The design, erection and operation of a Purlex plant at Buffelsfontein Gold Mining Company, Ltd.

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SYNOPSIS

As a result of a test programme over a period of approximately ten years and a decision to extend the uranium plant, it was decided to design and erect the first full-scale Purlex Plant in the Republic of South Africa.

The plant at Buffelsfontein G.M. Co. Ltd. was commissioned in August, 1968, and to date has produced results superior to those predicted in the original estimates.

The capital expenditure was R678 000 for a plant capable of treating 1 600 gallons per minute of pregnant solution.

Compared with the conventional resin ion exchange process, it has been found that the saving in production costs is 21 cents per pound U_3O_8 produced.

The amine loss is 10 parts per million, which is equivalent to a total organic loss of 250 parts per million.

The uranium loss in the raffinate is less than 0.001 g/l U_3O_8.

The final product is of a high quality and it is probable that it is pure enough to convert to the fluoride directly.

HISTORY

The history of amine solvent extraction for uranium goes back to 1952 when the Americans issued a report from Oak Ridge in which the process was described.

More reports were issued by workers at Oak Ridge and in 1957 the Research Department of Stilfontein G.M. Co. Ltd. decided to investigate the process as it might be applied to South African resin column eluates.

In 1959 studies were initiated at Cape Town University to investigate uranium extraction using amines.

In 1962 discussions with representatives of the Government Metallurgical Laboratory were held with the purpose of starting a joint project on solvent extraction.

Consequently, a Bufflex Pilot Plant was started at Buffelsfontein G.M. Co. Ltd. in 1963.

The Bufflex process comprised the treatment of the pregnant solution in a resin ion exchange plant followed by solvent extraction of the concentrated eluate.

From this date onwards, solvent extraction of uranium was investigated jointly by the G.M.L. (later the National Institute for Metallurgy) and Buffelsfontein G.M. Co. Ltd.[1,2]

It became evident that solvent extraction of uranium in the Republic of South Africa held so much promise that it was felt that the industry as a whole should participate. Consequently, a second pilot plant, employing a Purlex process, was erected at Buffelsfontein by the Chamber of Mines in 1966. The Purlex process comprised direct treatment of the pregnant solution in a solvent extraction plant. From then on a Steering Committee, consisting of members of the uranium industry, directed the operations of the pilot plant.

This pilot plant continued investigations until 1968 when it was closed down and moth-balled.

DESCRIPTION OF THE PURLEX PROCESS

The Purlex Process

The process can be defined as the almost quantitative recovery of uranium in the form of a high purity ammonium-diuranate, from aqueous pregnant solution, utilising an organic solvent which is immiscible in water. The Flow-sheet is illustrated in Fig. 1.

The Solvent

The solvent at Buffelsfontein consists of three components, usually mixed in the proportions shown in Table I.

TABLE I

Component	% by volume	Specific Gravity
Tertiary amine	4	0.809
Aromatics	35	0.870
Lighting Paraffin	61	0.778
TOTAL	100	0.812

Extraction

The extraction is carried out in four stages with the two phases moving counter-currently. The aqueous phase gravitates while the organic phase is pumped. Extraction mixers operate in the organic continuous phase, which makes it necessary to re-cycle approximately 90 per cent of the solvent.

Scrub

The loaded solvent from the extraction bank is washed, in five stages, firstly with ten per cent sulphuric acid and then with an ammoniacal solution to remove impurities such as iron, arsenic and silica.

Strip

The uranium is stripped in four stages from the solvent with an ammonium sulphate solution, assisted by hydrolysis with 2N ammonium hydroxide in each stage.

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AQUEOUS TO M4

AQUEOUS (OK LIQUOR)

AQUEOUS (PREGNANT SOLUTION.)

SCRUBBED SOLVENT

LOADED SOLVENT

M4   P4   M5   NH₄OH M9   S9

BLEED TO GOLD AND URANIUM RECOVERY.

S4   S5   M6

P3   M3   S6   M10   NH₄OH   Na₂CO₃   Na OH

S3   M7   H₂SO₄   S10   M13

M2   P2   S7   NH₄OH   M11   S13

S2   M8   S11   AMM. SULPHATE

M12   NH₄OH

P1   M1   WATER NH₄OH   M8α   S12

S1   S8   STRIPPED SOLVENT

S8α

AQUEOUS (RAFFINATE)   FRESH SOLVENT

LEGEND:

