CHAPTER 7.3

Pneumatic Flotation

Greg Harbort

Pneumatic flotation cells are typically classified as high-intensity flotation machines with rapid kinetics, resulting in short residence times, often half that of mechanical flotation cells and a quarter of that of flotation columns. Wash water is often used to enhance concentrate grade. Typically used in cleaning duties and coal flotation, pneumatic flotation cells are also used in roughing and scavenging in a wide variety of mineral commodities and wastewater treatment. Numerous publications have provided regular updates on their development, including those by Finch (1995), Rubinstein (1995), Aksani (1998), Lynch et al. (2007, 2010), and Cheng and Liu (2015).

Pneumatic flotation machines were among the first machines used in flotation. However, with the advent of mechanical subaeration cells, the use of pneumatic flotation machines significantly declined. In 1928, five pneumatic cell designs were in industry use. Although this number increased to 11 machine types by 1945, the majority were derivatives of the earlier Callow, Forrester, and MacIntosh pneumatic cells.

The modern pneumatic flotation cell can most effectively be classified according to the theoretical and technical considerations of Bahr (1971). Although a variety of aerators, pulp feed arrangements, and separating vessel designs exist (Harbort and Clarke 2017), the applied principles and fundamental design remain unchanged. Because of the absence of an impeller, high fluid velocities are required to disperse the air into fine bubbles and to maintain particle suspension. Researchers realized that bubble–particle attachment and the separation of the phases could be considerably improved when carried out in separate units. Aerators are typically situated outside the separating vessel. The bubbles are produced by several means including sized pores, annular gaps, swirling annuli, and plunging jets. Air can either be supplied under high pressure or induced under vacuum.

The characteristics of pneumatic flotation can be summarized as follows:

- No moving parts exist for production of fine air bubbles.
- Particle adhesion to air bubbles occurs in the aerator (forced bubble adhesion).
- The aerator and separating vessel are physically separated.

Flotation columns have many features in common with pneumatic flotation, including the lack of moving parts and the use of wash water. The major differences compared with pneumatic flotation are

- Air-bubble/particle adhesion takes place within the separating column.
- There is a countercurrent flow of air bubbles and solid particles.
- Bubble–particle contact is of significantly lower intensity.

In general, the height/diameter ratio of column flotation cells is larger than that for pneumatic flotation. For column flotation, ratios of 2.2:10 are reported, while for pneumatic flotation plants, ratios of 1.5:3 are common.

DAVCRA CELL

The Davcra cell was developed in the 1960s in the Metallurgical Research Section of the Zinc Corporation. William Davis, who directed the cell's design and development, had proposed that the major proportion of mineral particle–bubble interaction occurred within the high-intensity zone of a flotation machine and was virtually time independent. Therefore, the body of the flotation cell was responsible for only a minor portion of recovery, and flotation residence time in the normally accepted sense had little meaning (Davis 1964, 1966).

The Davcra cell's operation is illustrated in Figure 1. Pulp under pressure enters the nozzle tangentially and issues from the nozzle orifice as a swirling annulus. Air under pressure passes through the orifice as a central air core. On entering the body of pulp in the tank, the interaction of the pulp annulus and air jet causes a shearing of fine bubbles with intimate mixing of these bubbles through the pulp annulus. The bubble–particle aggregates formed float into a mineralized froth. The feed pulp is prevented from flowing directly out of the tailing discharge by a baffle. This ensures that all tailings must first pass through the quiet deaeration zone where remaining bubble–particle aggregates rise into the froth. The level of pulp in the cell is maintained by hydraulic or mechanical restriction of flow in the tailing line (Cusack 1968).

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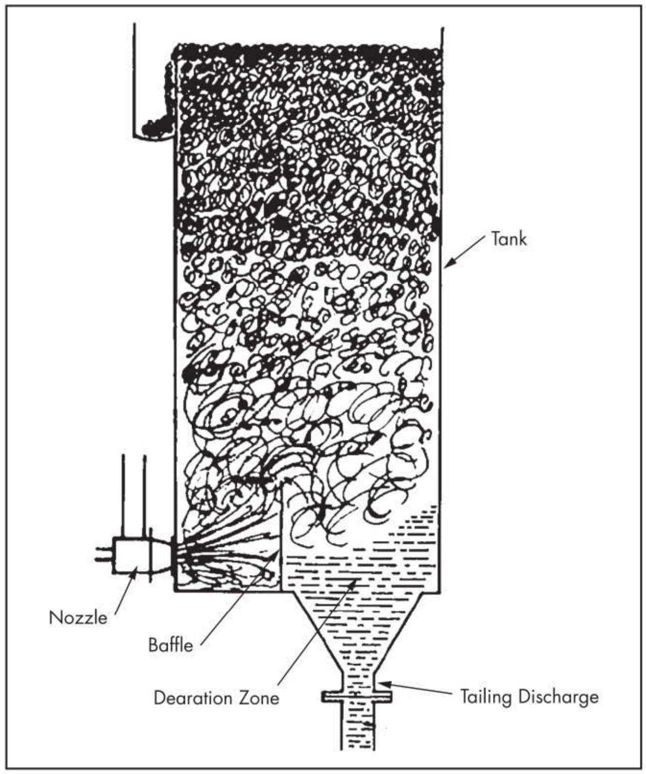
Extensive pilot tests were conducted with Davcra cell prototypes at Broken Hill, New South Wales, Australia, progressively scaling up from 8 t/h (metric tons per hour) unit capacity to 60 t/h, and then to 200 t/h (Figure 2). By 1968, production units were in operation at Broken Hill in both lead roughing and scavenging duties (Cusack and Oley 1971). The largest cells had a nominal unit capacity of 363 t/h, in a tank 5.4 m long, 1.7 m wide, and 3.7 m deep. The aeration nozzle was the major wear component, with the use of titanium diboride ceramic providing an operating life >4,000 hours. The nozzle typically operated at a pulp pressure of 186 kPa, with compressed air supplied at 193 kPa.

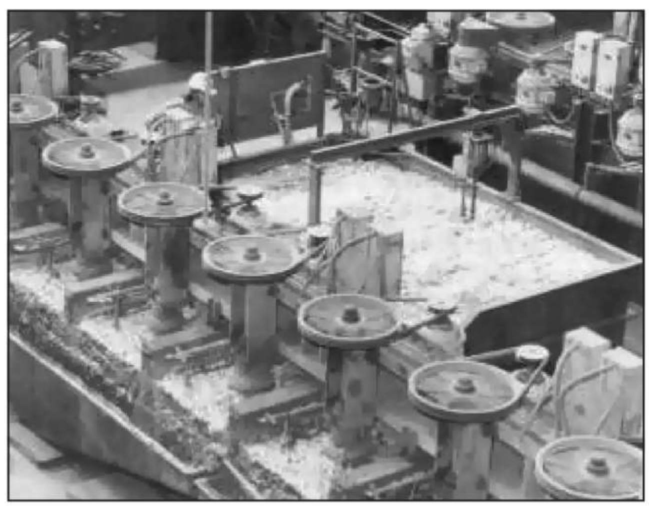
The Davcra cell was conclusively proven to achieve higher recoveries at around half the residence time of mechanical flotation cells. By 1975, the cell was the most dominant of the pneumatic flotation cells installed in industry. Sites that operated successfully with it included Leinster Nickel in Australia; Bougainville Copper in Papua New Guinea; Anglo American Platinum in South Africa; ZCCM Mufulira copper concentrator in Zambia; Palabora Mining Company copper concentrator at Phalaborwa, South Africa; and Territory Enterprises copper mine and Coal Cliffs mine in Australia.

The Davcra cell was a robust, reliable flotation machine. However, it was developed by a company whose primary business was mining, and after a review of noncore business, the Zinc Corporation stopped manufacturing the Davcra cell.



Source: Cusack 1968, reprinted with permission from the Australasian Institute of Mining and Metallurgy

Figure 1 Sectioned view of the Davcra flotation cell



Courtesy of Zinc Corporation

Figure 2 Davcra flotation cell (in background) at Broken Hill

BAHR CELL

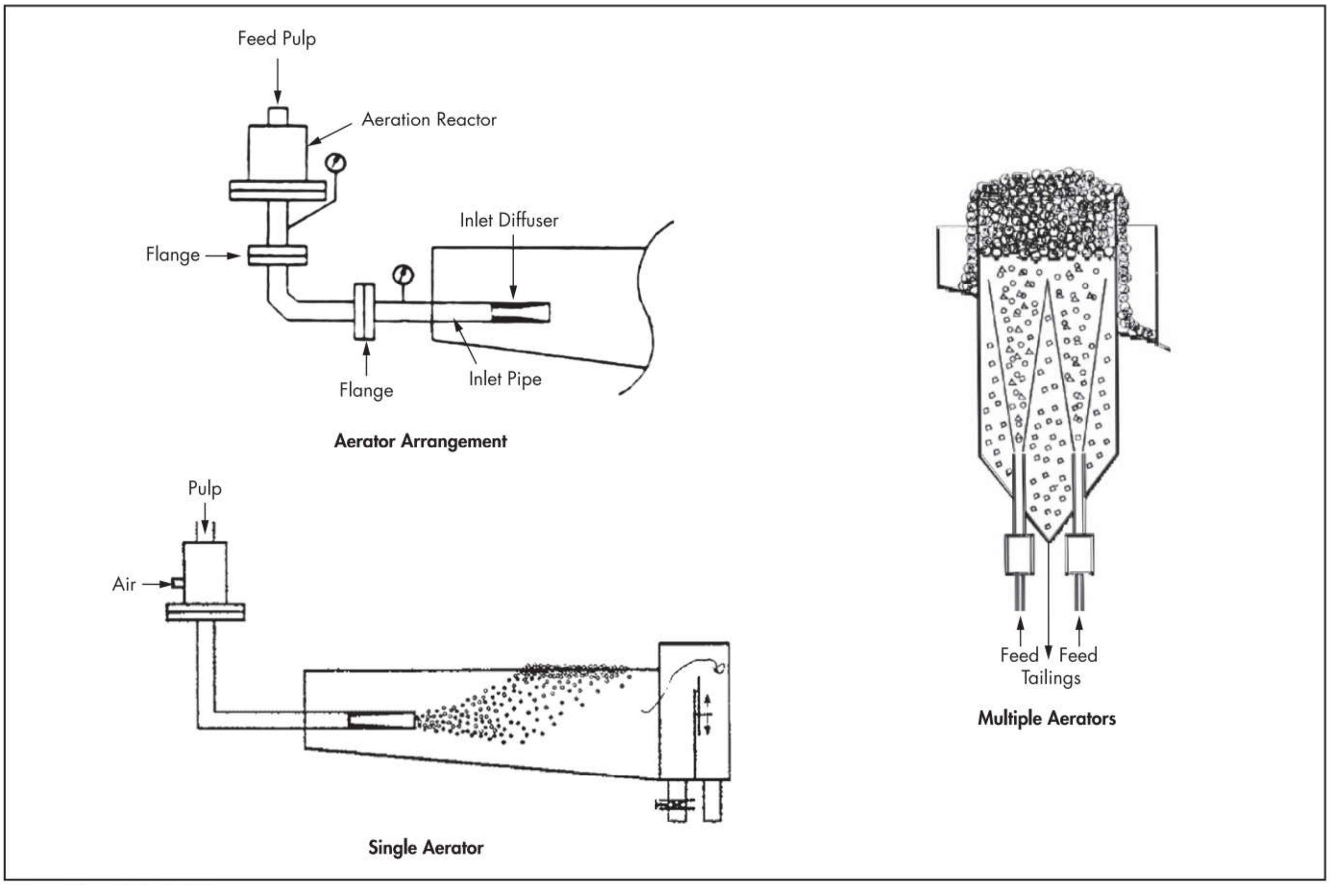
Modern flotation machine developments in Germany commenced in the 1950s and centered on pneumatic flotation cells. In the ensuing decades, significant work was conducted by Bergbau-Forschung GmbH, the Clausthal University of Technology, the Technical University of Berlin, and KHD Humboldt Wedag. Individuals who had a preeminent role in this development included Albert Bahr, Wolfgang Simonis, and Rainer Imhof, among others.

Early work by KHD focused on "cyclone flotation," where compressed air was forced through porous walls of the flotation tank to generate bubbles. The pulp was fed tangentially into the cell for contacting with the air. After extensive laboratory and pilot tests, an apparently workable design was developed by 1961 (Salzmann and Koch 1964). A production unit was installed in 1965 at the Lüderich lead–zinc flotation plant. Its success was not overwhelming, given problems with blockages in the porous media, and this line of development was discontinued (Cordes 1997).

Ongoing investigation throughout the 1970s would eventually result in development of the Bahr cell in the early 1980s. This pneumatic flotation cell was developed jointly by Bergbau-Forschung GmbH, Clausthal University of Technology's Institute for Mineral Beneficiation, and Ruhrkohle AG's Westerholt mining company.

The design utilized external aerator units where compressed air flowed through annular rings via channels into the pulp. The aerator units were located beneath the main flotation tank and entered the tank vertically (Figure 3). Bahr confirmed that the optimum air-to-pulp ratio was 1.3:1.5 during coal flotation with the pneumatic cell. The optimum air-to-pulp ratio for the flotation of sulfide ore was much lower (0.15:0.48). Both high airflow rates and low pulp flow rates resulted in larger bubbles and lower recoveries (Changgen and Bahr 1992).

Production units were successfully installed at several plants, including the Dorfner kaolin plant in Germany, which secured the exclusive use of the Bahr cell in kaolin flotation, as well as at several coal operations in Germany. The success of the Bahr cell in Germany resulted in a large pilot-scale testing program in South Africa on coals from the Waterberg, Soutpansberg, and some of the Natal coalfields. This resulted in the successful installation of Bahr cells in the Iscor Durnacol coal plant (Gorzitzke and MacPhail 1989) and Grootegeluk coal mine (Ventert and Van Loggerenbergt 1992).



Adapted from Bahr 1971
Figure 3 Bahr cell

EKOF CELL

In the 1980s, KHD Humboldt Wedag commenced investigation into free jet flotation machines. The concept was developed from research conducted by Sigfried Heintges, Ali Alizadeh, and Wolfgang Simonis at the Technical University of Berlin (Heintges et al. 1984; Alizadeh and Simonis 1985). As shown in Figure 4, the conditioned pulp was pumped through a nozzle (4). Upon exiting the nozzle, air was entrained by the free jet (5) and broken into fine bubbles when it impinged upon the pulp. A downcomer (3) carried the aerated mixture into the main flotation tank (1), with the immersion depth of the downcomer used to vary bubble size distribution and the air entrainment rate adjusted by varying the nozzle feed pressure. Although initial tests on coal produced acceptable results, it would be another 10 years before the self-aspirating, free jet flotation machine was introduced as a commercial product.

From 1985 to 1987, KHD made a substantial effort to develop aeration reactors and flotation cell design. Extended pilot-scale tests were conducted at the Hugo and General Blumenthal coal plants of Ruhrkohle AG in Germany. The eventual pneumatic cell design would combine elements of both the earlier unsuccessful "cyclone flotation" cell and the Bahr cell. Marketed initially by KHD's subsidiary Erz- und Kohleflotation GmbH (Ekof), it was known as the Ekoflot, or Ekof flotation cell.

The Ekof cell used ring-shaped ceramic slot aerators housed within a polypropylene body. Compressed air passed through the discs with a gap of 50–75 μm before entering the pulp flow in a high-velocity region created by a displacement body. The highly aerated mixture was piped down to the flotation tank where its entry was tangential to the tank wall, creating a slow swirling motion in the pulp. The concentrate was removed through a central launder, with the circular motion enhancing transport and removal (Figure 5).

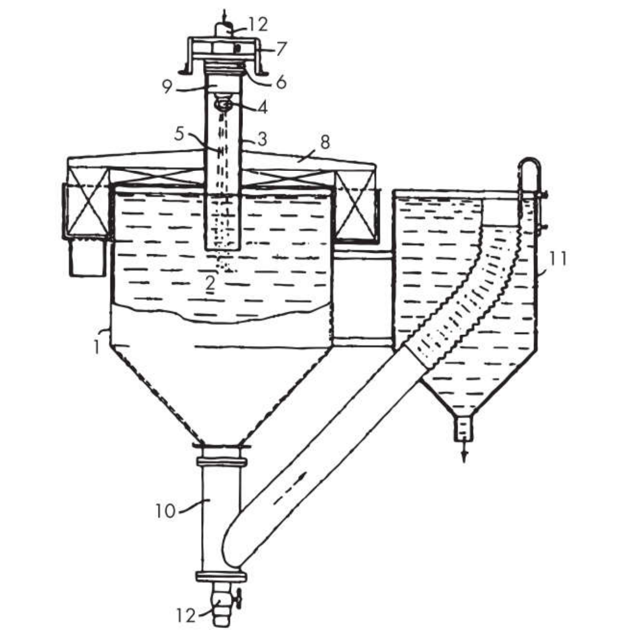
In May 1987, two 4.5-m-diameter Ekof cells with the tangential feed design were commissioned at the Pittston/Clinchfield coal operations in McClure, Virginia, United States. The units were operated in parallel, treating 1,600 m^3 of coal per hour. This installation was rapidly followed by one at the Fechner coal mine in Germany. The plant was commissioned in 1988 in a coal tailing retreatment duty using a 5.1-m-diameter cell, treating 1,600 m^3/h. At the time it was one of the highest capacity unit flotation cells installed.

