium Smelting with Changing Energy Systems and the Risk of Carbon Leakage. In Tomsett, A (ed) *Light Metals 2020*, The Minerals, Metals & Materials Society, pp. 726–734.
- Whitfield D, Akhmetov S, Al-Jallaf M, Blasques J, Al-Aswad K, Baggash I (2015) From D18 to D18+: Progression of Dubal’s Original potlines. In Hyland, M (ed) *TMS Light Metals 2015*, pp. 499–504.
- Moxnes B, Solheim A, Liane M, Svinsås E, Halkjelsvik A (2009) Improved cell operation by redistribution of the alumina feeding. In Bearne, G (ed) *TMS Light Metals 2009*, pp. 461–466.
- Shi J, Wong CJ, Yao YC, Bao J, Skyllas-Kazacos M, Welch BJ (2018) Advanced Feeding Control of Aluminium Reduction Process. Proc. 12th AASTC, paper 4c2.
- Yao YC, Cheung CY, Bao J, Skyllas-Kazacos M, Welch BJ, Akhmetov S (2017) Estimation of Spatial Alumina Concentration in an Aluminium Reduction Cell Using a Multilevel State Observer, *AIChE Journal*, 63 (7) pp. 2806–2818.