M - MIXER.   AQUEOUS PHASE. ⟶

S - SETTLER.   SOLVENT PHASE. ----▸

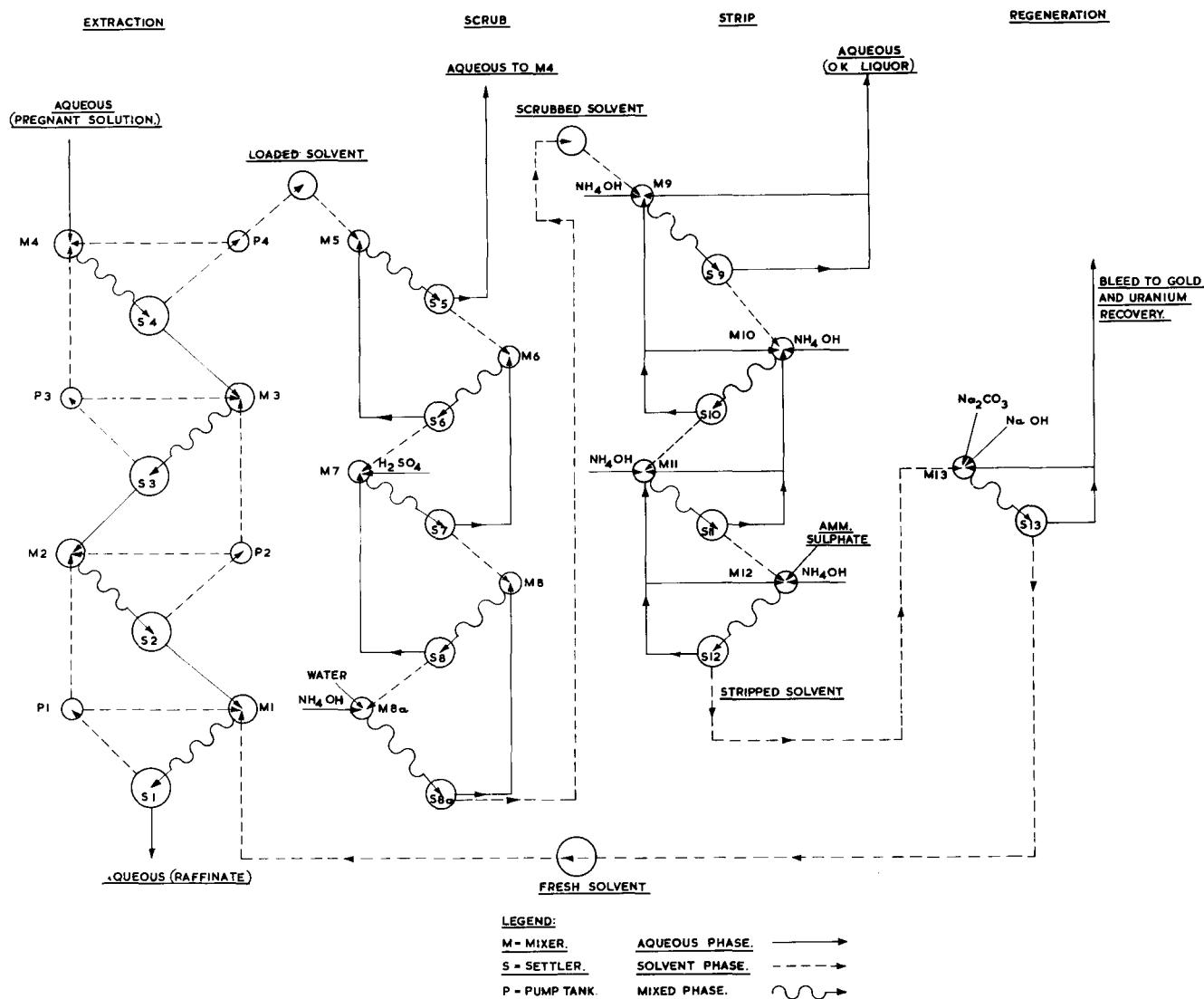P - PUMP TANK.   MIXED PHASE. ∿∿∿▸

Fig. 1

Ammonium-diuranate is precipitated with gaseous ammonia and settled out of the loaded ammonium sulphate solution. The "barren" ammonium sulphate solution is returned to the last strip mixer for re-use.

Regeneration

The stripped solvent is regenerated in one stage by the addition of sodium carbonate and caustic soda before being re-used in closed circuit.

ECONOMIC EVALUATION

The investigation of both BUFFLEX and PURLEX processes on a pilot plant scale enabled a fairly accurate economic evaluation to be made comparing the costs of uranium recovery by ion exchange, Bufflex or Purlex processes.

It should be noted that a decision had been made to extend the uranium plant at Buffelsfontein from its existing capacity of 150 000 tons per month to 250 000 tons per month. It was obviously vital that a clear cut picture should be drawn at the time to enable a decision to be made whether to increase the number of ion exchange columns or to convert the whole plant either to the Bufflex or the Purlex process.

The comparison of estimated operating costs is shown in Table II. It is clear from this comparison that the Purlex process appeared the most attractive. There was, of course, the calculated risk that some of the known problems, such as fungus growth and possibly other problems then unknown, might cause material difficulties in the operation of the full-scale plant. However, consideration of all the factors led to a decision to install a full-scale Purlex Plant.

DESIGN AND CONSTRUCTION

ERECTION OF THE FULL-SCALE PURLEX PLANT

General Layout

A site adjacent to the resin regeneration plant and the ion exchange eluate storages was selected. This allowed for the incorporation of redundant ion exchange equipment and the utilisation of existing precipitation facilities. In view of the generally favourable climate, an open air construction was decided on. This reduced the capital cost, eased the ventilation problem and diminished the fire hazard.

TABLE II

ECONOMIC EVALUATION OF THE ION EXCHANGE, THE BUFFLEX
AND PURLEX SYSTEMS

Cost in cents per lb. U_3O_8	Conventional Ion Exchange —Main Plant (b)	Bufflex Pilot Plant	Purlex Pilot Plant
Nitric Acid	13.12	—	—
Ammonia	2.90	2.41	2.78
Caustic Soda	0.85	0.13	0.33
A.S. Lime	1.15	—	—
Sulphuric Acid	0.33	0.49	0.09
Amine	—	0.37	2.55
Third phase inhibitor	—	0.06	1.83
Paraffin	—	0.32	1.39
Sodium carbonate	—	0.06	0.11
Hydrochloric Acid	—	0.04	0.07
Total Reagents	18.35	3.88	9.15
Electric Power	0.50	0.70	0.84
Maintenance stores (a)	0.77	1.00	1.15
Water	0.21	0.40	0.19
Resin replacement (a)	3.89	4.53	—
Labour (a)	3.71	4.01	2.42
Uranium loss in barren	9.24	9.24	2.09
TOTAL	36.67	23.76	15.84

Notes:

(a) These costs were all reconstructed, based on a monthly production of 130 000 lb. U_3O_8.

(b) Conventional ion exchange costs were based on actual performance during the five month period January to May, 1968. Uranium loss in the barren for the conventional ion exchange and the Bufflex systems were also based on actual losses during this period.

A control centre, which houses the instrument console, electrical switchgear and laboratory, overlooks the extraction sections. A concrete apron underlies the plant to contain spillage. Each mixer-settler unit is on a slightly lower elevation than the previous one on line to allow gravitation of one phase. The scrub, strip and regeneration sections are situated underneath the two extraction banks. This arrangement permits a more compact plant and facilitates supervision.

Design

The design was essentially based on parameters established on the pilot plant. Separate mixers and settlers are used on all stages. The mixers are used only for dispersion of the aqueous and organic phases and have no pumping action. The emulsion thus produced flows by gravity to a settler.

Extraction

Two extraction banks in parallel, each capable of treating 800 gpm of pregnant solution are used. This was preferred to one bank with a capacity of 1 600 gpm in view of the increased flexibility and also on account of the smaller scale-up factor from the pilot plant flow rate of 120 gpm

Extraction Settlers

The surface area of the settler was calculated on the basis of 1.0 gallon per minute of the dispersed (aqueous) phase per square foot of settler area. Allowing an adequate safety factor, a 35 foot diameter settler, equivalent to a surface area of 962 square feet was selected to treat a maximum of 800 gpm of pregnant solution.

An overall solution depth of 28 inches, comprising two phases of 14 inches each, was selected.

Circular settlers were installed for two reasons, namely:

(a) The initial enquiries indicated a prohibitive cost for stainless steel equipment and polyester fibreglass seemed the one likely material of construction. Fibre-glass vessels could not be manufactured in square or rectangular shapes without elaborate stiffening.