PNEUFLOT CELL

In the mid-1990s, strategic restructuring within KHD saw the Ekof company return to a flotation reagent focus, with flotation machinery development directly handled by KHD. The Ekof cell became known as the Pneuflot cell. In addition to the tangentially fed tank, there was increased interest in a cell design using the aeration unit feeding into a vertical downcomer that discharged into the tank. The aerator is self-inducing and engineered to increase the velocity of the slurry as it runs over air inlets. The resulting Venturi effect causes air to be introduced as the slurry is pumped through the system. The aerator is also designed so that when the air enters the slurry, it sheers off in small bubbles and sufficient energy is generated to achieve particle–bubble attachment. This energy is generated in the aerator by the decreasing diameter into the aerator.

The single aerator feeds into a downcomer that passes through a central froth crowder and discharges into a circular perforated pipe that distributes the highly aerated pulp evenly across the cell cross section, as shown in Figure 6.

Initially, both designs were available from KHD (Sanders and Williamson 1996). At the Michilla copper operation in Chile, vertically fed, self-aerating Pneuflot cells were installed in one duty, with tangentially fed cells using compressed air in another duty (Sánchez-Pino et al. 2008). Within several years, the vertically fed cell would become the dominant Pneuflot design. Significant installations included the Magnesita magnesite operation, where six tangential Pneuflot cells were installed, and the Michilla copper operation, both in Chile (Ekof, n.d.; KHD 1997, 2006). A small but not insignificant number of Pneuflot cells were installed between 1992 and 1996; however, for eight years after that, the Pneuflot cell fell out of favor in the flotation industry. One reported difficulty was a shortage of pilot-plant equipment and technical support outside Europe. It was, however, to make a significant return in 2005–2006, with record numbers of cells installed (Markworth et al. 2007), including the Liuqiao and Zhujiadian coal mines in China and the Dartbrook coal preparation plant of Anglo Coal Australia. In 2009, McNally Bharat Engineering, a part of the Williamson Magor Group, took over the coal and mineral division of KHD. Under the name MBE-CMT, it continued to market the Pneuflot cell. Installations under the new owners have included phosphate roughing and cleaning in Kazakhstan and iron ore flotation in Brazil and Chile. In 2012, twelve Pneuflot cells ranging in diameter from 2.5 to 6.0 m were installed in copper roughing and scavenging duties in Kazakhstan.



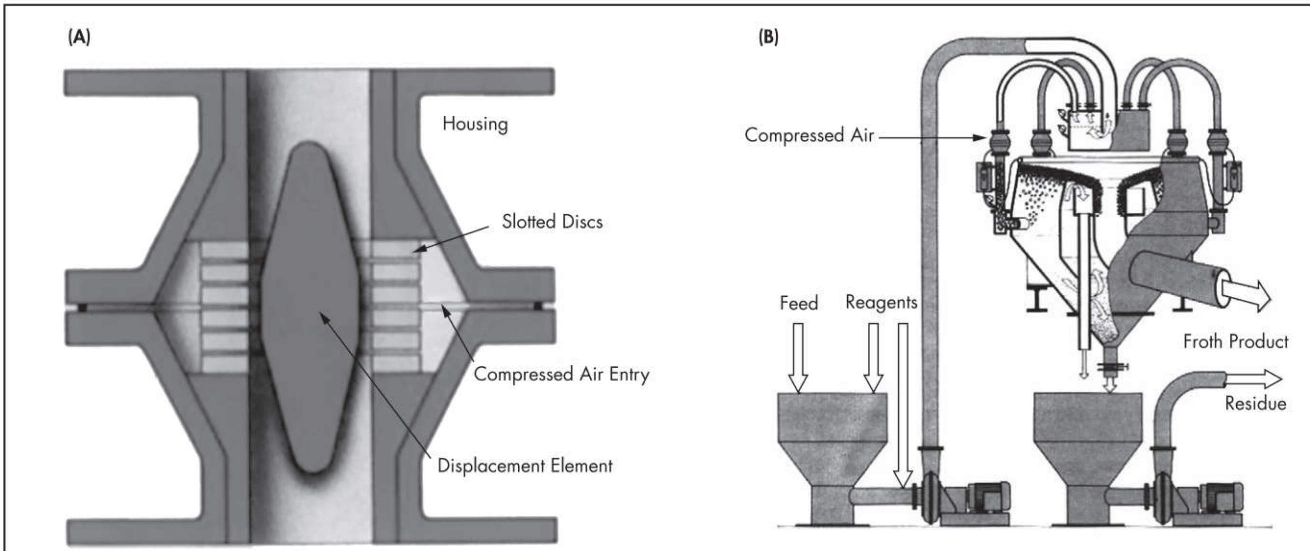
Note: Only relevant numbered items in the figure are discussed in the text.

Source: Heintges et al. 1984

Figure 4 Early-concept free jet pneumatic flotation cell

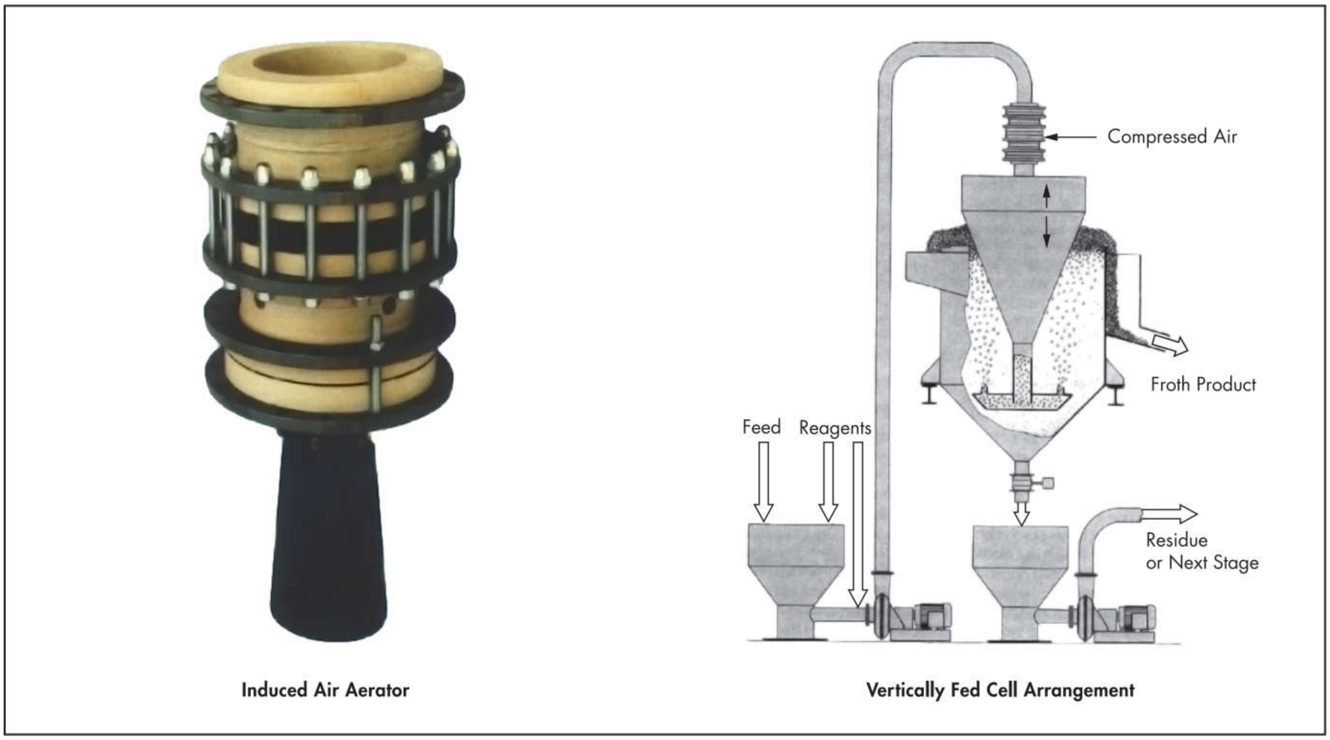
ALLFLOT CELL

Another German company that was to achieve success with its pneumatic flotation machines was Allmineral Aufbereitungstechnik of Duisburg with its Allflot cell. The key to the device was the aeration unit developed in the mid-1980s by Andreas Jungmann and Ulrich Reilard (Figure 7). In the aerator, compressed air passed through a series of annular

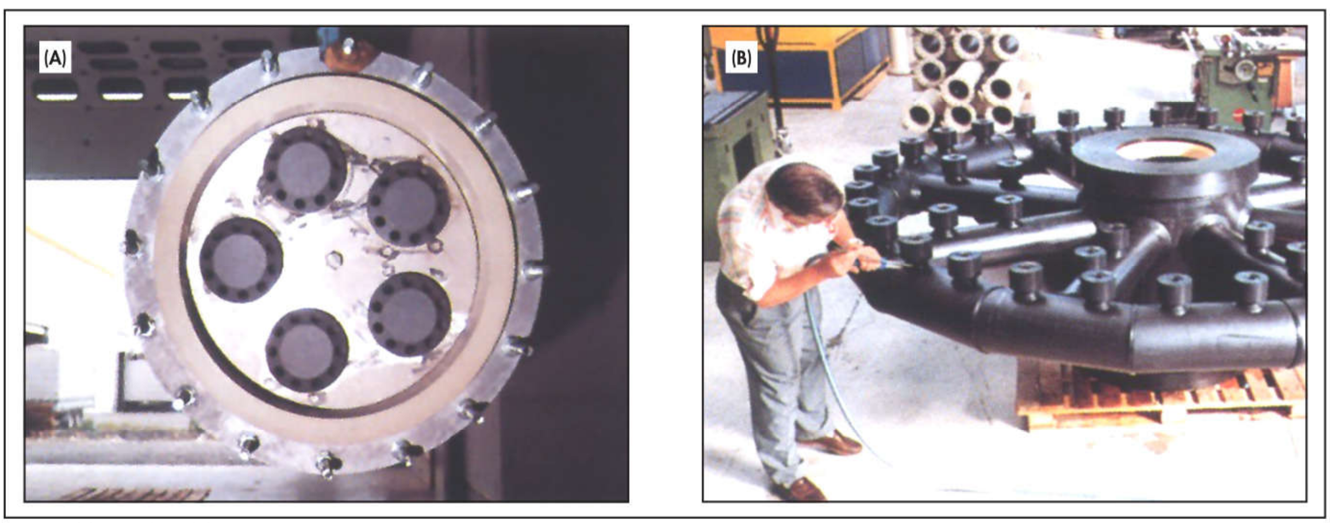


Courtesy of Ekof

Figure 5 (A) Slot aerator and (B) Ekof cell arrangement



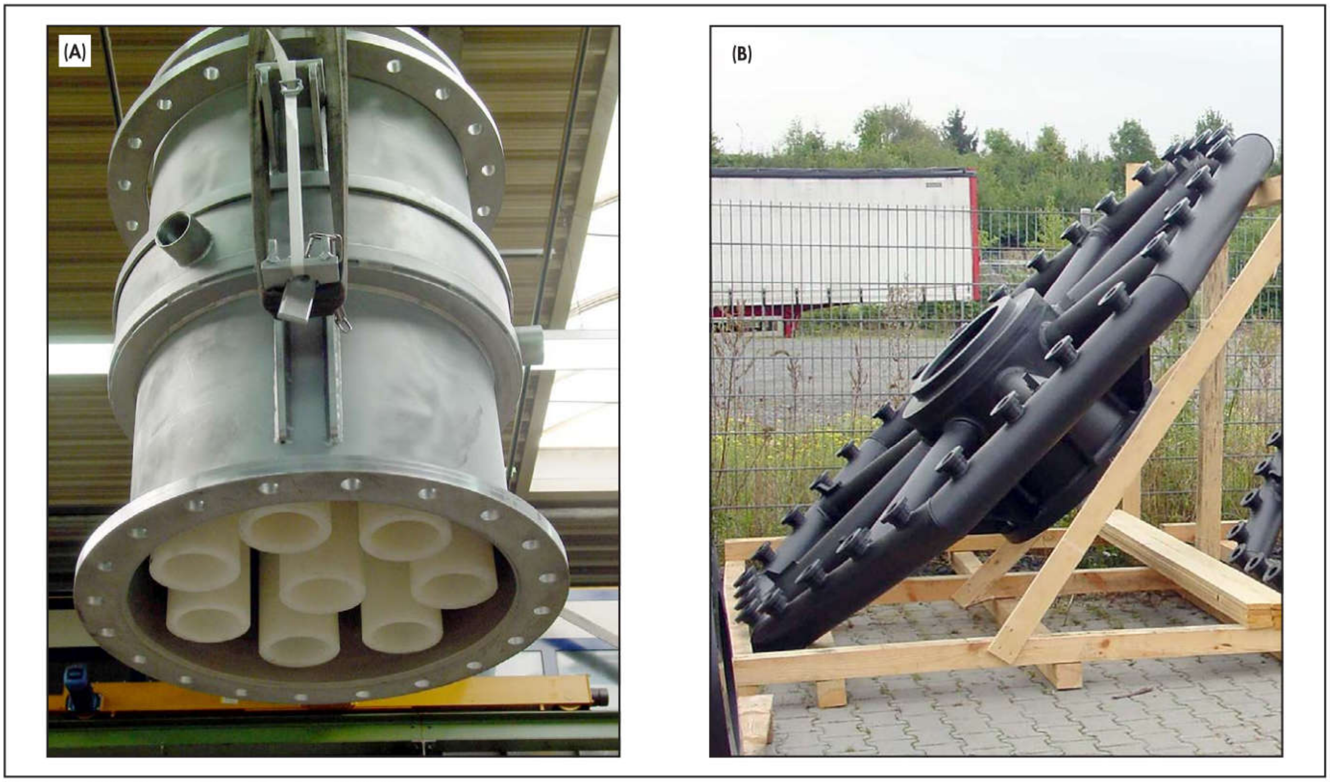
Courtesy of MBE Coal and Minerals Technology GmbH
Figure 6 Pneuflot cell



Source: Allmineral, n.d.(b)
Figure 7 (A) Feed side of 800 m^3/h Allflot aerator and (B) aerated pulp distribution manifold

slots surrounding the slurry stream. A high-velocity region created by a displacement body within the aerator produced intense contact between fine bubbles and pulp. A swirling motion was imparted to the aerated mixture by a series of screw-shaped strips within the mixing and dispersion sections. The highly aerated mixture entered a downcomer and passed to a dispersion ring to allow even dispersion of the mixture within the flotation tank (Jungmann and Reilard 1989; Allmineral, n.d.(a), n.d.(b)).

One major area of success for the Allflot cell was in coal flotation in Poland. Prior to 1990, a centrally managed economy operated there. Coal production costs were not an important issue, and losses on coal production were covered by government subsidies. After 1990, the transformation of Poland's centrally planned economy to an open-market economy saw significant changes in the country's coal industry. Nonprofitable mines were closed, profitable mines were modernized, and new coal preparation plants were constructed



Courtesy of Maelgwyn Mineral Services Limited
Figure 8 (A) Imhoflot V-cell nozzle arrangement and (B) ring distributor

(Blaschke and Gawlik 2004). There was also increased cooperation between German and Polish companies. Allmineral was one company that became a major supplier of equipment to Poland. Allflot cells were installed in the Jankowice, Bielszowice, and Knurów coal preparation plants treating –0.5-mm material (Blaschke 2000; Nycz and Zieleźny 2004). Tighter controls also saw increased quality requirements for coal-fired power plant feed stocks with the use of flotation in steaming coal preparation plants being introduced. This resulted in the installation of Allflot cells with a capacity of 600 m^3/h into the Jaworzno plant in 2002 and the Janina plant in 2004. Also of note was a copper slag flotation plant installed in Chile treating copper slag.