(b) Circular settlers allowed for the use of peripheral aqueous and organic overflow launders; the inner edges of the launders forming the overflow weirs. The large solution volumes involved made the long launders attractive as the shallow flow over the weir would minimise short-circuiting in the settler, would improve phase depth control and would reduce the volume of solution overflowing after power failures.

The height of the organic overflow launder is set at the total solution depth — 28 inches. The height of the aqueous overflow launder is set at 25 inches, but can be altered when required. Launders were equipped with P.V.C. lips to facilitate levelling off.

The emulsion flows in a launder from the mixer to the settler. A slight swirl is induced in the emulsion as it enters the settler to improve the distribution. The emulsion strikes a platform situated in the organic phase just above the interface level.

A radial punched plate baffle is located in the settler at a radius of 7 feet 6 inches to avoid short-circuiting of solutions and to contain the dispersion band.

The aqueous discharge from a settler flows into the next mixer while the organic phase is discharged into a solvent distribution tank.

The Raffinate After-Settlers

The raffinate from the last extraction settlers of both banks gravitates into a central raffinate collecting tank from where it is pumped to three 30 feet diameter by 10 feet high raffinate after-settlers operating in parallel.

Three redundant eluate storages, used in the ion exchange process, were lined with fibre-glass and modified to serve as after-settlers.

Raffinate is fed diagonally opposite to the overflow port. A portion of the entrained solvent separates from the aqueous raffinate and floats on top while the raffinate dips underneath a baffle and is discharged.

The solvent is reclaimed intermittently, utilising a launder, attached to the side of the tank. The solution level in the tank is elevated by throttling the delivery valve; the solvent overflows into the launder and gravitates to the spillage sump in the Purlex Plant.

Extraction Mixers

The residence time in the mixer is 30 seconds.

A mock-up mixer and settler, capable of handling 800 gpm of pregnant solution and identical in design to that envisaged for the ultimate plant, were installed ahead of schedule for testing purposes. Trouble was experienced in operating the mixer in organic continuous phase using the normal power input. Experiments indicated that proper distribution of the aqueous phase into the mixer was necessary when handling large volumes, and a peripheral distribution shelf was successfully used for

this purpose and was incorporated in the design of the final extraction mixers.

The mixer is a cylindrical tank equipped with the aqueous distribution shelf near the top with the outlet opening lower down to prevent flooding of the shelf and to ensure free-fall of aqueous from the shelf into the mixer. The outlet port is surrounded by a semi-circular baffle extending to about 12 inches from the bottom to prevent by-passing; the mixed phase passes underneath this baffle to the outlet port. A conductivity probe is situated in the baffled compartment and activates an alarm on the annunciator panel when flip-over to aqueous-continuous phase occurs.

The Impeller

A standard impeller, as used on the pilot plant, was used.[1],[2] A short impeller shaft is used and is situated above the interface level. This ensures that the impeller starts in the organic phase after a stoppage and facilitates immediate operation in organic continuous phase.

The Scrub Section

A five stage scrub was incorporated with a view to the production of high purity uranium at a later stage.

The loaded solvent from both extraction banks is delivered into a surge tank ahead of the scrub bank. The solvent gravitates into the first scrub mixer and then through the entire bank. The aqueous phase is pumped using centrifugal pumps. The aqueous effluent from the first scrub settler is returned into either or both of the last extraction mixers.

Ten per cent sulphuric acid is supplied to the third scrub mixer, while filtered domestic water and 2N ammonium hydroxide are supplied to the last mixer. The reagent feeds for scrubbing are equipped with ratio controllers, activated by the solvent flow to the first scrub mixer.

The Scrub, Strip and Regeneration Settlers

Settlers have been conservatively designed on the flow of the solvent and based on one gallon of solvent per minute per square foot of surface area. The installed 16 feet diameter settlers should, therefore, permit a maximum solvent flow of 200 g.p.m. The phase depths are identical to those in the extraction banks.

The emulsion from the mixers is delivered into the organic phase through the side of the settler. Three baffles distribute the phases through the settler in a zig-zag pattern. The aqueous pump tank is incorporated in the settler. The aqueous transfer pump is mounted against the settler, drawing directly from the aqueous compartment. The settlers are equipped with two view panels on the sides. Two crud drains per settler have been allowed, one from the first baffled compartment and one from the last. The ports are at the interface level through the sides of the settlers.

The Scrub, Strip and Regeneration Mixers

The retention time is five minutes. The power consumption is five horse-power per 1 000 gal. of mixer volume, the impeller details being calculated on the same principle as for the extraction mixers.

The geometrical design of extraction, scrub, strip and regeneration mixers is practically the same. The latter three are not equipped with aqueous distribution shelves in view of the smaller aqueous volume handled. Furthermore, those mixers intended to operate in an aqueous

continuous phase have longer impeller shafts so that on starting up the impeller will be in the aqueous phase.

The Strip Section

Solvent from the last scrubbing stage is pumped to a surge tank ahead of the strip section whence it gravitates through the strip and regeneration sections.

Ammonium sulphate is delivered to the last strip mixer at the rate of one-third of the solvent flow and is pumped counter-currently to the solvent flow. The pump deliveries were equipped with by-pass connections back to their own mixers to facilitate operating of any mixer in aqueous continuous phase when necessary.

2N ammonium hydroxide solution is fed automatically to all strip mixers to effect a gradual stage-wise pH increment from 3.8 in the first mixer to 5.5 in the last mixer.

The pH electrodes were initially situated in the aqueous phase in the settlers and the mixers were operated organic continuously. The long time lag, however, caused cycling and the electrodes had to be moved into the mixers. The pumps were incapable of transferring the aqueous phase and, in addition, circulating an adequate volume to keep the mixers in an aqueous continuous phase. Air-lifts were therefore installed to circulate the aqueous phase. The feed to the air-lifts was taken slightly below the interface level, from the first compartment of the settler. The air-lifts, apart from circulating adequate volumes of the aqueous phase, also return accumulated crud into the mixers, thereby assisting in its dispersion.

The loaded ammonium sulphate from the first strip settler (termed O.K. liquor) is transferred to an after-settler and from there to a storage. Facilities are provided to reclaim solvent from the after-settler.

O.K. liquor is heated to 30°C by an automatically controlled steam heater in a vessel ahead of the single five feet by five feet precipitation agitator, where the uranium is precipitated with gaseous ammonia at pH 7.3. The effluent from the five feet by five feet precipitator is settled in a 30 feet by 12 feet thickener. The overflow gravitates into the ammonium sulphate storage and the underflow is filtered on two six feet by six feet rotary filters in series. A de-ionised water wash is applied to both.

The precipitated uranium slurry settles and filters better than the conventional slurry. The required settling area is approximately one square foot per 10 gallons of liquor per hour, while the filter duty of a six feet by six feet rotary filter is about 7 800 pounds U_3O_8 per 24 hours at a feed S.G. of 1.250.

Regeneration

Regeneration is performed in one mixer-settler unit. A centrifugal pump circulates the aqueous phase and another transfers the regenerated solvent to the elevated solvent storage.

Fresh six per cent sodium carbonate solution is fed continuously from a steady-head through a rotameter into the mixer at the rate of 0.25 gallons per minute. The pH is automatically controlled at ±8.5 by the addition of 10 per cent caustic soda solution. The mixer operates in an organic continuous phase and the pH electrodes, therefore, are in the settler.

Excess aqueous solution, equivalent to the addition of fresh sodium carbonate and caustic soda solution

overflows from the aqueous weir compartment into a four feet diameter by four feet nine inches high fibreglass vessel. A 10 per cent sulphuric acid solution is added to pH 2.5. The acidified solution is circulated with an airlift to effect proper agitation and is then transferred into the Gold Plant leaching circuit. In this manner small amounts of gold and uranium are recovered.

Crud Treatment Facilities

Two six feet diameter by eight feet high fibre-glass crud tanks have been provided. Both tanks were equipped with perspex view strips and a common pump suction manifold with take-off points spaced at 18 inch intervals to facilitate reclamation of solvent. A pump was provided for circulation and reclamation.

Instrumentation

A centralised system of control instruments was installed. All functions are relayed to a console, placed in an air-conditioned control room. An annunciator panel reveals alarm conditions of flow, tank levels, conductivity, pH and running equipment. The instruments are of the pneumatic type and a refrigerator unit is used to clean the mine air prior to being used.

The following solution flows are controlled, measured and recorded on charts by differential pressure type instruments:—

(a) Pregnant solution.
(b) Solvent feed to extraction.
(c) Solvent feed to scrub.
(d) Solvent feed to strip.
(e) Dilute sulphuric acid, dilute ammonia and domestic water to scrub.
(f) Ammonium sulphate to strip.

The scrub solutions are automatically controlled in ratio to the solvent feed to scrub. This ratio may be changed, within certain limits, to suit plant conditions.

Conductivity indicators were placed in all mixers of all stages.

Automatic recorder-controller meters control the pH in the strip and regeneration sections.

Level indicators on the console indicate the solution levels in the pregnant solution storage, solvent storage, solvent surge tanks, O.K. liquor stock tank and in all reagent solution stock tanks.

The make-up of 2N ammonium hydroxide solution was automated. A level controller regulates the feed of demineralised water to the make-up tank. The water addition is measured and a computed volume of ammonia gas is added automatically. Temperature and pressure compensators allow for variation in temperature and pressure of the ammonia gas. A steam operated heat exchanger, placed close to the ammonia feed point, ensures complete vaporising of ammonia liquor. A flexible ammonia hose between the ammonia gas injection point and the control instruments eliminates vibration on the instruments.

Reagent Storage and Handling Facilities

Amine is received and stored in 45 gallon drums.

Paraffin is delivered in bulk by road transport and stored in an 18 000 gallon storage tank. Batches of paraffin are pumped and metered into the regeneration mixer as required.

Aromatic liquid is delivered in bulk by rail tanker into an 18 000 gallon storage tank, from where it is transferred and metered into the regeneration mixer when required.

Solvent is made up in small batches to avoid a sudden variation in the relative concentration of one component while a batch of this is being fed into the circuit.

The four 16 feet diameter reagent storage tanks originally used for resin regeneration were converted into reagent solution make-up and storage tanks. The Goldfields type lime agitator was converted to a sodium carbonate solution make-up and storage tank.

WATER TREATMENT

Demineralised Water Plant

The demineraliser consists of two stages, viz a split bed and a mixed bed ion exchanger. Domestic water first passes through the split bed system, comprising a column with 250 cubic feet of cation exchange resin and a second column with 200 cubic feet of anion exchange resin, then through a 15 cubic feet mixed bed column to remove the last traces of impurities, mainly silica.

A nine feet diameter resin pressure column, a seven feet diameter resin pressure column and a 14 feet by 18 feet conical resin storage vessel, previously used for resin regeneration, were converted to a cation column, an anion column and a demineralised water storage, respectively.

The effluent column from the mixed bed demineraliser is equipped with a conductivity meter to detect electrolytes.

Filtered Scrub Water

In order to remove suspended impurities, such as rust, from domestic water, prior to use as scrubbing water, it is filtered through a sand bed.

A nine feet diameter obsolete resin regeneration column was converted into a sand filter and a 14 feet by 18 feet obsolete conical resin storage into a filtered domestic water storage.

Fire Precautions

In assessing the fire hazard, the inflammable and explosive nature of the solvent should be considered. Paraffin has the lowest flashpoint at $\pm110°$F, which in consequence, is also the flashpoint of the solvent mixture.

It is accepted that suitable fire fighting equipment should be installed in situ, as the limited time factor in the case of oil fires prevents the effective use of outside facilities. An automatically operating foam system and a series of water hydrants were installed at Buffelsfontein. A water curtain is provided for the front part of the control room to protect the operator and the instruments. The water curtain is activated by fixed temperature nozzles.

Platforms and walk-ways were fabricated from steel egg-grating to assist natural ventilation and to prevent the accumulation of solvent spillage. Fibre-glass containers were equipped with stainless steel lids.

The Buffelsfontein Plant, in view of its open air construction, is equipped with conventional T.E.F.C. motors with flame-proof stop-start stations. The spillage pump, situated directly above the spillage sump is equipped with a flame-proof motor. The switchgear is located in a pressurised room, the fan intake being

away from the plant. The control room is pressurised from the same source and the temperature is controlled. The plant is illuminated with conventional sealed flood lights mounted well clear of the plant. Four lightning arrestors have been installed; one on each corner of the plant.

Anti-static Precautions

All constructions are bonded and earthed, especially those of fibre-glass construction or with non-metal linings. Stainless steel earthing strips are strung through fibre-glass transfer pipe lines, but these suffer from the disadvantage that the flow through small-bore pipes is hindered. The ends of delivery lines dip into the solution in the receiving vessel to prevent free fall.