IMHOFLOT CELL

Also tracing their roots back to Clausthal University and the University of Berlin are Imhoflot pneumatic flotation cells. Rainer Imhof had been involved to some degree with the Germany pneumatic flotation machines for several decades. He continued their design and development, resulting in the Imhoflot pneumatic flotation process that has been successfully marketed by Maelgwyn Mineral Services of the United Kingdom.

The initial Imhoflot cell design was the V-cell. Two of its key components are shown in Figure 8. Pulp was pumped to a series of nozzles and the resultant energy aspirated air, which was sheared into fine bubbles. The aeration unit, or bubble generator, was constructed of fused silicon carbide components to protect it from wear (Imhof et al. 2005). The aerated mixture traveled through a downcomer tube and was introduced upwardly into the cell by a ring distributor system and nozzles.

An early installation of the Imhoflot V-cell was the Compañía Minera Huasco iron ore concentrator in Huasco, Chile. The twin-stream flotation plant incorporated six 4.5-m-diameter Imhoflot cells to treat 600 t/h of magnetic separator concentrates for removal of silicates. Extensive testing was conducted over several years with construction of the new flotation plant occurring in 2002 (MMS 2002a, 2003). Another significant installation took place in 2002 with another 4.5-m-diameter V-cell introduced to a coal tailings process plant in the Lugansk region of Ukraine. An Imhoflot plant was also used successfully for the El Romeral iron ore operation owned by Compañía Minera del Pacifico (CMP) of La Serena, Chile. The flotation plant incorporates two 5.2-m-diameter Imhoflot V-cells and has a design capacity of 965 m^3/h. The process is based on reverse flotation to remove silica impurities from the iron ore concentrate and is similar to that at CMP's Huasco pelletizing operation (MMS 2006a). A V-cell flotation plant was also installed at the Belaruskali potash plant in Turkmenistan in 2006. Operating in a saturated brine solution, the two 3.5-m flotation cells have a design throughput of 350 m^3/h treating 50–60 t/h solids (MMS 2006b). Figure 9 shows details of the V-cell arrangement and installation.

The second Imhoflot cell design was the G-cell, which incorporated centrifugal forces within the flotation tank. The aerator was self-aspirating and uses a high-shear ceramic multi-jet venturi system operating at around 250 kPa back pressure. Aerated feed is fed tangentially into the flotation tank. Unlike the Pneuflot tangentially fed cell, where rotational forces are not significant, the G-cell produces specific rotational speeds in the pulp, achieving centrifugal acceleration of 9.8 m/s^2. The resultant force on the pulp creates an angled pulp–froth interface. This enhances froth flow to the concentrate launder (Imhof et al. 2007). Figure 10 shows an assembled G-cell and the centrifugal flow of pulp within the tank.

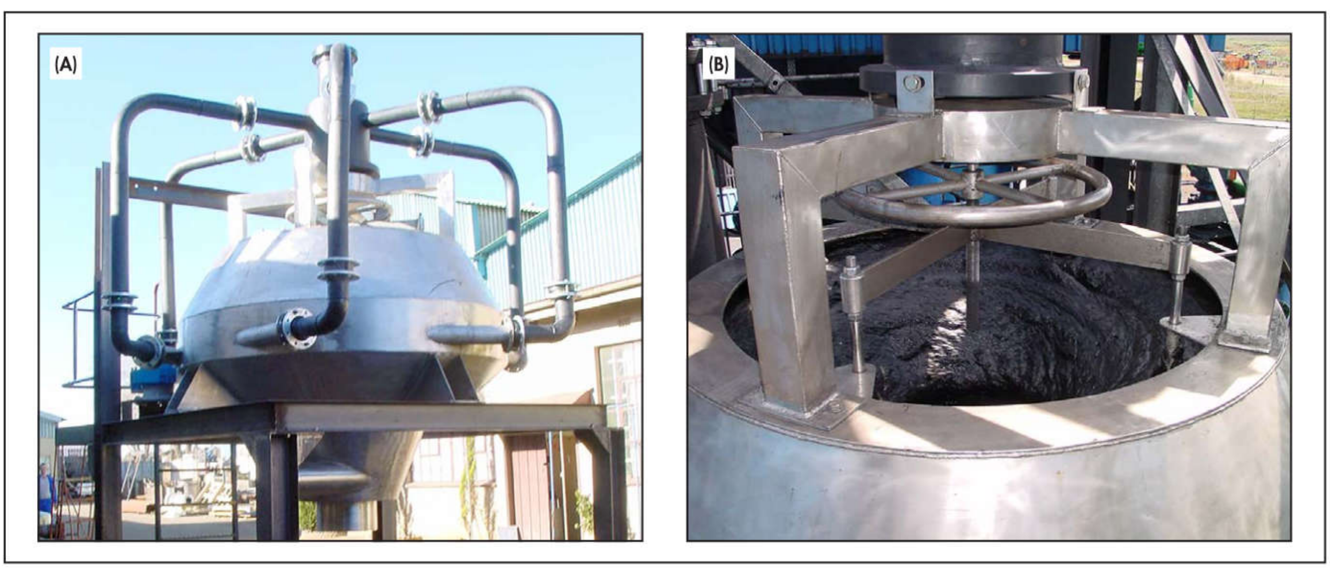
In 2002, Ingwe Coal Corporation installed a 2.2-m-diameter G-cell with a feed capacity of 300 m^3/h in its Koornfontein operations in South Africa (MMS 2002b). During 2004, two 1.2-m G-cells were installed at the Barbrook mine of Caledonia Mining Corporation in South Africa to upgrade rougher flotation concentrates (MMS 2004). By 2005, another three 1.8-m-diameter Imhoflot cells had been installed in the Dorfner kaolin plant in Germany. The 2007 Imhoflot G-cell installation at ZapSib in Siberia, Russia, represented a quantum leap for the G-cell in terms of both scale-up and installation size, utilizing six 3.6-m-diameter cells to recover coal.

Courtesy of Maelgwyn Mineral Services Limited
Figure 9 Installation in Chile

JAMESON CELL

The Jameson cell is currently the most widely used pneumatic flotation cell, both in terms of installed capacity and number. The principles of Jameson cell operation have been discussed by numerous authors including Jameson (1988), Jameson et al. (1988), Atkinson (1994), Evans et al. (1995), and Harbort et al. (2002, 2003). The principles can be described with reference to Figures 11 and 12.

In its design, slurry is pumped to the Jameson cell slurry distributor and split between several downcomers. The downcomer is where primary contact of bubbles and particles occurs. The liquid feed is delivered to the nozzle under pressure, which results in a free jet of liquid being created as it passes through an orifice plate. As the free jet travels through the air-filled section of the downcomer, contact with the air results in a slight slowing of the jet, minor expansion of the jet diameter, and undulations on the jet surface, which in turn entrain small amounts of air. The free jet impinges on the liquid mixture within the downcomer, and the pressure of the impact creates a depression on the liquid surface, called the induction trumpet. Because of the fluted entry shape of the induction trumpet, air is channeled into the area at the base of the free jet. The free jet passes through the induction trumpet carrying with it this layer of entrained air. Additionally, the periodic collapse of the induction trumpet results in further air entrainment. Once the jet enters the main body of liquid in the downcomer through the bottom of the induction trumpet, it is referred to as a plunging jet. The high-shear rate of the plunging jet results in the entrained layer of air being broken down into a multitude of small bubbles, typically of 500-µm diameter, which are carried down the downcomer length by the inertia of the injected slurry and gravitational forces. The kinetic energy of the plunging jet is dispersed through the creation of a region of intense turbulence where momentum is transferred to the surrounding mixture and expands to occupy the downcomer cross section, creating recirculating eddies of aerated liquid. This region of jet dissipation is known as the *mixing zone* and is defined as the downcomer volume occupied by (1) the fluid inside the submerged jet below the induction trumpet



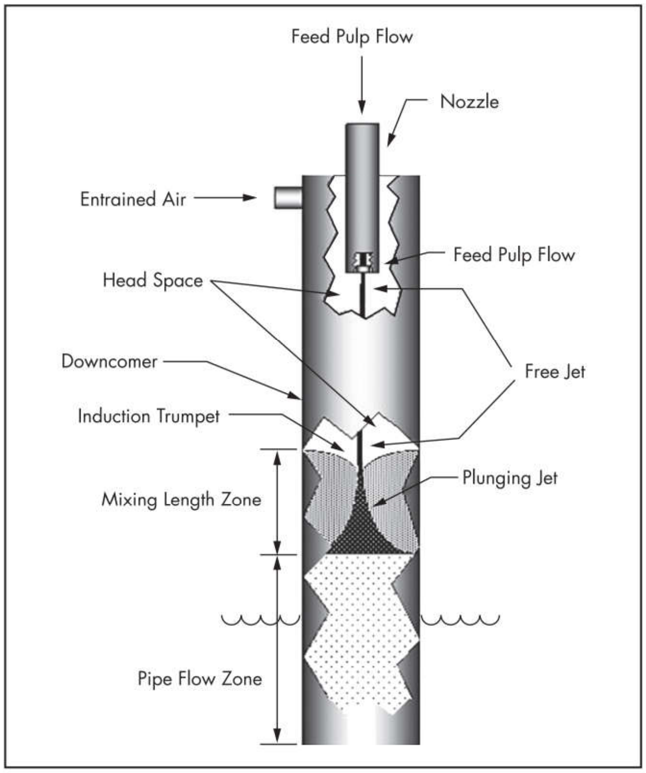
Courtesy of Maelgwyn Mineral Services Limited
Figure 10 (A) Assembled G-cell and (B) centrifugal flow of pulp

and (2) the body of recirculating fluid between the submerged jet and the downcomer wall.

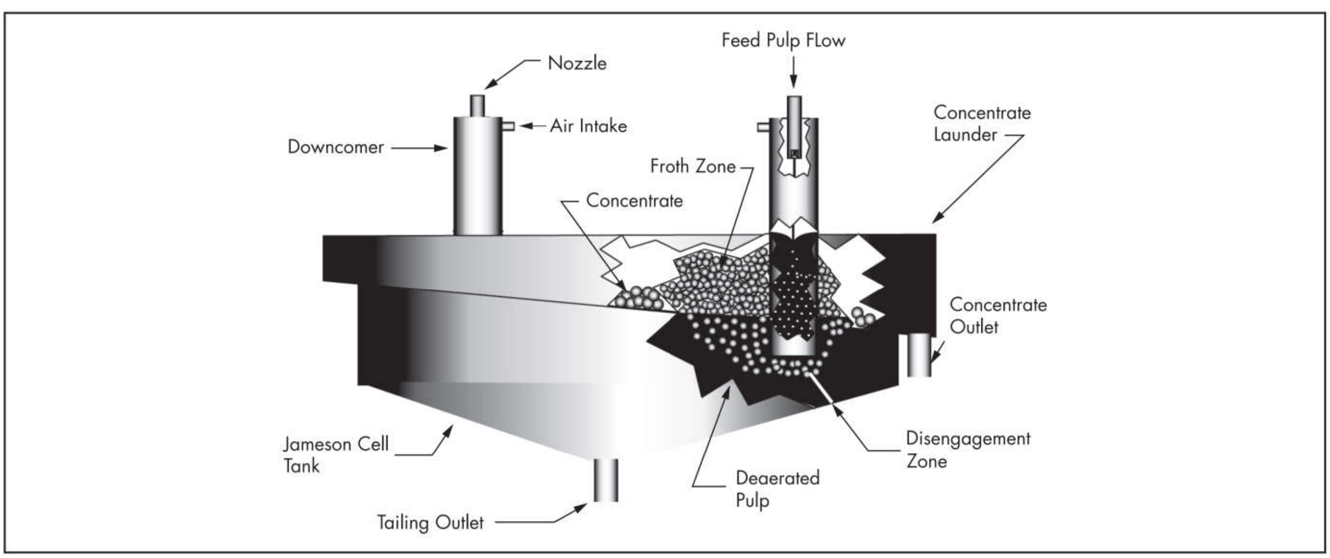
Within the mixing zone, ongoing bubble coalescence and bubble creation continue to occur. Beneath the mixing zone is a region of uniform multiphase flow known as the pipe flow zone. Because the net downward slurry velocity counteracts the upward buoyancy of the bubbles, the bubbles pack together to form a moving expanded bubble bed of high void fraction. The pipe flow zone is characterized by a bubbly flow at lower air rates and moderately churn-turbulent flow at higher air rates.

The bubbly mixture exits the downcomer into the tank pulp zone where secondary contacting of bubbles and particles occurs and bubbles disengage from the pulp. The velocity of the mixture and large density differential between it and the remainder of pulp in the tank results in recirculating fluid patterns, keeping particles in suspension without the need for mechanical agitation. Tank void fraction measurements and computational fluid dynamics modeling (Koh and Schwarz 2009) show that bubble patterns in general form a central, air-swept cone surrounding each downcomer (Figure 12) as described by Taggart (1945). The Jameson cell tank contains areas of high, localized air void throughout the pulp zone (Figure 13). The rising swarm of bubbles is governed by several factors including recirculating patterns within the tank, pulp flow volumes, and airflow volumes. The tank froth zone is where materials entrained in the froth are removed by froth drainage and/or froth washing.

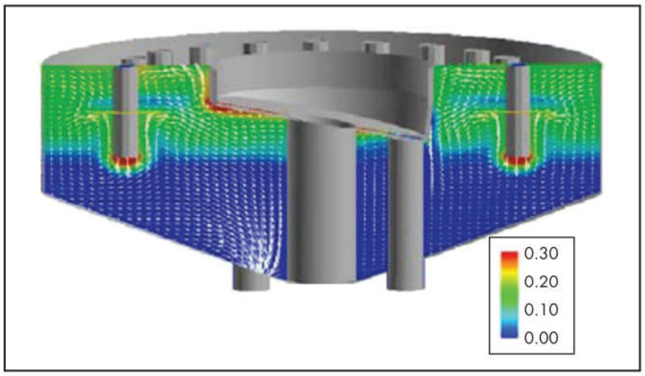
Between June and December 1986, research and development staff at Mount Isa Mines in Queensland, Australia, evaluated column flotation on a range of streams in the lead–zinc concentrator and the copper concentrator for their upgrading properties in a pilot-plant-scale column operated correctly with a wash-water addition. As a result of the work, the company decided to install flotation columns. Before the column installations were completed, concerns were raised on possible future maintenance issues with the simple sparging systems that had to be used. Professor Graeme Jameson visited Mount Isa to describe in a presentation the options for bubble preparation in columns, including a newly developed laboratory-scale device developed at the University of Newcastle in Australia. The device was sent to Mount Isa for evaluation by research and development staff, based on the methodology created with the pilot-plant column. The device did not include a wash-water addition system. As a result of the benefits observed from the proper addition of wash water in the preceding tests of the pilot-plant column, this technology was



Source: Harbort 2006
Figure 11 Jameson cell downcomer



Source: Harbort 2006
Figure 12 Jameson cell



Source: Koh and Schwarz 2009
Figure 13 Air void fraction with Jameson cell

added to the device through the initiative of the research and development staff.

Research and development staff at Mount Isa Mines demonstrated the following:

1. The use of wash water in the device allowed the device to improve its performance (based on the position of its grade–recovery curve); further, this improved performance was equal to the pilot-plant column's performance for a given feed, again based on the position of the grade–recovery curve.
2. Successful scale-up of the device was demonstrated; the performance of a series of progressively enlarged versions of the device was evaluated during a period of collaboration between Mount Isa Mines and the University of Newcastle, where the enlarged versions were made at the University of Newcastle and sent to Mount Isa for evaluation.

In specific applications, equivalent performance could be achieved in the short tank design, now known as the Jameson cell (Jameson et al. 1988).

In 1989, two 1.9-m-diameter Jameson cells were installed in the Mount Isa lead–zinc concentrator cleaning lead slimes (Jameson and Manlapig 1991). This installation showed a vast difference in the flotation kinetics of the Jameson cell when compared to both mechanical cells and flotation columns. The enhanced Jameson cell kinetics were attributed to the short downcomer residence time allowing collection to occur prior to galena oxidation.