It was established that the introduction of ions into the solvent during the process increases the conductivity of the solvent and reduces the danger of static feed-back. This naturally does not apply to the fresh solvent, especially the unmixed paraffin, and regulations regarding earthing and maximum permissible pipe velocities were observed.

Synthetic clothing was seen as a possible source of static electricity and its use is prohibited. Personnel are supplied with denim overalls.

Situation of the solvent plant in relation to other structures

The proximity of the ammonia storage vessels on the Buffelsfontein plant was considered a hazard. A 10 feet high brick wall was erected directly between the storages and the Purlex Plant to act as a heat shield and a three feet high brick wall was extended around the storages to prevent burning solvent from entering the enclosure. The area surrounding the plant is kept clean and proper fire breaks are being maintained.

Security

A fence has been erected around the plant and admittance is controlled. Suitable notice boards, warning personnel and visitors against the fire hazard, have been erected. Smoking and the use of open flames in the security area are prohibited. The use of electrical and friction tools, which may cause sparks, is controlled by the use of a permit system. Plant personnel receive regular training in fire prevention and fire-fighting.

MATERIALS OF CONSTRUCTION

316 Stainless Steel

Apart from the cost, this is the most suitable material for vessels, pipes and valves. However, it is attacked on the welds by dilute sulphuric acid solution owing to the formation of non-resistant carbides. This problem was overcome by using a low carbon type, viz: 316L. The extraction settlers and the covers for the fibre-glass settlers were fabricated from 316L.

Polyester fibre-glass

This material is suitable for vessels and piping and showed no corrosion. Fibre-glass lining of mild steel vessels proved very successful. The scrub, strip and regeneration settlers, the scrub organic surge tank, the strip organic surge tank and all flow measuring boxes were fabricated from fibre-glass. Virtually all pipe runs in the plant, varying in size from two inches diameter to 12 inches diameter, are of fibre-glass. Standard flanges, imbedded in fibre-glass, were used at joints. All the

mixers, the pump tanks and the solvent storage tanks were fabricated out of mild steel lined with fibre-glass.

The O.K. liquor after-settler and storage, the uranium precipitator and thickener, the ammonium sulphate stock tank and the three raffinate after-settlers were mild steel rubber-lined vessels originally in the conventional plant. These have been lined with fibre-glass. In two cases, the fibre-glass lining was successfully applied on top of the old rubber lining.

Rubber Linings

Various types of rubbers were tested without success. The solvent normally penetrated the rubber and destroyed the bond between the lining and the metal. Nitryl rubber pump components are being used successfully on solvent and acid applications. It is also used for gaskets and valve diaphragms. Neoprene is subject to swelling and distortion.

Polyvinyl chloride

Polyvinyl chloride hosing has been used successfully on the pilot plants, but is attacked by solvents with a high aromatic content.

Acrylonitrile Butadiene Styrene

This material was used successfully on the pilot plants and is not affected by aromatic compounds. These pipes suffer from the disadvantage that extensive support is required, preferably in angle or channel iron.

Polyurethane, polypropylene, penton and prodorfilm linings

These had been successfully tested on the pilot plants, but found no application in the main plant, except for polyurethane base paints.

Nebar

Nebar gaskets were used with good effect in the mock-up and in the construction of the main plant.

PUMPS

Kestner vertical type pumps are used for most applications. These pumps are glandless, self-priming and resistant to chemical attack. Surging of a Kestner pump may be avoided by extending the suction pipe upwards to just below the priming level.

Fibre-glass airlifts are used successfully to circulate the aqueous phase in the strip bank.

Paints

Epoxy and polyurethane base paints offer good chemical resistance, but the hard surface chips readily on sharp edges and corrosion sets in between the film and the metal. Polyurethane paint was applied to all steel-work and to the concrete apron of the plant. Chipping and blistering resulted in numerous failures.

Enamel paints were moderately successful on the pilot plants and, considering costs, must be regarded as the most suitable paint available.

Capital Expenditure

The capital expenditure on the plant, which included alterations to the old plant and modifications to the precipitation section was R678 000. This amount included expenses during the mock-up experiments and enough solvent to start the plant.

Plant Control

The two separate extraction banks provide useful flexibility and the plant may be operated over a wide range of pregnant solution flow rates, viz. from 400 gpm to 1 600 gpm, without impairing the efficiency significantly. One bank is stopped in the event of the pregnant solution flow rate dropping below 800 gpm. This also facilitates maintenance on the extraction bank without stopping the entire plant.

Stopping and starting of the plant offer no problem and may be effected within a few minutes. Power failures, however, result in the overflow of a considerable volume of solvent on to the spillage apron.

The efficiency of the extraction process largely depends on maintaining the correct ratio between the flow rates of pregnant solution and fresh solvent. This ratio is influenced by the concentration of anions in the pregnant solution, the concentration of the amine in the solvent mixture and by the efficiency of the solvent regeneration process preceding the extration stage. The ratio is, therefore, not constant and requires periodic adjustment. Ideally the flow ratio should yield a barren raffinate (nil U_3O_8) and a saturated loaded solvent (1 g/l U_3O_8 per one per cent amine contained in the solvent). In practice an acceptable compromise is sought and it was found that the uranium content of the solvent leaving No. 1 extraction settler gives the best guide for solvent flow control. The optimum value varies from system to system and is best obtained by trial and error.

The importance of optimum solvent loading is often over-emphasised. Operating below the optimum solvent loading only results in a slight increase in sulphuric acid and ammonia consumption, while consistently high uranium extraction more than compensates for the increase in reagent cost.

Consistent solvent flow rates through the scrub and strip sections are essential for efficient operation; this is facilitated by the solvent surge tanks between the extraction and scrub and between the scrub and strip stages. The constant flowrate facilitates accurate scrub reagent addition and pH control in the strip and regeneration sections.

The flows of domestic water and dilute ammonium hydroxide to the fifth scrub stage are equivalent to 20 per cent and one per cent of the solvent flow, respectively while the flow of 10 per cent sulphuric acid to the third scrub stage is controlled at two per cent of the solvent flow. Scrubbing technique is described in greater detail under the purity of the final product.

The flow rate of the strip solution is not critical and it is kept constant under all conditions to provide a constant flow of liquor to the uranium precipitator. This flow rate is equivalent to about one third of the solvent flow.

Plant Tests

Certain tests, as given in Table III are carried out by the operating personnel.

Operating Costs

Actual working costs are given in Tables IV and V.

OPERATING TECHNIQUES

OPERATIONAL AND METALLURGICAL PROBLEMS ENCOUNTERED

General

The commissioning of the Purlex Plant was uneventful and apart from anticipated difficulties, which had also been encountered on the pilot plant, viz. fungal growth, periodic crud formation and occasional high solvent losses, the plant soon settled down to efficient operation.

However, consistently severe crud formation and abnormally high solvent losses were experienced during the period July to October, 1969. This coincided, probably incidentally, with the period while aromatic solution was introduced as a third phase inhibitor in place of the isodecanol. The mixture of isodecanol and aromatics was suspected, but pilot plant tests revealed that the low acidity of the pregnant solution—resulting from the re-circulation of primary filtrate to the leach—was mainly responsible for the poor results. Inadequate clarification of the pregnant solution aggravated the solvent loss. This was rectified by spiking the pregnant solution with sulphuric acid and by improved clarification.

Extraction was hampered by obstructions in the two inch diameter fibre-glass solvent transfer lines. These were replaced with three inch diameter lines.

Fungus

The species in the Purlex Plant was identified as belonging to the family "Aspergillus" and is not of the same species normally encountered in uranium leaching circuits. This discounts the theory that fungus originates from animal glue used as a filter aid in the filtration sections. The fungus is believed to be airborne and multiplies rapidly in oxygen-bearing media between a temperature of 25°C and 45°C. It has been established that isodecanol is a nutrient for fungus while aromatic solutions, although not a fungicide, do not support growth.

It is not clear why the Buffelsfontein plant should be plagued with fungus, while other solvent extraction plants are virtually free of fungus. It has been suggested that the adjacent sewage works may be a breeding spot for fungus.

The presence of fungus is undesirable in all sections of the plant. It grows in both phases and attaches itself to containers. Accumulations on settler weirs affect the organic and aqueous phase depths. It is normally saturated with solvent and, when released into the aqueous phase, carries substantial quantities of solvent with it. Fungus generally has the same specific gravity as solvent and is, therefore, difficult to separate from solvent.

In an attempt to eliminate or reduce the fungus problem, isodecanol addition to the circuit was replaced by an aromatic solution in April, 1969. Owing to the large volume of solvent in the circuit, bleeding of the isodecanol is a lengthy process and the immediate effects were, therefore, slight. The isodecanol concentration is now down to about 0.1 per cent and the fungal growth during the hot summer period was much less compared with the previous summer.

Crud

Attempts to analyse crud have been unsuccessful as the residue disintegrated upon drying. Recent work, however, indicated that sulphur may be a major constituent of crud in the strip section.

TABLE III

PLANT TESTS

Test	Purpose	Method
U_3O_8 analyses on solvent leaving No. 1 settler.	To control fresh solvent flowrate to extraction.	Colorimetric method using PAN reagent.
Third phase test on regenerated solvent.	To detect a deficiency of third phase inhibitor.	Nitric acid test.
Test for silica on effluent from water demineraliser.	To detect breakthrough on the demineraliser and prevent contamination of demineralised water supply.	Ammonium molybdate test.
Sulphuric acid content of pregnant solution	To control above 2.5 g/l to avoid high solvent loss	Titration with standard sodium carbonate solution
Phase ratios of emulsions in extraction and strip mixers.	To control flowrates of solvent and aqueous circulation in extraction and strip section.	Allow emulsion to separate in measuring cylinder.
Conductivity check test of emulsion in mixers.	To ascertain whether mixer operates in organic or aqueous continuous phase.	Portable conductivity probe.
pH checks on aqueous solution in each strip stage, the regeneration stage and final precipitation stage	To check accuracy of automatic pH controllers. Inaccurate pH control may cause local A.D.U. precipitation	Tabel model Knick pH meter.
Flowrates of pregnant solution, solvent to extraction, solvent to strip, reagents to scrub, ammonium sulphate to strip and sodium carbonate to regeneration.	To check automatic flow controllers—thus improving plant efficiency and obtaining an accurate metallurgical balance.	Measuring boxes and stop watch.
Flowrate of O.K. liquor to precipitation	To control the residence time in the precipitator — this affects the physical properties of the A.D.U.	Weir measurement.
Specific gravity of A.D.U. slurry to primary and secondary rotary filter.	To control filtration efficiency.	
Concentration of scrub acid.	Scrub reagent control.	Titration.
Concentration of ammonium hydroxide.	Scrub reagent control and control of ammonium sulphate concentration.	Titration.
Concentration of sodium carbonate.	Regeneration reagent control.	Titration.

PLANT SAMPLING

Sample	Analyses	Purpose
Pregnant solution.	U_3O_8, Fe^{++}, Fe^{+++}, H_2SO_4, Mn, S_4O_6, CNS, NO_3, SiO_2, An.CO. and suspended solids.	Uranium balance, acid control and possible effect of impurities on the process and the quality of the product.
Raffinate	U_3O_8	Uranium balance and to assess extraction efficiency.
Raffinate	Amine	To assess amine loss.
Solvent to scrub.	U_3O_8	To assess U_3O_8 loading.
Aqueous effluent from scrub bank.	U_3O_8 and Fe	To assess U_3O_8 and Fe removal in scrub bank.
Solvent to strip.	U_3O_8 and Fe	To assess the quantity of uranium stripped in the scrub section and the efficiency of iron removal. To calculate uranium balance in the organic circuit.
Solvent to regeneration.	U_3O_8	To assess strip efficiency and calculate uranium balance in the organic circuit.
Regenerant discarded.	U_3O_8 and Au	To calculate uranium and gold content of the regenerant.
Demineralised water.	Ca, Fe, Cl, SO_4 and SiO_2	To control operation of the demineraliser
O.K. Liquor	U_3O_8 and SO_4	To calculate U_3O_8 content of O.K. liquor and control sulphate content of the strip solution.
Regenerated solvent.	Amine and third phase inhibitor.	To control solvent make-up and assess losses.
Moisture of A.D.U. despatched	SO_4	To control filter wash and thus soluble sulphate content of A.D.U.
A.D.U. despatched.	U_3O_8	To calculate uranium despatched provisionally. (NUFCOR results are finally accepted.)
A.D.U. slurry from secondary filter.	Fixed sulphate and soluble sulphate.	To control precipitation conditions and final filtration.

NOTES:

The raffinate, fed to the after-settlers and the effluent from the after-settlers are sampled to assess the efficiency of the after-settlers and the final amine losses.