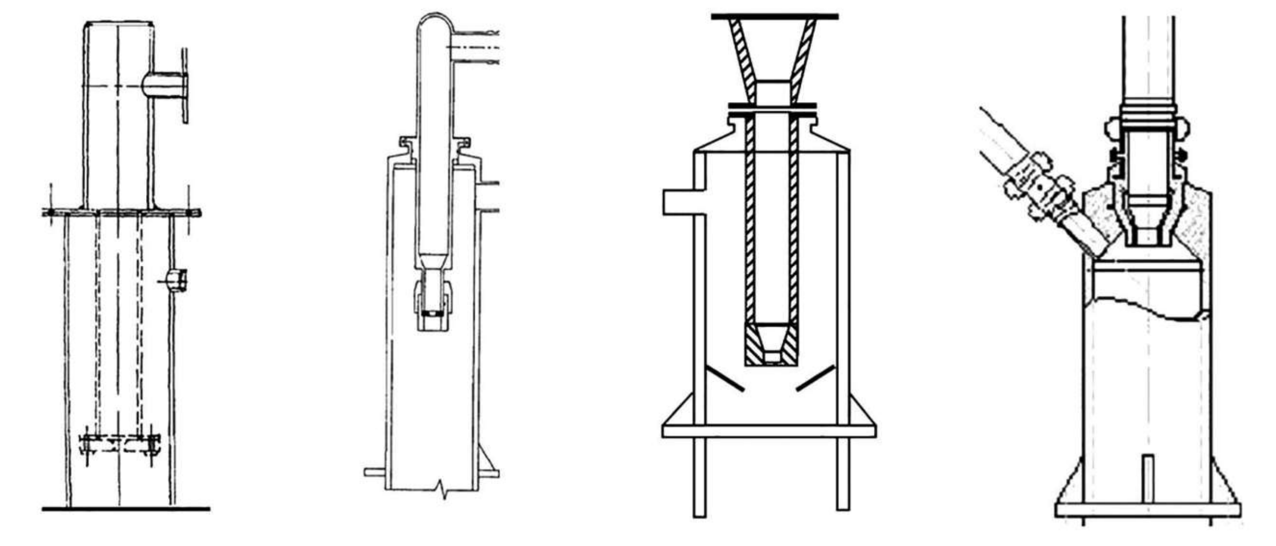
The initial success, however, was not replicated at other sites where the cell was tested. Conventional column flotation and the Jameson cell were both evaluated in the cleaning of Zn, Pb, Cu/Ag, and bulk rougher concentrates at the Luina concentrator in Australia in 1988. Both the column cell and the Jameson cell were suitable in cleaning Cu/Ag rougher concentrate. The very slow flotation rates and low rate differentials between pyritic gangue and sulfide values made the large volume, multistage conventional cleaning circuit more successful in other cleaning applications. Based on these results, a hybrid of Jameson technology and a conventional column was installed at the Hellyer mine in Australia for Cu/Ag cleaning (Lane et al. 1991). The hybrid system was chosen to maximize recovery in the recovery zone of the column. The feed was to be introduced to the column through a Jameson downcomer via a variable-speed pump. Because of pump problems, the Jameson downcomer column feed system was bypassed in the commissioning stages and the column operated with a standard sparging system. In 1990, Cominco tested the Jameson cell at the Sullivan concentrator in British Columbia, Canada, to evaluate the technology, potentially for the Red Dog concentrator in Alaska, United States. The Jameson cell was found to perform similarly to, but not better than, the existing mechanical cells at Sullivan. Size recovery analysis was evaluated on some of the Jameson product streams. The fine bubble size and high flotation rate suggested superior performance for the separation of fine (<20 µm) particles. No such superior performance, compared with the existing mechanical cell performance, was observed (Fairweather 2005). A Jameson cell was also tested as part of the Pasminco column flotation program in 1989. As discussed by Grimes (1991), it was unsuccessful because of the materials transportation difficulties presented by the coarse grind and high solids densities encountered in the company's South Concentrator zinc cleaning circuit. Several modifications to the pumping system failed to overcome the problems, and work was terminated in favor of devoting more time to the conventional column test program.

In 1989, two 3.0-m-diameter Jameson cells were installed at Mount Isa's new Hilton lead–zinc concentrator. The installation represented a major innovation. The lead rougher and rougher/scavenger concentrates were cleaned, with the Jameson cells baffled in such a way as to create four cells in series (Rohner 1993). Because of the larger diameter, an internal circular launder was installed. In recognition of earlier issues with unstable performance because of fluctuating feed pressures, a complex control logic was implemented to maintain pump box level and pressure. Incorrect downcomer positioning between the inner and outer concentrate launders resulted in low-grade concentrate being produced, and the inner launder was eventually closed. The pressure control logic did not work effectively, and the high percent solids of the feed resulted in low air entrainment rates. Although performance was generally below expectations, the installation did provide knowledge allowing the successful implementation of downcomer placement and internal launder design in subsequent Jameson cells.

The concept of control logic linking the feed pump level, dilution water, pump speed, and downcomer pressure was expanded for two Jameson cells installed in a zinc cleaning duty at the Kidd Creek mine in Canada. The logic followed similar operation for hydrocyclones, varying pump box water addition, varying pump speed, and automatically opening or closing downcomers depending on pressure. The system was inherently unstable, and the company eventually decided the Jameson cells could operate better on manual control.

In 1989, successful tests were conducted at the Peko Wallsend copper concentrator in Tennant Creek, Australia, which resulted in two 1.4-m-diameter cells being installed in December 1989 in a final cleaner role (Jameson et al. 1991). This represented the first installation in copper. The most significant Jameson cell installation of this period was at Newlands coal preparation plant in Australia where in 1988 and 1989, six 1.5-m × 3.5-m rectangular flotation cells were commissioned.

Over the next several years, the number of Jameson cells installed increased rapidly, such that by 1994 it had become the dominant nonmechanical flotation cell in use in mineral froth flotation. It remained in this position until 1999 and, during this five-year period, represented approximately 41% of installed



Source: Cowburn et al. 2005
Figure 14 Downcomer developments from 1989–2000

pneumatic plus column flotation capacity. Installations of note were Philex Mining copper cleaning and roughing circuits, treating 1,000 t/h; the Maricalum Mining copper cleaning and roughing circuits, treating 750 t/h; and the Alumbrera copper cleaner circuit (Harbort et al. 1994, 1997, 2000). During this time, the Jameson cell achieved almost total dominance in the Australian coal industry (Murphy et al. 2000), with installations at Blackwater, Goonyella, Riverside, North Goonyella, and Hail Creek, to name a few.

The period 1990–1999 was a time of extensive experimentation and development, with much conducted on operating sites. Initial work related to stabilizing the downcomer feed pressures to give stable flow and air entrainment. Typical feed pressures required are 150 kPa, to achieve a jet velocity of 15 m/s. The Jameson cell and feed systems were modified to operate at a higher volumetric throughput than the nominal fresh feed flow, with the difference composed of recycled tailing. Tailing was recycled to maintain a stable feed pump hopper level, either via hydraulic head equalization or via an instrumented control loop. Initially the hydraulic system was used in metalliferous operations, with the instrumented system used in higher flow coal operations. The instrumented system gradually became the standard. The amount of tailing recycled has varied from an initial 25% of downcomer feed to 40% as an enhancement to recovery and may now reach 80%.

In metalliferous operation, orifice wear was also an issue. Materials of construction of the orifice plate were investigated in 1991, including high-chromium hardened steel and various ceramics (Harbort et al. 1994). High-density alumina was deemed to have excellent wear properties and became the standard.

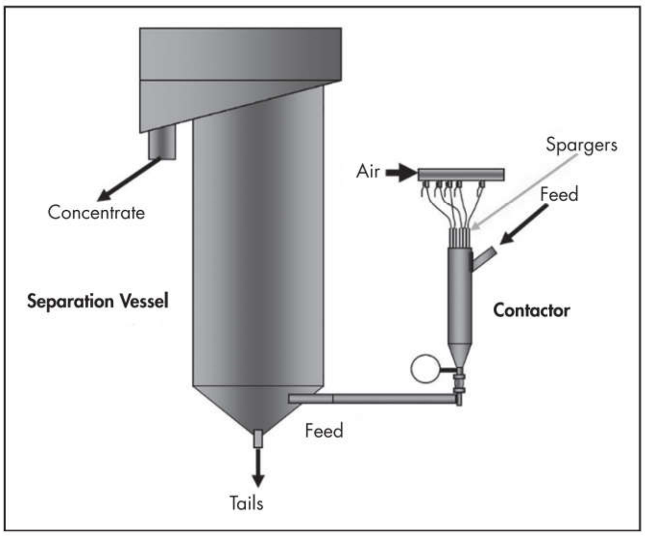
As discussed by Cowburn et al. (2005), a substantial effort was placed on improvement of performance of the nozzle and downcomer, resulting in the 1995 development of the "slurry lens." The major changes in the design are shown in Figure 14, which illustrates the initial steel downcomer and nozzle design, the use of high-density polyethylene, the first use of ceramic lining for roughing duties, and the current Jameson cell arrangement. The location of the slurry lens (instead of the previous orifice plate) allowed an increase in the mixing zone in the downcomer, improving residence time and allowing operation at higher air-to-pulp ratios, from ~0.6 to ~1.2, with superficial air rise velocities (Jg) up to 1.2 cm/s.

Substantial increases in downcomer and tank unit capacities also occurred. The cell diameter increased from 3.5 m in 1993 to 6.5 m in 2000, with downcomer capacities increasing from 30 m^3/h to 90 m^3/h. The larger flows increased the interaction of aerated slurry exiting neighboring downcomers, causing increased pulp turbulence. Significant testing was conducted to optimize the shape, location, and porosity of the bubble diffuser to minimize this turbulence. Diffusers were installed in an attempt to create uniform bubble rise velocities across the surface of the cell by slowing the superficial gas velocity in the high void faction area immediately around the downcomer.

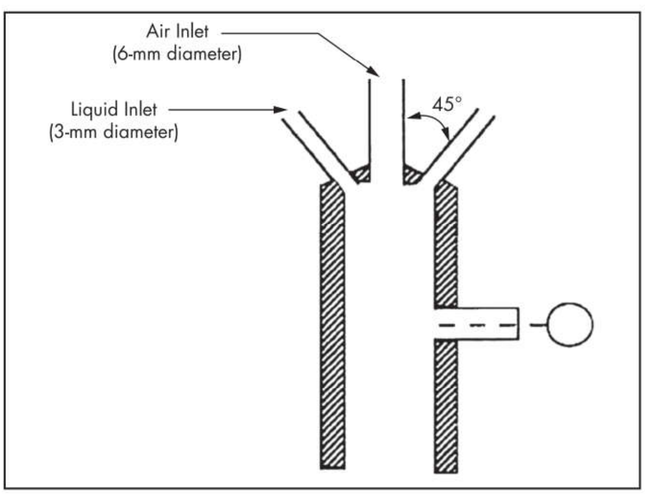
The froth zone and carrying capacity limitations were also an area of investigation considering the Jameson cell's high unit capacity and frequent use in high mass recovery duties such as coal flotation and base metals cleaning. The 2000 review by Murphy et al. provides a detailed description of Jameson cell carrying capacity based on particle size distribution, aeration rate, and cross-sectional area. Both in-froth and above-froth wash-water addition systems have been used in the Jameson cell. In-froth washing produces a drier concentrate and, with washing occurring closer to the froth–pulp interface, allows increased time for bubble drainage in the froth phase, generally increasing washing efficiency. A reported difficulty of in-froth washing is blockage of water holes and uneven distribution of wash water. The current Jameson cell wash-water practice is to use above-froth wash-water addition systems in a shower arrangement. The increased use of recycle and higher Jg has resulted in greater dependency on wash water to achieve specified concentrate grade. Where earlier installations operated at a wash water to concentrate water ratio of 0.8:1.2, the modern norm is to operate with ratios of 1.2:1.5.

In 2000, Glencore Technology made a strategic decision to largely withdraw from metalliferous duties and focus on coal flotation. This resulted in the Jameson cell achieving wide acceptance in coal not only in Australia but also in the United States. During 2000–2012, 76 Jameson cells were installed in various coal plants. Jameson cells were also installed in the burgeoning Chinese coal industry, with varying success (Ren and Yang 2015).

By 2006, the Jameson cell had commenced a return to base metals flotation (Young et al. 2006). In addition to traditional cleaning roles, focus was placed on preflotation and scalping duties. Preflotation involves removing a fast-floating gangue prior to the main flotation circuit, rather than floating and then depressing the gangue material. Scalping involves recovering rapidly floating material to produce a final-grade concentrate at the head of either roughing or cleaning circuits, provided effective liberation and the correct electrochemical conditions exist. After scalping at the head of the rougher circuit, the valuable mineral remaining in the circuit is floated as a lower-grade concentrate that is then upgraded in the cleaner circuit. The removal of the liberated mineral prior to a regrinding stage allows the regrind and cleaning circuits to be designed and operated more appropriately to the middling material. This allows a greater efficiency of separation of composite particles. Minimizing the quantity of regrinding decreases slimes generation and reduces the losses that inevitably result from their presence (Gray et al. 1998; Pokrajcic et al. 2005). Huynh et al. (2014) also discuss the use of Jameson cells for the removal of penalty elements and their effect on circuit configuration. Jameson cell operation in prefloat duties include the Red Dog Pb-Zn installation (Smith et al. 2008), the Mount Isa copper concentrator (Carr et al. 2003), and the Century Pb-Zn mine (Rantucci et al. 2011). Operations in the scalper duty include the Phu Kham copper concentrator in Laos (Bennett et al. 2012), and the Telfer copper–gold concentrator (Seaman et al. 2012) and Prominent Hill copper concentrator, both in Australia (Barnes et al. 2009).



Adapted from Amelunxen 1993
Figure 15 Contact cell arrangement



Source: Ityokumbul 1998
Figure 16 Early contact-cell aerator arrangement

CONTACT CELL

The contact cell was invented in 1992 by Roger Amelunxen as a high-capacity machine capable of producing high-purity concentrate grades (Amelunxen 1993). Feed slurry is placed in direct contact with air in a pressurized chamber outside the flotation tank (Figure 15). The resultant aerated pulp is released into the flotation tank where the mineral-loaded air bubbles are allowed to float and separate from the gangue. As with other pneumatic flotation cells, the sizing of the tank is not a function of residence time but rather mixing characteristics and carrying capacities of the froth. Figure 16 illustrates the arrangement of an external contactor tested by the Florida Industrial and Phosphate Research Institute (Ityokumbul 1998). The contactor is operated at a typical pressure of 140 kPa, with tank residence time reported being two-thirds to one-half of mechanical flotation cells.

Contact cells have been installed in several sites, treating various commodities. In Canada they have operated at the Agnico Eagle and Eskay Creek mines in copper cleaner duties. In the United States, they have operated at the Mountain Path rare earths operation in California and in Phoenix, Arizona, in copper cleaning. Four contact cells were installed in the Hustadmarmor mine in Norway for lime flotation.

At the Bulyanhulu concentrator in Tanzania, two contact cells were installed to treat gravity gold tailing. At the start of operations, frequent transitions between quartz-dominant ore and argillite-dominant ore impacted throughput and flotation performance. To mitigate these effects, the contact cells were added to recover gold from the grinding and flash flotation circuits. Figure 17 details a cutaway of the Bulyanhulu contact cell, mapping air holdup within the tank, which indicated a complex aeration pattern. A high air holdup and potentially turbulent region near the distributor outlet was evident, although this was rapidly dissipated. An area devoid of air exists at the lower extent of measurement at the cell wall. It is expected that this air-barren region would extend into the lower third of the tank below the distributor. The aerated pulp distributor is designed so the pulp exits at both ends, and this resulted in a reduction of the air content in the center of the tank. This was sufficient to cause air from high air holdup regions to move toward the center of the cell. There is also some distortion caused by the internal launder. The expected result is deterioration in the concentrate quality near the internal launder, with the problem increasing with higher aeration. At the time of the measurements, the internal launder was lidded and closed because of concentrate quality problems (Harbort and Schwarz 2007).

JET FLOTATION MACHINES

China has a long history of pneumatic flotation cells, as discussed by Hu and Liu (1988). One of the longest standing types of pneumatic cells is the jet flotation machine, under the models XPM, FJC, and FJCA.

The XPM series was first put into operation in 1967 at the Nanshan Colliery. Initially only a limited number of XPM jet flotation machines were installed in industry, although by the year 2000, 89 cells had been installed, overwhelmingly in coal flotation, with sizes ranging from 4 to 16 m^3. Reasons for the large number included the small unit capacity and the number of mechanical flotation cells that were retrofitted with the jet aerators.

In each cell, four aerators were arranged radially. A recirculating pump withdrew part of the pulp from the flotation tanks and pumped it through a jet chamber. Guide vanes within the jet created a circular motion, reportedly enhancing performance. The jet generated a vacuum, drawing in air. The aerated pulp passed through a downcomer prior to being dispersed via an umbrella into the flotation tank. The flotation tanks acted as a bank, with feed flowing down the bank to a regulated tailing outlet (Wu and Ma 1998). Figure 18 provides details of the FJC jet flotation machine, which has essentially the same arrangement as the XPM.