Owing to solvent adhering to the sides of containers, the sample should not be transferred from one container to the next. A 1 000 cc sample is taken in a sample bottle which is slightly bigger than 1 000 cc. The bottle is stoppered with a ground glass stopper and submitted for assay in the same bottle.

TABLE IV

OPERATIONAL COSTS FOR THE YEAR OCTOBER, 1968 TO SEPTEMBER, 1969

Item	Unit cost cent	Cost—cent per lb. U_3O_8
Paraffin	24.50	2.293
Amine	67.55	2.757
Solvesso-150	50.70	0.686
Iso-decanol	16.34	0.553
Ammonia	6.073	3.191
Caustic soda	7.00	0.668
Sodium carbonate	3.40	0.219
Sulphuric acid	0.51	0.168
Hydrochloric acid	5.45	0.001
Total reagents		10.536
Water	0.01644	0.185
Electrical power	0.372	0.835
Labour		2.423
Maintenance stores		1.155
(a) Total production costs		15.134
(b) U_3O_8 loss in raffinate	8 U.S. dollars	2.273
Grand Total (a+b)		17.407

TABLE V

OPERATIONAL COSTS FOR THE FOUR MONTHS NOVEMBER, 1969 TO FEBRUARY, 1970

Item	Unit cost cent	Cost—cent per lb. U_3O_8
Paraffin	24.50	1.488
Amine	67.55	2.064
Solvesso-150	50.70	2.160
Iso-decanol	16.34	0.100
Ammonia	6.073	2.776
Caustic soda	7.00	0.582
Sodium carbonate	3.40	0.194
Sulphuric acid	0.51	0.085
Hydrochloric acid	5.45	0.002
Total reagents		9.451
Water	0.01644	0.182
Electrical power	0.372	0.782
Labour		2.328
Maintenance stores		1.617
(a) Total production costs		14.360
(b) U_3O_8 loss in raffinate	8 U.S. dollars	0.796
Grand Total (a+b)		15.156

The formation of crud is undesirable as it prevents efficient phase separation. It eventually causes high solvent losses, incomplete stripping, incomplete regeneration and, in severe cases, will curtail normal operation.

Experience has taught that the following factors promote the formation of crud:—

(a) Undissolved solids in plant solutions promote "slime-crud" formation in the extraction bank. The suspended solids in the pregnant solution should not exceed 20 ppm.

(b) The presence of third phase accelerates crud formation. The required concentration of third phase inhibitors may be determined, but should be maintained at levels well above the minimum requirements. Concentrations of either two per cent isodecanol or 35 per cent aromatics are recommended for use in conjunction with an amine concentration of five per cent. Higher amine concentrations require increased additions of third phase inhibitors.

(c) Erroneous pH control in the scrub and stripping stages can cause severe crud formation. Local precipitation of A.D.U. provides a nucleus for crud formation.

(d) The use of a sulphuric acid-ammonium hydroxide scrub apparently aggravates crud formation; this is eventually carried through to the strip bank.

(e) The intensity of the leach appears to influence crud formation. This has not been proved conclusively, but there are indications that leach liquors, resulting from intense leaching, may be more susceptible to crud formation. The relatively high dissolved silica and polythionate content of these liquors may, upon ageing and cooling, cause precipitates, forming a nucleus for crud formation.

CRUD AND FUNGUS TREATMENT

Removal from Plant Containers

Plant settlers are cleaned out on a routine basis. The interval between clean-ups depends on the rate of fungus and crud formation. The only effective method of removal at Buffelsfontein is complete draining of tanks followed by sweeping and brushing of the settlers.

Treatment

A treatment tank should be available, preferably big enough to accommodate the entire contents of at least one extraction settler.

In view of the fact that the two original crud treatment tanks were too small, the O.K. liquor after-settler was converted to a crud treatment tank. The O.K. liquor storage tank was modified to serve as after-settler and storage.

The contents of an extraction settler or the entire strip bank may be transferred into the crud treatment tank and the plant may be recommissioned immediately on fresh solvent.

Uranium-bearing aqueous solution is recovered from the treatment tank and re-introduced into the extraction bank, while the solvent, crud and fungus are retained in the treatment tank for treatment before returning to the circuit.

Continuous vigorous handling and treatment with acidic solutions appear to be the most effective way of breaking up crud and fungus.

A centrifugal circulating pump provides vigorous agitation which assists in the destruction of deposits.

Once the organic fraction has been liberated, the specific gravity of the crud residue increases and it may be separated by settling.

Solvent Losses

The economic feasibility of the process is largely determined by the extraction efficiency and the level of the solvent losses. The losses of the various components of the organic phase are normally in direct proportion to the relative concentration of each in the organic mixture with the possible exception of aromatics. The amine loss at Buffelsfontein is of the order of 10 ppm.

FACTORS INFLUENCING SOLVENT LOSSES

Extraction Mixers and Settlers

The design of mixers and settlers should conform to the standards discussed under plant design. Mixers should provide intimate mixing of phases without producing stable emulsions. Operation of mixers in aqueous continuous phase results in inflated solvent losses. Pregnant solution flow rates should be kept constant and designed flow rates should not be exceeded. Air, drawn into the settler with the emulsion, forms bubbles to which solvent is attached. These air bubbles drift on the interface, burst when they strike the sides of the settler and transfer solvent into the aqueous phase.

There is a difference of opinion among operators as to whether the operation of intermediate extraction mixers and settlers has a pronounced effect on the final solvent loss or whether the solvent loss is largely determined by the operation of the final stage. A policy of operating all stages correctly has been adopted.

Special care should be exercised when refilling the final settler after draining. Failure to replenish the entire volume of the aqueous phase, before returning the organic phase, may result in some solvent getting into the aqueous weir compartment.

Stopping and Starting of the Plant

Stoppages must be limited since the solvent equilibrium in the circuit is disturbed when the plant is stopped as a result of the marked difference between the dynamic and static solvent inventories in the extraction settlers. It is, therefore, advisable to reduce the pregnant solution flow slowly before stopping the plant and to increase the flow slowly when restarting, allowing the plant to attain equilibrium without spilling solvent onto the apron. If possible, the organic circulation should be kept going as this will ensure an adequate supply of solvent to the mixer upon starting up and obviate the flip-over of mixers to aqueous continuous phase.

Solvent Composition

It was found advantageous to maintain the amine concentration as low as possible. The proportion of entrained amine lost is thus smaller and reduces the overall cost of the solvent.