From 2000 to 2010, the number of jet flotation machines installed in China grew. During this period, nearly half of all pneumatic and column flotation machines installed worldwide were installed in China. Approximately 180 FJC jet flotation machines were installed in the Chinese coal industry during this period. However, the machines suffered from high wear in the downcomer and umbrella diffuser.

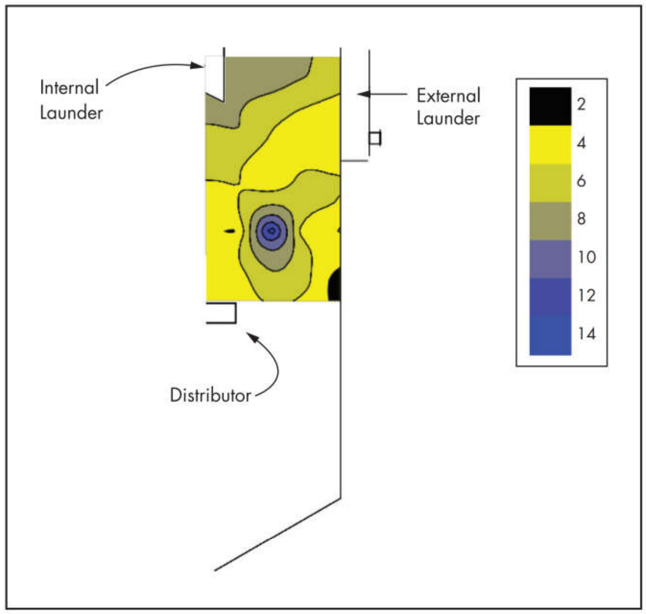
In 2006, the aerator mechanism was redesigned, and the new jet flotation machine was designated the FJCA. As shown in Figure 19, the FJCA design immersed the aerator, dramatically shortened the downcomer, and removed the umbrella diffuser. FJCA machines of 20-m^3 unit capacity were successfully installed in the high-throughput Linhuan and Wanfenanggang coal preparation plants with good aeration performance (Wu et al. 2010).

FLOKOB CELL

Major restructuring of the Polish coal mining industry commenced in 1990. The change from a centrally planned to a market economy enforced a different method for managing the coal mining industry. Unprofitable coal mines were gradually closed. In 1995, 65 coal mines were operating, but in 2004, only 41 coal mines existed, in which 43 coal preparation plants operated. There were also eight independent coal mines in Poland, which were closed coal mines purchased by new owners (Horsfall et al. 1994). Polish coals were difficult to float, and the FLOKOB pneumatic cell gained wide use in the country (Blaschke 2000; Blaschke and Gawlik 2004).

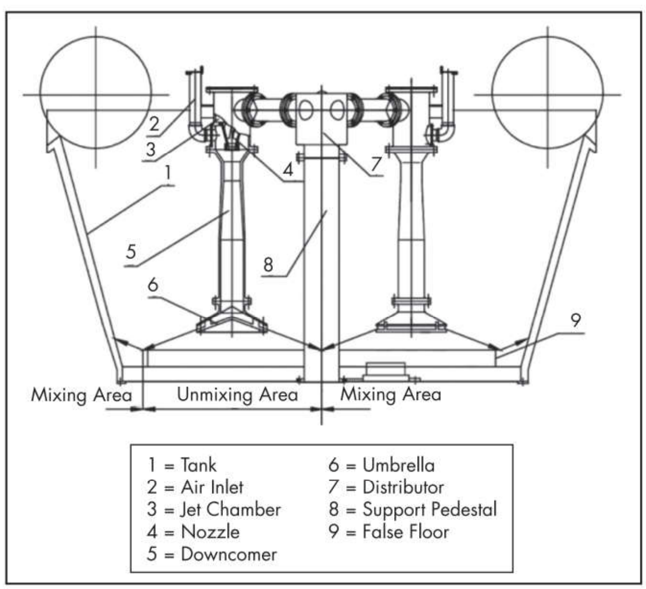
Research into the conditions and possibilities for the use of pneumatic flotation machines for the treatment of coal slurries commenced in the early 1980s in Poland (Brzezina 1982). By 1990, the Central Mining Institute in Katowice was at the leading edge of pneumatic cell flotation research, with the FLOKOB cell introduced to the Polish mining industry in 1991.

The publication by Brzezina and Sablik (1991) details the development and experimentation of the FLOKOB-3, a 3-m^3 flotation cell with a capacity of 60 m^3/h. Tests were conducted on various coals and investigated capacity, aeration, and immersion of the aerator downcomer. By 1996, the FLOKOB cell had successfully been scaled up to a 40-m^3 machine with a capacity of 520 m^3/h. Installations of significance included the Szczyglowice mine and the Rydultowy colliery preparation plant.



Source: Harbort and Schwarz 2007

Figure 17 Air holdup (%) within a contact cell



Source: Wu et al. 2010

Figure 18 FJC jet flotation machine

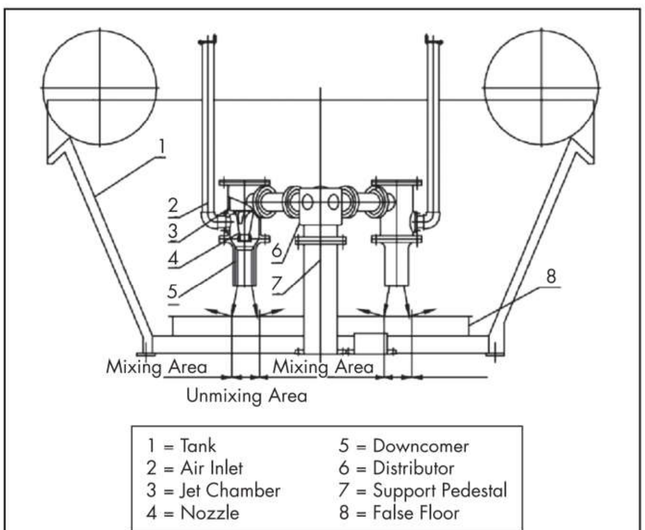
The FLOKOB cell operated with a dual aeration system, as shown in Figure 20. Feed was distributed to several aeration units. The slurry (at 120 kPa pressure) was mixed with air supplied at 30 kPa and entered a 1-m-long downcomer before entering the flotation tank. A portion of tailing was extracted from the cell and fed into a secondary aeration system submerged in the lower section of the cell (Brzezina and Sablik 1994).

AIR-SPARGED HYDROCYCLONE

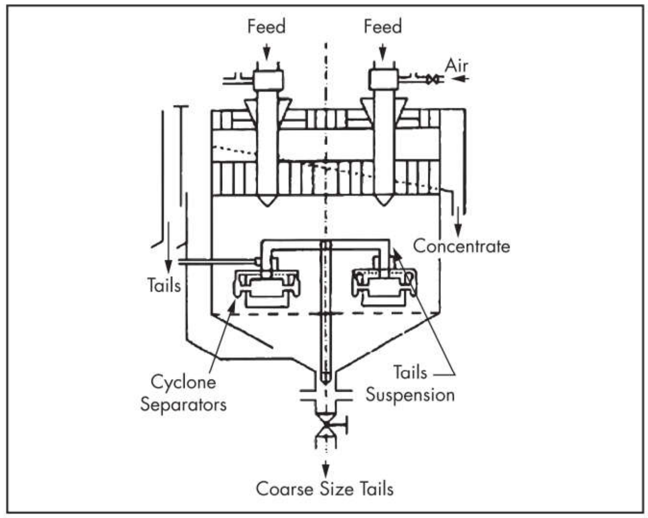
One of the most extensively investigated pneumatic flotation machines is the air-sparged hydrocyclone (ASH). The ASH was initially developed at the University of Utah based on the concept of Miller (1981) with initial work supported by postgraduate researcher Van Camp (1981).

The basic features of an ASH are a porous wall through which air is sparged and a tangential flow of slurry orthogonal to the airflow. The ASH was envisioned to take advantage of a controlled high force field developed by cyclone fluid flow, together with a high density of bubbles, to achieve effective flotation of fine particles, with an increase in flotation kinetics. As discussed by Miller et al. (1982), many different designs incorporating these features were tested prior to a preferred design being selected. As shown in Figure 21 (Ye et al. 1988), the slurry is fed tangentially through a conventional cyclone header, passing through the separator as a thin layer in swirl flow, and travels downward countercurrent to the froth phase, which is moving upward in the center of the device. Hydrophilic particles are thrown to the porous cylinder wall and are discharged in the reject. Hydrophobic particles encounter the air bubbles sparged radially through the porous wall. The high-shear force at the wall generates small air bubbles and provides for intense particle–bubble interaction. Particle–bubble attachment occurs and the hydrophobic particles are transported into the froth phase; froth exits axially at the top of the cyclone through a vortex finder.

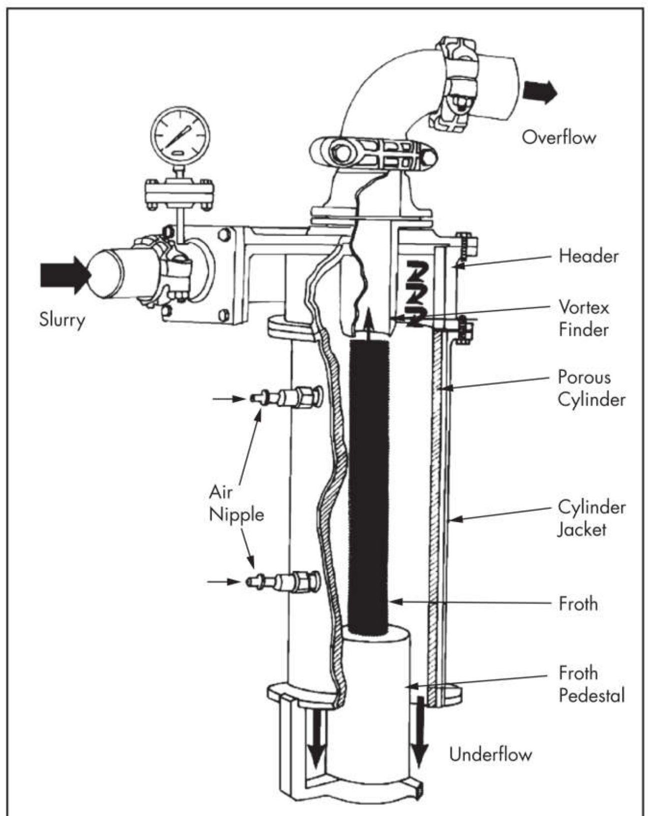
The effective recovery of fine particles is primarily because of the high centrifugal force field present in the ASH, with force fields of the order of 100 g increasing the inertia of the fine particles and facilitating the effective flotation of particles as small as 1 µm. The effectiveness of fine particle flotation is further enhanced by the presence of numerous, freshly formed, small air bubbles. An ASH bubble generation study by Lelinski et al. (1995) examined the influence of surfactant concentration, water flow rate, and porous tube pore size. Depending on the experimental conditions, the average bubble size ranged from about 100 to 300 µm in diameter.



Source: Wu et al. 2010
Figure 19 FJCA jet flotation machine



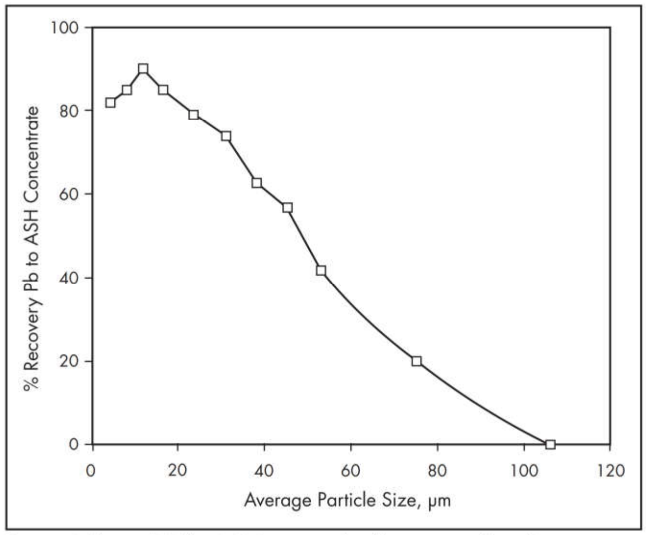
Source: Brzezina and Sablik 1994
Figure 20 FLOKOB flotation cell



Source: Ye et al. 1988
Figure 21 Air-sparged hydrocyclone

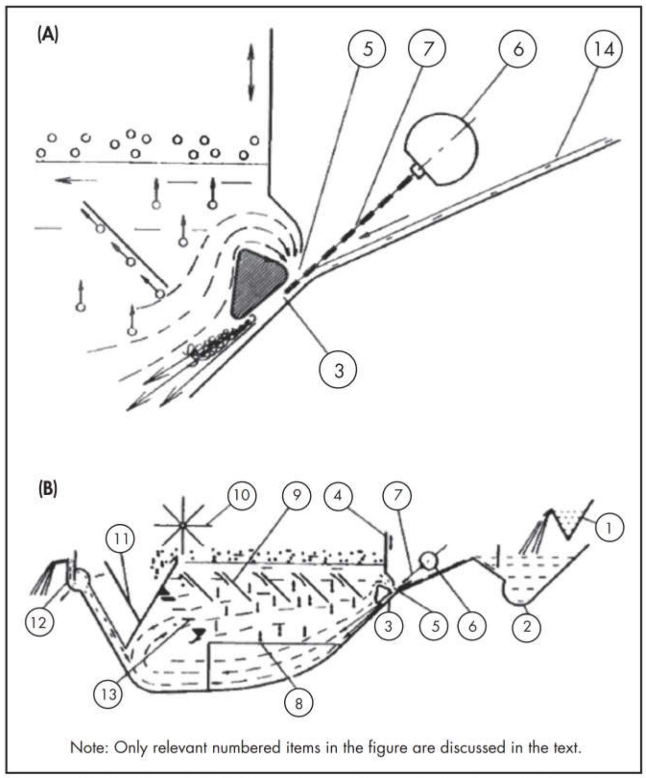
The ASH is a machine of immense potential, although that potential has not resulted in significant installations in the minerals industry. In general, for most flotation testing with the unit, a relatively high reagent addition is required to achieve the same recovery as that achieved in mechanical bench-scale flotation. This high reagent demand was thought to arise because of the high centrifugal force generated by swirl flow during ASH flotation. Higher reagent addition is required to stabilize the bubble–particle attachment in the relatively high centrifugal field. With larger-diameter ASH units, the centrifugal force is smaller, and therefore the reagent addition was expected to be reduced significantly (Miller and Yu 1993). A significant decrease in the concentrate grade occurs if the length of the vortex finder exceeds a critical value. Some plant-site continuous operation of the ASH was limited to about six hours because of crud formation on the inner surface of the porous tube. This crud gradually built up, plugged the pores of the porous tube, and caused poor flotation separation apparently because of inadequate air distribution (Miller et al. 1999).

Many successful pilot tests have been reported. Miller et al. (1986) discuss gold flotation tests where a feed grade of 0.34 g/t Au was upgraded to 170 g/t, at 75% gold recovery. The results were achieved with a residence time of less than 1 second, compared to a mechanical bench-scale test that achieved a comparative performance in 10 minutes. Miller and Van Camp (1982) report coal flotation where comparable results were achieved with an ASH at <0.5-second residence time and with mechanical flotation having a 2-minute residence time. Pyrite was effectively recovered from a coarse Witwatersrand (South Africa) quartzitic ore with an ASH. Recoveries of pyrite from 85%–93% at grades of 35%–40% sulfur were obtained. The optimal recoveries and grades were obtained for particles between 38 μm and 106 μm. The results were achieved with a residence time of <1 second and were comparable with those attained in mechanical batch flotation tests after 5 minutes (Van Deventer et al. 1988). At the Cadjebut operation, Australia, the major reason for losses of galena during lead beneficiation was that the circuit was unable to recover the galena present in the finest size fractions. A pilot ASH test significantly enhanced performance in lead roughing recovery for particles smaller than 20 μm, as shown in Figure 22. Although the tests were not progressed, "ASH based flotation was considered to be viable and of considerable potential benefit to the Cadjebut concentrator … at high specific capacity without necessitating major alterations to the overall flotation reagent consumption" (Baker and Willey 1993).



Source: Baker and Willey 1993, reprinted with permission from the Australasian Institute of Mining and Metallurgy

Figure 22 ASH size-by-size lead recovery at Cadjebut



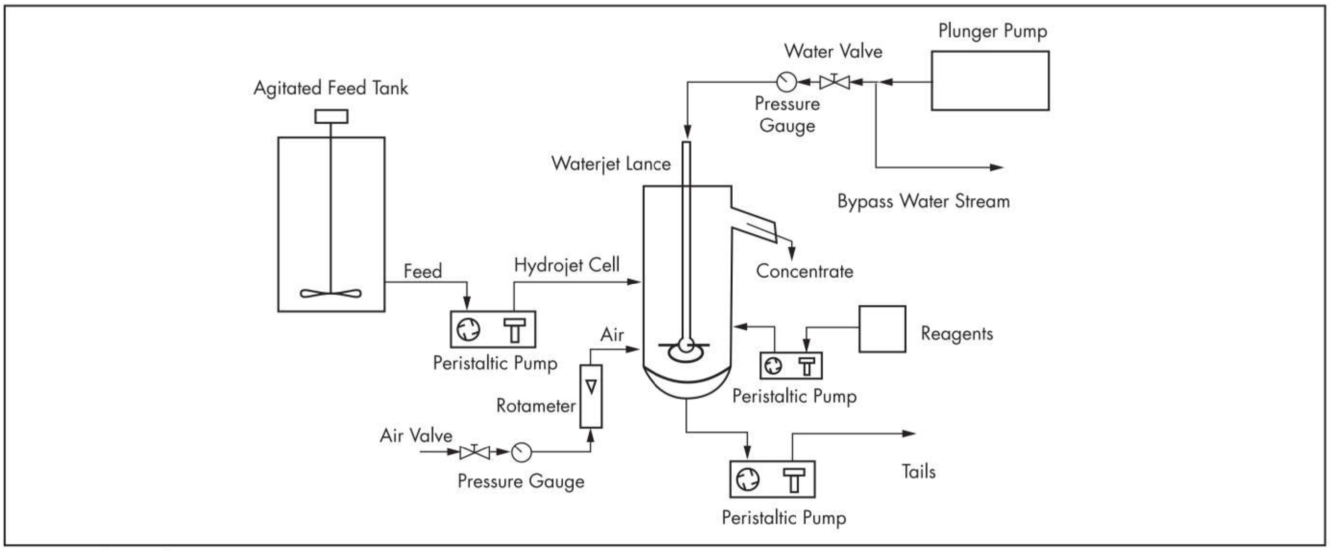
Source: Chudacek et al. 1994, reprinted with permission from the Australasian Institute of Mining and Metallurgy

Figure 23 FASTFLOT (A) aerator and (B) cell arrangement

FASTFLOT CELL

The FASTFLOT pneumatic flotation cell underwent extensive plant testing at the Pasminco Broken Hill operations, Australia, during the 1990s (Chudacek et al. 1994, 1995).

In the machine, high-velocity water jets are used to entrain air and then enter the flotation tank generating froth. The FASTFLOT arrangement and operation are shown in Figure 23. Mixing intensity, shear rate, and acceleration levels were controlled by the velocity of the thin jets (7), controlled by the operating pressure in the manifold (6). The jets travelled at approximately 50 m/s, accelerating the surrounding air envelope before plunging between two layers of mineral slurry (5) and entering a diverging mixing nozzle (3). The shear rate, because of the velocity differential between these three streams, resulted in shredding of the gas envelope into very small bubbles and provided very intense agitation accompanied by high turbulence. The aerated flow was discharged into the cell bulk compartment where it flowed past longitudinal stabilizers. Froth was diverted by bubble guides (9) toward the froth discharge end of the cell. As the layer of froth travelled toward the froth weir, it drained liquid and



Source: Carbini et al. 2001
Figure 24 Hydrojet experimental arrangement

entrained gangue mineral particles and thickened because of crowding against the weir. It was subsequently discharged by a rotary paddle (10) into the froth launder (11). The slurry flow at the discharge end of the cell was partly released via flexible ducts into the discharge launder (12), which was adjustable vertically to control cell liquid level. The rest of the slurry flow was diverted upward by a curved bottom where, in turn, the slurry flow was diverted horizontally by a flow guide (13) toward the recycle gate (4) that controlled the recycle flow (5).

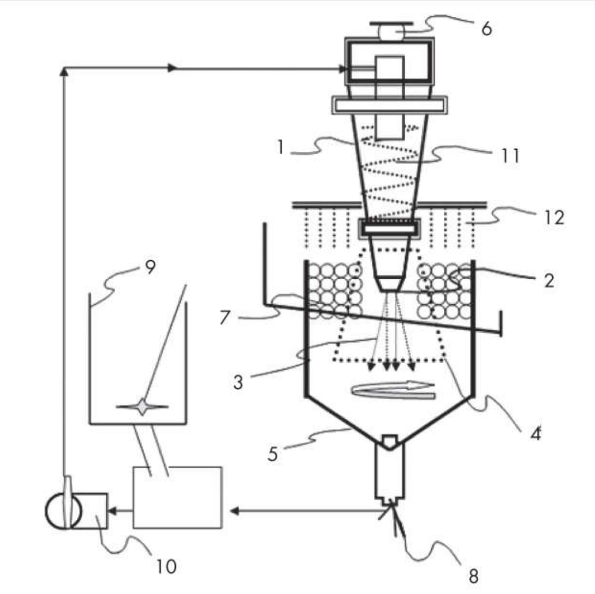
Operation at Broken Hill with a 1.5-m^3/h pilot cell achieved rougher zinc recoveries 10% higher than the operating plant at comparable concentrate grades. The FASTFLOT residence time was one-half to one-third that of the plant.

HYDROJET CELL

The Hydrojet cell also made use of high-pressure water jets for flotation. Developed at the Department of Geoengineering and Environmental Technologies, University of Cagliari, Italy, it made use of high-pressure (9 MPa) water lances immersed in a flotation tank. The lance exit was close to the curved base of the tank where compressed air also entered (Figure 24). The velocity of the water jet exiting the lance was 100 m/s, generating high turbulence and fine bubble formation. Successful tests were conducted on coal, barite, and zinc roughing. The comparative test on zinc roughing indicated that the Hydrojet cell in comparison to mechanical cell flotation would produce 10% higher recovery at equivalent concentrate grade and residence time (Carbini et al. 2001).

CYCLOJET CELL

The Cyclojet cell is a high-intensity jet, pneumatic flotation cell developed in Turkey in 2006. Figure 25 provides an arrangement of the cell. As described by Hacifazlioglu and Kursun (2012), feed pulp at a pressure of 10–60 kPa is pumped to the inlet of a hydrocyclone (1) in which the overflow has been closed (6). The tangential entry and conical shape generate centrifugal forces, resulting in a conic pulp jet (3) exiting the cyclone apex (2). The conic jet enters a conic tube (4), two-thirds of which is submerged in the flotation tank (5).

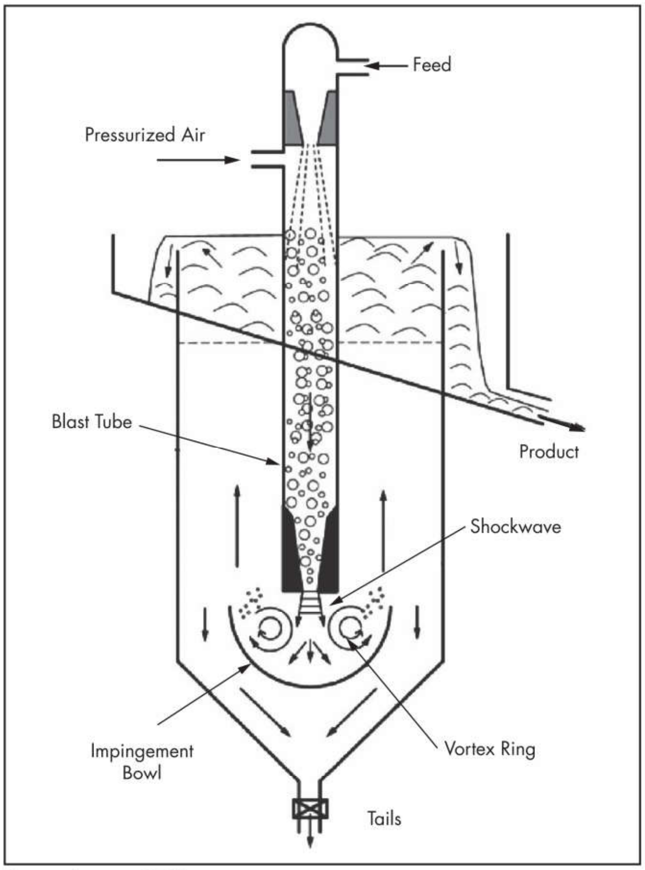


Note: Only relevant numbered items in the figure are discussed in the text.

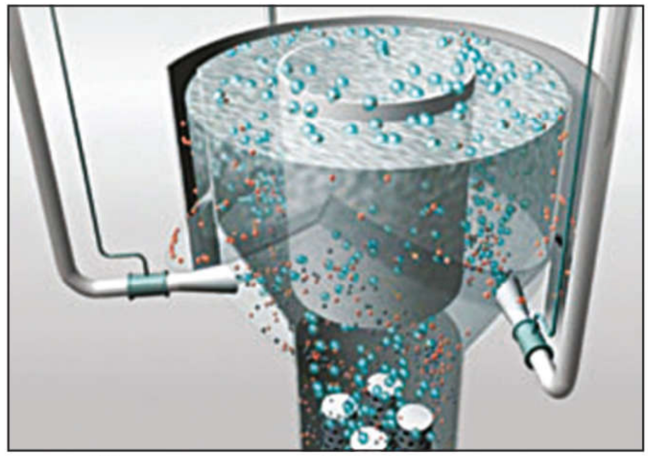
Source: Hacifazlioglu and Kursun 2012
Figure 25 Cyclojet arrangement

An air gap is created by the effect of the cyclonic jet within the tube, generating a vacuum and allowing the suction of air from the atmosphere via a restricted air entry in the tube. The shear forces generate bubbles reportedly 0.3–0.5 mm in diameter. The aerated pulp exits the conic tube into the tank where disengagement of bubbles occurs to form a typical froth layer (7). Froth washing (12) can be utilized and a portion of the tailing (8) can be recirculated to the feed to maintain stable feed pressures.

The Cyclojet cell has been successfully tested in coal, barite, and feldspar flotation. Comparative tests between the



Source: Jameson 2010
Figure 26 Concorde cell



Courtesy of Siemens
Figure 27 Simine Hybrid Flot cell

Cyclojet cell and the Jameson cell are reported to have produced similar recovery and concentrate grades (Demeekul et al. 2016).

CONCORDE CELL

In the Concorde cell pneumatic machine, preaerated feed is raised to supersonic velocities before passing into a high-shear zone in the flotation cell (Jameson 2010). A diagrammatic sketch of the Concorde cell is shown in Figure 26. It consists of two stages. In the first, small bubbles are formed in a blast tube under pressure. The feed enters as a vertical jet and mixes with air under pressure. In the second, the aerated mixture then passes through a choke. Downstream of the choke, a shock wave is created where there is a large change in pressure over a small spatial distance, and the bubbles reduce in size. Further particle–bubble contact occurs in the vortex ring that forms in the impingement bowl. Pilot-plant tests on platinum group metals in South Africa indicated that the rate of flotation in the Concorde cell was 100 times faster than the existing mechanical cell circuit.

SIMINE HYBRID FLOT CELL

The Simine Hybrid Flot cell was developed in 2007 by the Siemens Industrial Solutions and Services Group in the framework of a joint pilot project with the Minera Los Pelambres copper mine, Chile (Gaete 2007). As illustrated in Figure 27, feed pulp and gas are mixed inside the mixing chambers of an ejector system before being sprayed by high-pressure nozzles into the cell. A sparger system is installed in the lower section of the flotation tank to enhance recovery. The cell requires up to 70% less power when compared to systems of a similar capacity of 100–400 m^3/h (Krieglstein 2009).

Tests at the Minera Los Pelambres mine showed that the Simine Hybrid Flot cell in a selective copper–molybdenum process could achieve an increase in molybdenum recovery of more than 2% over the existing operation. The recovery increase was significantly in the –10 µm fraction (Siemens AG 2009). Two 16-m^3 Simine Hybrid Flot cells were installed as pre-roughers in the plant's copper–molybdenum selective process. A similar arrangement was planned for Codelco's Andina Division copper–molybdenum circuit. Successful piloting as an additional rougher line in the copper–molybdenum selective process was also conducted at Codelco's El Teniente Division (Carter 2011).

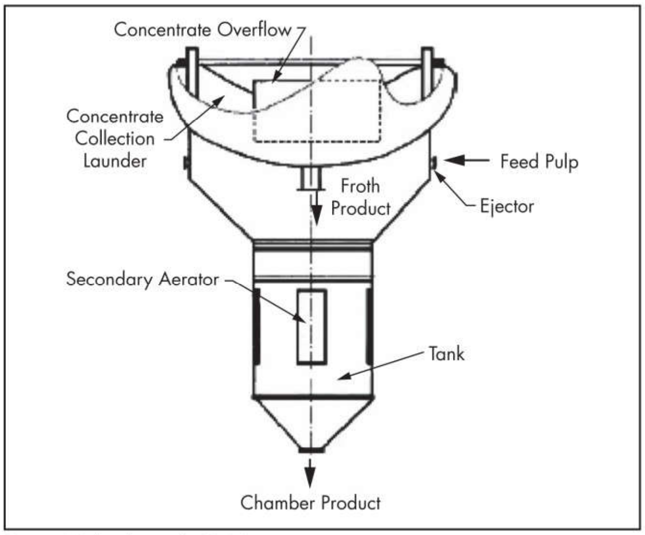
Ongoing development led to a new ejector working in a self-priming operation with a pulp pressure of 3.2–3.4 bar (Krieglstein et al. 2014). Bubble sizes of 500–750 µm have been reported (Hartmann et al. 2012), with a residence time approximately half that for installed mechanical cells.

CFM CELL

Another pneumatic machine that utilizes the dual aeration concept is the CFM cell. The CFM series was developed by the Institute of Material Studies and Metallurgy, Ural Federal University, Russia.

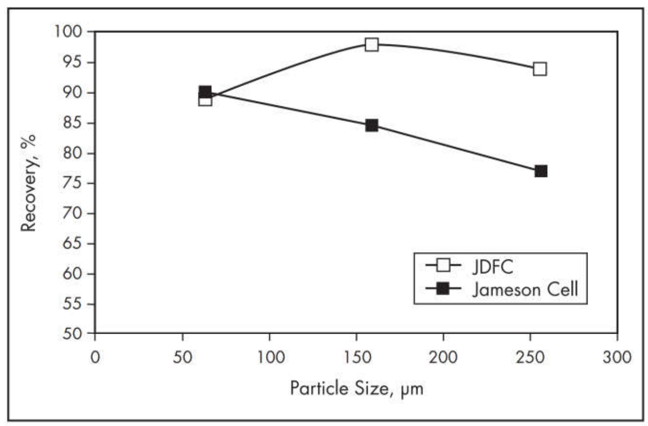
As detailed in Figure 28, pulp and air are mixed in ejectors and injected via a nozzle in the upper portion of the cell. Secondary aerators are positioned in the lower portion of the cell to enhance performance. An industrial CFM-1400VMG has a working chamber volume of 13.2 m^3 and a capacity of up to 500 m^3/h. Air consumption is up to 600 m^3/h, with a typical power consumption of 6 kW·h. The upper chamber diameter is 3.5 m, the lower chamber diameter is 1.4 m, and the height is 5.0 m. A residence time of 3.2 minutes has been reported to produce comparable performance to that of a mechanical cell circuit with a residence time of 10 minutes.

Four CFM-1400M cells have been installed at the JSC Svyatogor processing works (Russia) in various duties, including zinc and copper roughing. They have provided a 3.2% Cu



Source: Viduetsky et al. 2014
Figure 28 CFM pneumatic flotation machine (with ejectors and secondary aerators)

Source: Oteyaka et al. 2014a
Figure 29 Jet diffuser flotation cell modified downcomer exit



Source: Oteyaka et al. 2014a
Figure 30 Comparison of jet diffuser flotation cell and Jameson cell talc recovery by size

and a 2.1% Zn increase in plant concentrate grades, coupled with a 5.4% increase in Cu recovery and a 5.5% increase in Zn recovery (Viduetsky et al. 2014).

JET DIFFUSER FLOTATION CELL

The jet diffuser flotation cell (JDFC) was developed at Kütahya Dumlupinar University, Turkey, during a research project to increase coarse particle recovery in Jameson cells (Öteyaka et al. 2014a, 2014b). The main aim of the new design was to provide quiescent flow and more homogeneous dissipation of bubbles from the downcomer exit and decrease the turbulence occurring in the flotation tank. This was achieved by altering the shape of the downcomer exit to a fluted exit such that the diameter of the exit was approximately double the downcomer diameter (Figure 29). Pilot-scale operation of the JDFC was compared with an unmodified Jameson cell floating talc (Figure 30), indicating improved performance for particles >50 µm. In addition, the bubble plume exiting the downcomer decreased in length from two downcomer diameters for the unmodified downcomer to approximately half a downcomer diameter for the JDFC, reducing the risk of mineral-laden bubbles being carried into the tailing stream.

MODIFIED JET FLOTATION CELL

The modified jet flotation cell (MJC) was developed as part of a Brazilian project evaluating the use of Jameson cells for the recovery of oil emulsions from water in offshore oil platforms (Santander et al. 2011). Modifications are described with reference to Figure 31 and included placement of a concentric blind-end tube around the exit of the downcomer to allow floated material to enter immediately into the froth phase and a packed bed at the upper part of the concentric tube to stabilize the froth and facilitate the rise of the oil floc/bubble aggregates. Both laboratory- and pilot-scale tests were conducted with feed streams containing oil droplets of ~10 µm to 20 µm, and concentrations between 100 mg/L and 400 mg/L petroleum. The studies indicated that the MJC achieved an oil recovery of 85%, 5% higher than the Jameson cell.

FREE JET–TYPE FLOTATION SYSTEM

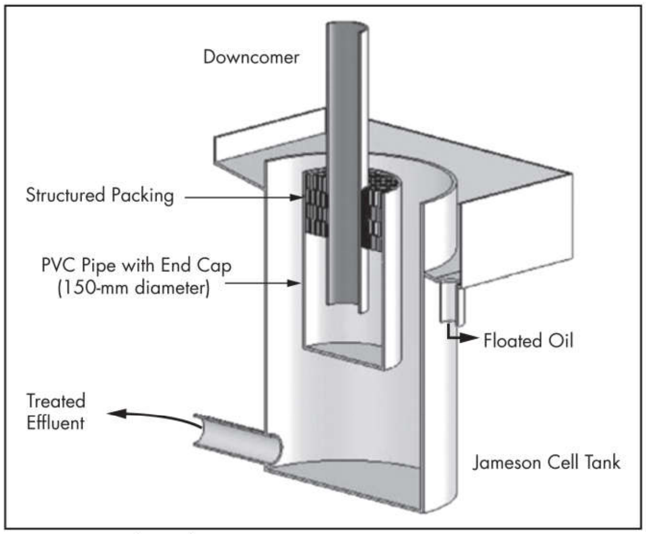
The free jet–type flotation system (FJFS) was evaluated by Istanbul Technical University, based on modifications to free jet flotation cells originally conceived at the Technical University of Berlin in the 1980s (Güney et al. 2016). The FJFS, as illustrated in Figure 32, makes use of a multiple orifice plate–single nozzle, operated at 100–120 kPa, to generate multiple free jets and entrain air from atmosphere via two air inlets. The aerated pulp enters a downcomer whose exit is close to the froth–pulp interface. Below the downcomer, a deflector plate is situated. The distance between the downcomer exit and the deflector creates a secondary turbulence zone. As the distance between the downcomer exit and deflector is increased, recovery increases with a deterioration in concentrate quality. Tests were conducted on asphaltite sourced from the Şırnak-Silopi region in southeast Anatolia, Turkey.

REFLUX FLOTATION CELL

The Reflux flotation cell allows the volumetric feed rate to be increased to nearly 10 times the typical conventional level, achieving low cell residence time, in the order of 25 seconds.

As described by Jiang et al. (2014) and Figure 33, bubbles approximately 500 µm in diameter are generated under a high shear rate within a rectangular, multichanneled, cuboidal downcomer. High liquid fluxes shear bubbles from a sintered surface sparger mounted flush to the channel wall before disengaging the downcomer flow into a vertical column. High feed fluxes, up to 15 cm/s, and high gas fluxes, up to 5.5 cm/s, ensure a high gas holdup beneath the downcomer. Bubble–liquid segregation is achieved using an arrangement of parallel inclined channels incorporated below the main vertical column.

A series of experiments was conducted to investigate flotation of fine coal particles from an extremely dilute feed of 0.35% solids using the Reflux flotation cell. The results demonstrated that the successful recovery of coal particles (92%) was possible (Dickinson et al. 2015).



Source: Santander et al. 2011
Figure 31 Modified jet flotation cell

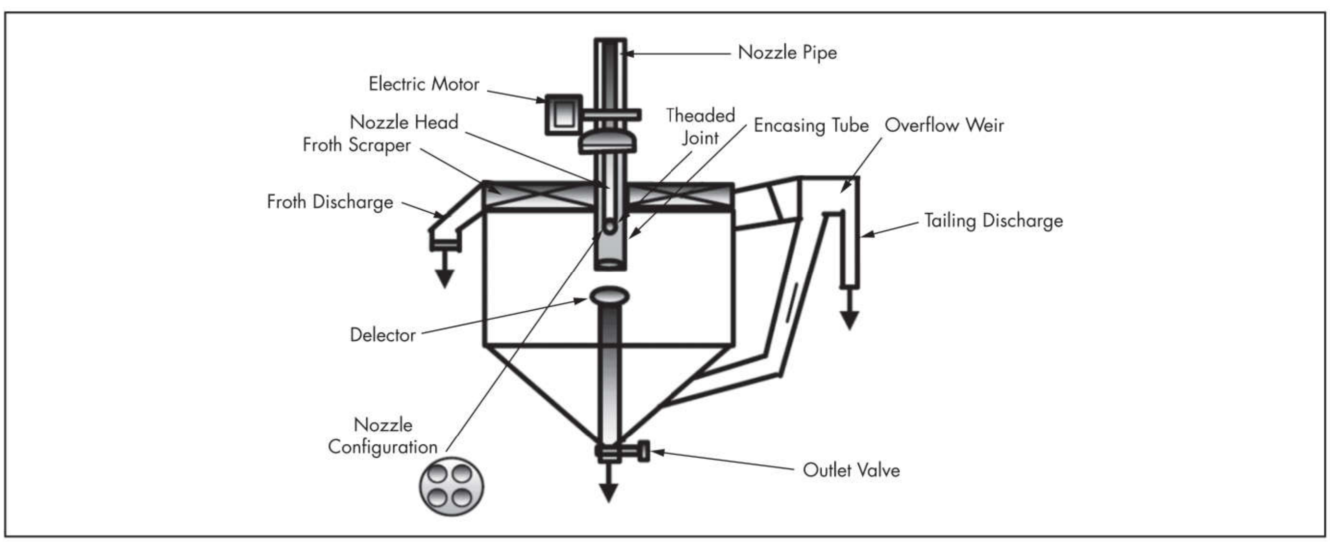
TURBOFROTH

The Turbofroth pneumatic cell was developed in South Africa as an alternative to tall flotation columns. The cell operated a double aeration system where primary aeration was accomplished either through downcomers as described by Steinmuller et al. 2000 (Figure 34) or via an external static mixer system in series with feed pipes that distributed the pre-aerated slurry into the top of the column. A second aeration step was provided at the base of the column via air spargers. The dual aeration system allowed reduction in machine height to between 3 and 5 m. During the 1990s, the Turbofroth was installed in numerous locations in Africa treating coal, copper, lead, zinc, and platinum group metals (Arnold and Terblanche 2001).

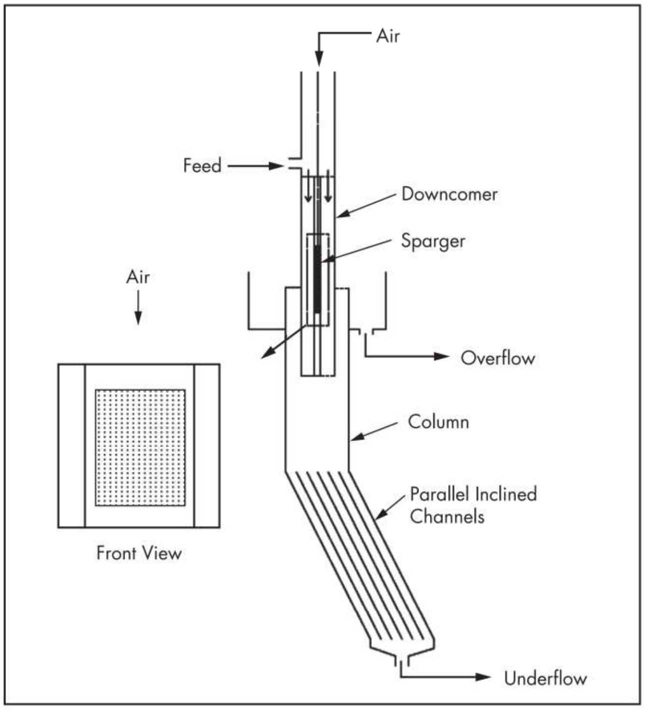
MULTICELL

The Multicell was developed by Anglo Coal during a flotation machine characterization program at Goedehoop Colliery, South Africa (Opperman et al. 2002). Throughout the program, researchers found that high mixing energy was required to achieve acceptable recoveries and optimal use of reagents. During Jameson cell tests, researchers found that the mixing power through the orifice plate in the downcomer was insufficient to achieve the recoveries expected on the Goedehoop fine coal.

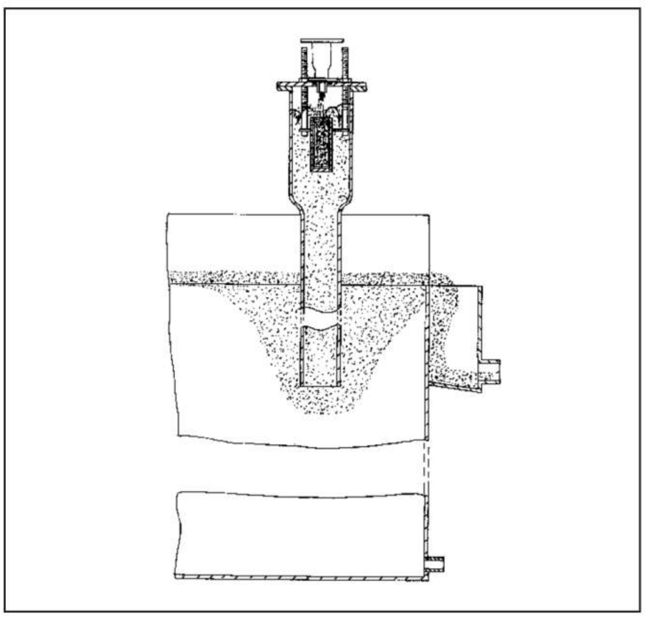
Feed slurry is introduced into the chamber of a vertical spindle pump with its suction positioned at the base. Compressed air is blown into the suction of the pump at a controlled flow rate. Some of this air escapes from the pump alongside the shaft and is sheared by perforated paddles attached to the shaft. The slurry–air mixture exits the chamber through four perforated windows at the bottom of the pump shaft. This forms the first part of flotation in the Multicell and may be considered similar to the mixing and bubble generation zone within a mechanical cell. Pulp flows downward where it is drawn into the pump at the bottom of the tank. Inside the pump, the airstream is sheared into very small bubbles. The mixture of air and pulp is pumped to a second tank through a flow distributor. In this tank, mineral captured in the froth is allowed to separate from gangue material. This tank allows



Adapted from Güney et al. 2002
Figure 32 Free jet type flotation system



Source: Jiang et al. 2014
Figure 33 Reflux flotation cell



Adapted from Steinmuller et al. 2000
Figure 34 Turbofroth pneumatic cell

for an absolute quiescent zone, so little product is dislodged to tails and little mechanical entrainment of gangue in concentrate occurs. The Multicell was successfully scaled up to a feed capacity of 250 m^3/h.

CENTRIFUGAL FLOTATION CELL

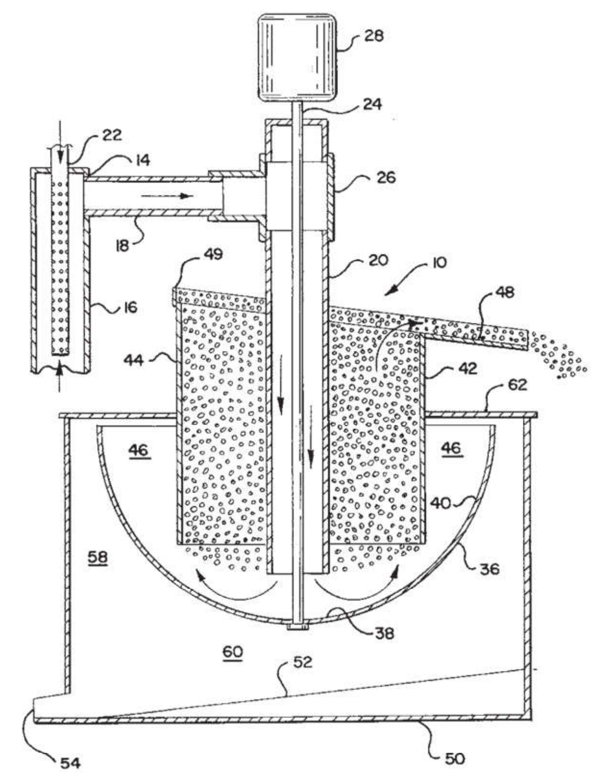
Centrifugal flotation cells (CFCs) represent a series of pneumatic flotation machines where centrifugal forces are applied by various means to enhance flotation. Initially proposed in the 1980s, significant development was undertaken by Clean Earth Technologies in 1995 to separate oil from water. CFCs were subsequently applied to mineral froth flotation. The thesis by Jun-Xiang (2001) provides a detailed review of CFCs' early development and theory of operation.

The first CFC consisted of a rotating drum where separation occurred, a pulp feeding pipe, and an air injector. In practice, the pulp entered at the bottom of the rotating drum near the center. The centrifugal force caused the slurry to migrate outward across a screen of bubbles that were injected at high velocity through a jet at the bottom of the drum. Angular momentum was continuously being transferred to the fluid, and in a short time, the fluid inside the cell formed a rotating motion, creating a centrifugal force field in the range of 50–150g. The mineralized froth had a density smaller than that of the fluid, so it migrated inward, where it overflowed a weir and collected in the froth launder. The tailing was discharged through tailing ports (orifices) along the outer rim of the drum. One of the persistent problems with this CFC was the rapid blockage of the tailing discharge ports in the outer rim of the rotating drum, if the feed slurry had a pulp density higher than 30% solids.

To overcome this issue the CFC-1 was developed. This version is described by U.S. Patent 5,928,125 (Jian et al. 1999) and is illustrated in Figure 35. The conditioned pulp is pumped into the aerator (14) comprising a vertical sparger section (16) with a 2.0-µm compressed air tube sparger (22), a horizontal section (18), and a downcomer (20). The high superficial velocity slurry stream shears off the bubbles that have formed on the sparger wall. The aerated slurry continues downward through the downcomer into the bottom of a rotating bowl (36). An elongated shaft (24) extends vertically through a bearing housing and collar (26) mounted on the upper portions of the downcomer. The upper end of the shaft is connected to a variable-speed motor (28). The lower end of the shaft supports a rotating vessel (36). A froth column (42) with an upright annular wall (44) is placed between the downcomer (20) and the sidewalls of the rotating vessel. The upright annular wall provides a vertical weir that extends to a height above the sidewalls of the rotating vessel. As the vessel rotates, the air bubbles move toward the downcomer, rise to the pulp surface, and exit as overflow through the froth column in a concentrate launder. Tailing moves outward and upward, passes the annular passageway, and exits via the tailing outlet. The final prototype operated with a pulp feed rate of 0.3 m^3/h, an aeration rate up to 0.6 m^3/h, and a bowl rotating at between 100 and 400 rpm.

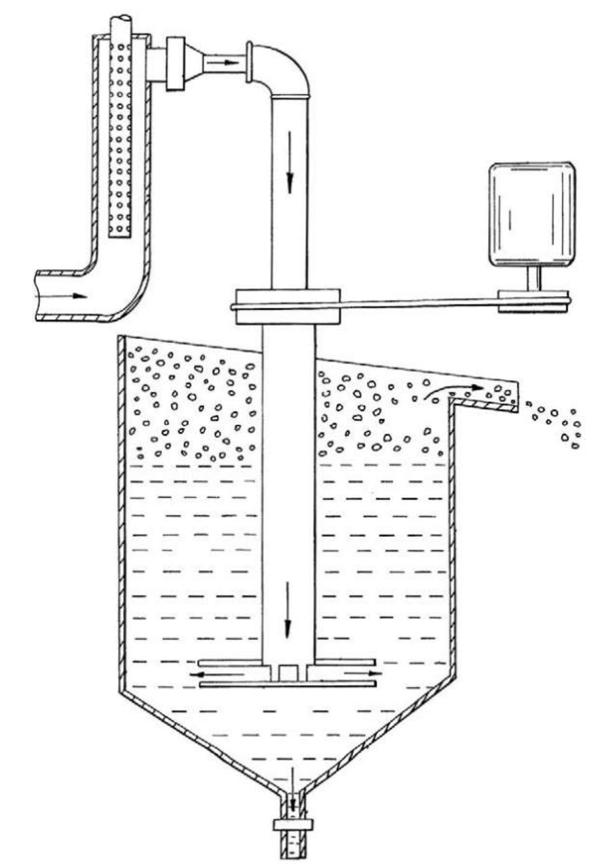
Further development led to the CFC-2, which had a substantially different structure. This version is described by U.S. Patent 6,126,836 (Jian et al. 2000) and is illustrated in Figure 36. The aeration unit was the same as for the CFC-1, but instead of rotating an internal tank, the downcomer was rotated. To prevent downward egress of froth-laden bubbles to tailing, a base plate was installed to close the bottom of the downcomer. Aerated pulp flowed horizontally through downcomer exit ports, with a plate parallel to the base plate enhancing lateral and radial discharge of the stream. Centrifugal force moved the rotating slurry to the wall of the flotation cell where mineral-laden bubbles rose to the froth layer and tailing descended to the tailing outlet.

Laboratory tests comparing the CFC-1 and CFC-2 performance against Denver flotation cell bench tests were



Note: Only relevant numbered items in the figure are discussed in the text.

Source: Jian et al. 1999
Figure 35 CFC-1



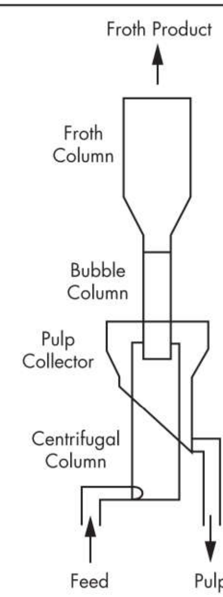
Adapted from Jian et al. 2000
Figure 36 CFC-2

conducted. These indicated that at comparable recovery and concentrate grade, the CFCs when floating galena achieved flotation kinetics 1.5–3.0 times greater than the Denver laboratory cell. When floating chalcopyrite, the CFCs achieved flotation kinetics 3.0–5.0 times greater than the Denver laboratory cell. Ultimate recoveries were comparable, with enhanced recovery of –10 µm mineral.

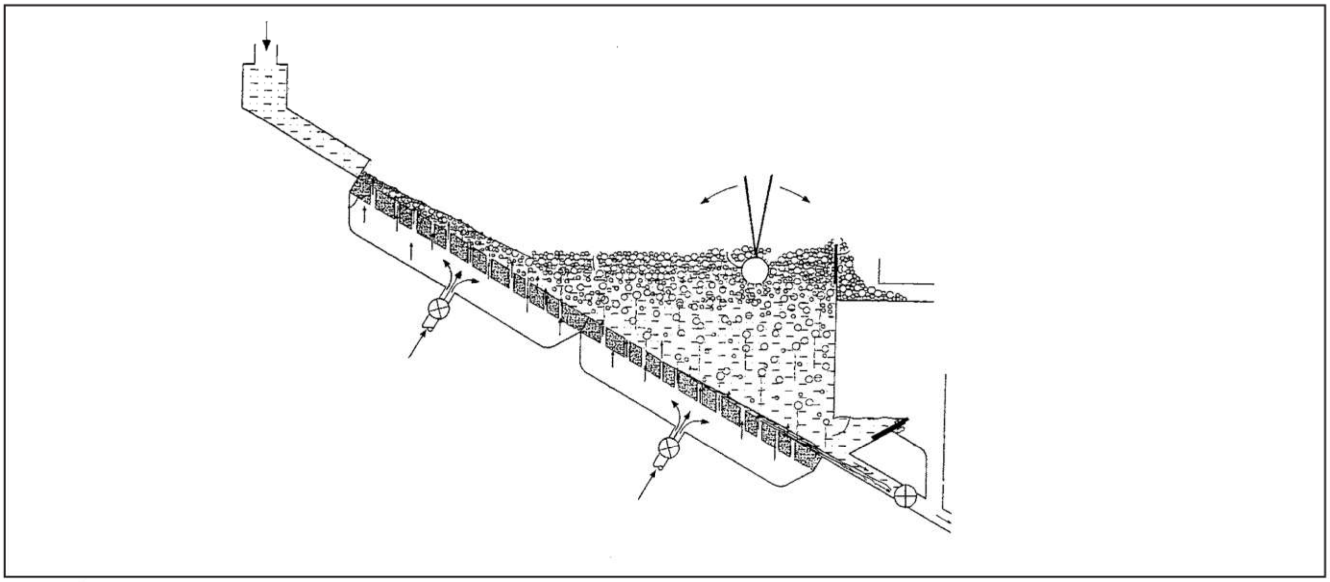
Research into CFCs has continued over the years, with more recent studies conducted by Yazd University in Iran (Ghaffari and Karimi 2012), where a significant computational fluid dynamics study was undertaken.

CYCLO-COLUMN CELL

The cyclo-column cell is another pneumatic flotation machine that utilizes centrifugal forces. As illustrated in Figure 37, the principal element is a circular, centrifugal column inside which a centrifugal force field is generated by pumping a preaerated flotation feed tangentially into its lower end. The column is closed at the bottom and open at the top. As a result of the tangential entry, the feed swirls inside the column as it moves upward. During the process, bubble–particle attachment takes place and the resulting bubble–particle aggregates collect around the central axis of the column. Inside the centrifugal column, there is a second column, called the bubble column, which captures the bubble–particle aggregates and transports them into the froth column above. The froth column is wider, enabling entrained gangue to drain from the froth,



Source: Yalcin 1995
Figure 37 Cyclo-column cell



Adapted from Urizar 1996

Figure 38 Urizar cascade flotation machine

Courtesy of Amerigo Resources Ltd.

Figure 39 Cascade flotation at El Teniente

resulting in a clean concentrate. The material that does not enter the bubble column flows into a pulp collector and reports as tailing (Yalcin 1995).

CASCADE FLOTATION MACHINES

Cascade flotation machines historically have been simple and inexpensive. A modern design is shown in the development of Urizar (1996), where air is forced through a porous media into a thin film of flowing pulp (Figure 38).

Although rarely used in the modern flotation era, the size of the cascade flotation plant treating tailing from Codelco's El Teniente copper concentrators warrants mention. In 1992, Minera Valle Central S.A. started to recover copper from the fines fraction of El Teniente tailings using a simple cascade flotation system and a flotation cleaning circuit. Subsequent expansions introduced additional cascades, mechanical roughing cells, grinding and regrinding circuits, and flotation columns for cleaning. By 2003, the cascade flotation circuit had a capacity of 100,000 t/d and consisted of four lines of concrete cascade flotation cells, each 1.80 m × 0.70 m × 54 m in the first stage, and four lines in the second stage, each 1.80 m × 0.70 m × 126 m, providing 148 cascades in total (Figure 39). The cascade circuit reportedly produces approximately 4,500 t/d of low-grade concentrate at about 0.42% Cu. This concentrate is pumped directly to the rougher flotation (Maycock 2003).

PIPE FLOTATION

In 1996, Syncrude operated a pilot plant to evaluate the potential for addition of reagents and air to pipelines to allow initial bubble–droplet contacting. The pilot study showed that the bitumen recovery target of 95% was exceeded. The froth quality achieved was 59% bitumen and 20% solids (Mankowski et al. 1999). The new process operated at a temperature of 25°C, offering lower energy costs and lower capital costs. The arrangement is described by U.S. Patent 0249431 (Cymerman et al. 2006a).

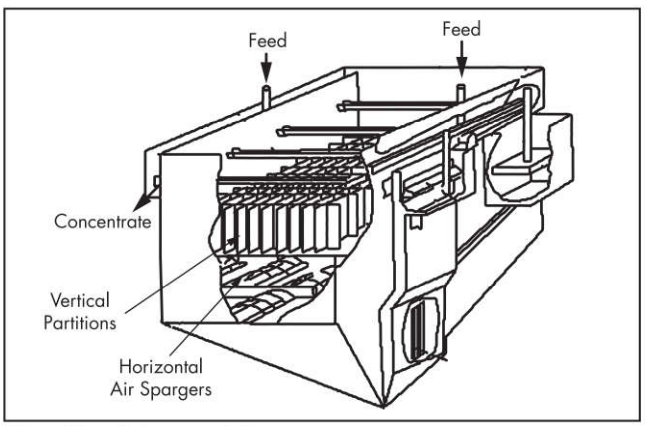
The system was implemented in Syncrude's Aurora operations, which commenced in 2000. Oil sand slurry was transported from the mine site via pipelines (3–5 km) with a relatively high slurry velocity (4 m/s). Air was injected into the pipelines. The aerated oil sand slurry was then introduced into primary separation vessels through a tangential entry feed well. Bitumen droplets attached to air bubbles to form a primary froth for recovery. Wash water was added under the froth to reduce the amount of solids reporting to the bitumen froth. The implementation was partially successful with a bitumen recovery of 83% versus the target of 95% (Cymerman et al. 2006b).

RUBINSTEIN FPPM

Rubinstein's FPPM pneumatic cell was a Russian development in the 1970s (Rubinstein 2003). As shown in Figure 40, horizontal air spargers provided aeration at the base of the tank. The collection zone was divided into compartments of 0.04 m^2 cross-section by vertical partitions, allowing optimization of the slurry and airflow patterns in the cell. Two 40-m^3 cells (2.2 m × 4.4 m cross section by 6.0 m height) were installed in the Kuznetsk coal basin. Each cell had a unit capacity of approximately 400 m^3/h.

DE-INKING PNEUMATIC FLOTATION CELLS

Froth flotation is the most widely used separation process in modern paper mills, and the development of flotation de-inking cells has been pursued more aggressively than the technologies of any other segment of the pulp and paper industry. Flotation is traditionally the standard European de-inking system for old newspapers and is gaining increased use in other areas. The earliest work on utilization of froth flotation for de-inking of waste papers dates to the 1930s with the flotation de-inking patent by Hines (1933). The first commercial flotation de-inking system was installed in the United States in 1952; Europe installed its first one in 1959. As discussed by Jiang and Ma (2000), flotation de-inking has many similarities to mineral flotation with performance affected by factors such as pH, consistency, temperature, ink/fiber particle size, chemical types, water hardness, and air bubble size. Several types of de-inking pneumatic flotation machines are of interest to the minerals industry and are discussed in the sections that follow.



Adapted from Rubinstein 2003
Figure 40 Rubinstein FPPM pneumatic cell

Swemac Hellberg Cell

The Swemac Hellberg cell aerator provides independent control of bubble size and the air-to-liquid ratio, allowing improved selectivity. It features an annular flotation cell into which the liquid to be treated is introduced tangentially. As the liquid whirls around the cell, foam is blown radially inwardly by an airstream into a central region in which continued whirling of the foam facilitates its separation from the air. The foam drops into a recovery zone while the air is drawn off and recirculated (Hellberg 1980).

Escher Wyss Cells

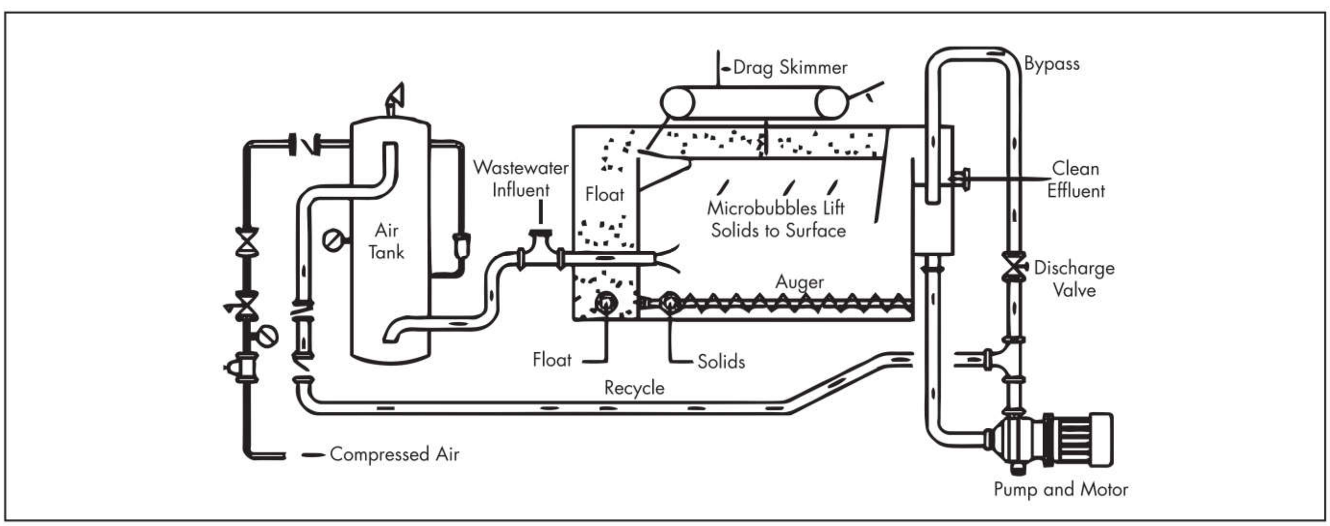
The Escher Wyss FZ-U cell made use of a rectangular tank with an air distributor located at its base. Compressed air was forced through apertures in the distributor to form bubbles. One advantage of the cell was its ability to be stacked vertically (Holik and Mueller 1975). The CF (compact flotation) cell was designed for the removal of small ink particles, and the CFS (compact flotation–speck removal) cell was designed for the removal of large ink particles. Both cells are round with tangential stock feed through multiple lines, and foam removed by gravity overflow. The primary differences between these cells are in the aeration sections and in the tank construction. The CFC introduced in 1992 was a modification of the CF and the CFS cells. It is a totally enclosed modular design and may be installed in either a single cell or a stacked two-cell arrangement (Sauvé 1999).

Tubular Injector Cell, Elliptical Cell, and EcoCell

The tubular injector cell utilized a vertical venturi-type aerator with flotation conducted in a closed horizontal, cylindrical vessel. The tubular shape was subsequently compressed from cylindrical to elliptical to promote faster air removal. It was further refined into the EcoCell following replacement of the venturi with the Escher Wyss CFC design.

Shinhama Hi-Flo Flotator and Kamyr Gas-Sparged Cyclone

Both the Shinhama Hi-Flo flotator and the Kamyr gas-sparged cyclone (GSC) make successful use of the ASH for de-inking



Source: Shammas and Bennett 2010
Figure 41 Dissolved air flotation arrangement

flotation. Where the GSC is a direct development of the ASH, the Hi-Flo is larger and does not use a porous medium for bubble generation.

DISSOLVED AIR FLOTATION

Dissolved air flotation is widely used to treat wastes from a wide variety of sources including paper making, refineries, ship's bilge and ballast waste, de-inking operations, metal plating, meat processing, laundries, iron and steel plants, soap manufacturing, chemical processing and manufacturing plants, barrel and drum cleaning, wash rack and equipment maintenance, glass plants, soybean processing, mill waste, and aluminum forming. The air is released from a supersaturated solution because of the reduction of pressure. As a result of this pressurization–depressurization, very small gas bubbles are formed and rise to the surface with oil and suspended solids attached. The amount of gas dissolved in solution and consequently the amount of gas released upon reduction of the pressure are both direct functions of the initial air pressure. Following pressurization, the water proceeds from the saturator, through a pressure-reducing valve, into the flotation basin; there the bubbles will first nucleate on any available low-energy sites on solid particles. If no sites are available, bubbles will nucleate homogeneously in the liquid phase. The bubbles will then grow until their growth is diffusion limited. Bubble sizes typically range from 45 to 115 µm, depending on operating conditions (Shammas and Bennett 2010). A typical dissolved air flotation arrangement is shown in Figure 41.

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