A reduced amine concentration also increases the ratio between the concentration of third phase inhibitor and amine without increasing the first, resulting in reduced crud formation.

Composition and nature of the pregnant solution

The undissolved solids content should be kept down, preferably below 20 ppm.

The free acid concentration of the pregnant solution appears to affect the solvent losses. At Buffelsfontein it was found necessary to maintain the acidity above 2.5 gram per litre sulphuric acid.

The pregnant solution should be free from detergents which may cause the formation of stable emulsions. Effluents from sewage treatment works could act as a source of detergents. Soluble oils used for lubrication and detergents used in the plants for cleaning purposes should be controlled.

The temperature of the pregnant solution will not only affect evaporation losses during the process, but will also influence the rate of fungal growth. Ageing and cooling of the clarified pregnant solution may result in the formation of silica and sulphur precipitates, which again may promote crud formation.

Evaporation losses

High aqueous solution temperatures increase evaporation losses. Aeration of organic solutions should be avoided and containers should be equipped with suitable leak-proof covers. The ambient temperature affects evaporation. Although an open-air plant assists natural ventilation, protection against the sun would be advantageous.

Spillage

Spilling and handling losses may be minimised by providing a well-graded impervious apron underneath the entire solvent plant. Spillage should be returned into the circuit as soon as possible.

Fungus and Crud

The growth of fungus and the formation of crud must be kept to a minimum. Adequate facilities should be provided for the efficient treatment of fungus and crud and for the recovery of the solvent.

After Settlers and Coalescers

The incorporation of the three after-settlers at Buffelsfontein was advantageous. No attempt was made to distribute or disperse the incoming raffinate. A distribution baffle, subsequently installed in one after-settler, did not improve the operation significantly. Solvent recovery in the after-settlers was equivalent to 2 ppm of amine when the amine content of the feed from the extraction bank was of the order of 13 ppm and about 1 ppm when the feed contained 7 ppm of amine. The after-settlers are particularly useful when operational difficulties are encountered. Emulsion may be pumped into the settlers and the solvent reclaimed after an adequate settling period.

Tests were conducted with a small coalescer employing a pre-coated cloth. About 25 per cent of the amine in the raffinate leaving the after-settler could be recovered, while 25 per cent adhered to the precoat. The filter blinded within three hours and the throughput was of the order of 2.3 gallons per minute per square foot of filter area.

An experimental after-settler, packed with $1\frac{1}{2}$ inch raschig rings, reduced the amine content of the final effluent to 5 ppm regardless of the feed value. The volume of rings used was equivalent to one cubic foot per gallon of raffinate per minute.

Regeneration

The initial installation did not allow for the continuous feed of fresh sodium carbonate solution and the bleed of excess regenerant. The entire aqueous phase was discharged once per week and replaced with fresh 10 per cent sodium carbonate solution. The correct pH value

was maintained by manual addition of a 10 per cent caustic soda solution.

This system did not function satisfactorily as the carbonate solution was gradually displaced from the circuit by the caustic soda solution. The excess solution overflowed into the spillage sump where it mixed with other spillage before being returned into the extraction bank. The regenerant became badly discoloured and changed into a viscous syrup. The circulating rate of the regenerant was also reduced after stoppages or power failures on account of the fact that the static inventory of regenerant in the settler was less than the dynamic inventory, the balance having overflowed during the shutdown. Regeneration, strip and extraction efficiencies were impaired on account of the above weaknesses in the system.

The system was modified to facilitate a continuous small feed of fresh sodium carbonate solution and a continuous bleed of excess regenerant out of the circuit. The addition of caustic soda solution was automated. Tests were done to assess the optimum concentration of the sodium carbonate solution. A six per cent solution gave good results while a more dilute solution caused the formation of stable emulsions, possibly as a result of the presence of precipitated uranium.

THE RECOVERY OF GOLD AND URANIUM FROM DISCARDED REGENERANT

The discarded regenerant contains about 60 ounces of gold and 270 pounds of U_3O_8 per month. Recovery of this gold and uranium without contaminating the circuit with poisons, has been investigated.

It was found that the gold could be recovered quantitatively by zinc dust precipitation, but the uranium could not be re-precipitated by the direct addition of alkali. However, acidification with sulphuric acid to pH 2.5, followed by aeration, breaks down the carbonate complex and permits precipitation of about 80 per cent of the uranium content.

The acidified regenerant, after aeration, is transferred into the gold plant leaching circuit where the uranium is precipitated in the alkaline pulp while the gold remains in solution and is extracted in the filtrate.

PURITY OF THE FINAL PRODUCT

Ultra high purity of the final product offered no economic advantage. However, it was decided to attempt the production of a high purity A.D.U. for research and the plant was designed accordingly.

The contaminants in the final products produced in the pilot plants and in the main plant are illustrated in Table VI.

It is difficult to assess whether the final product is of nuclear quality owing to the lack of realistic specifications. The stringent specifications of certain countries cannot be satisfied, but the product was used successfully by some investigators.

Some of the impurities, which are absorbed more selectively than uranium, cannot be eliminated at present and their presence, no doubt, depends upon their concentration in the ore and on the rate of dissolution during the uranium leaching process. Molybdenum falls into this category.

Impurities with a lower affinity than uranium may be partially removed in the scrub section. Iron and arsenic fall into this category.

TABLE VI

ANALYSES OF CONTAMINANTS IN THE FINAL PRODUCTS OBTAINED IN THE BUFFLEX AND THE PURLEX SYSTEMS

Element	Bufflex ppm			Purlex Pilot Plant ppm			Purlex Main Plant ppm		
	High	Low	General	High	Low	General	High	Low	General
Al	13	L0.3	3.5	100	3.4	10	50	3	5
As	135	L10	23	14	L10	L10	30	9	20
B	1	L0.3	L1	3.4	L0.3	1	1	0.2	0.5
Be	L1	L0.3	L1	L0.3	L0.3	L0.3	L0.1	L0.1	L0.1
Bi	L1	L1	L1	L1	L1	L1	4	0.5	1
Cd	L1	L1	L1	L1	L1	L1	1	L0.3	1
Co	9.5	2	3.7	95	27	50	230	52	110
Cr	L1	L3	L3	3	L1	2	10	2	5
Cu	1	L0.3	L1	5	L0.3	2	15	0.1	2
Fe	30	3	4.7	300	32	75	500	200	300
Ga	2	L1	L1	L1	L1	L1	L1	L0.3	0.3
Ge	L1	L1	L1	L1	L1	L1	L3	L1	1
In	L1	L1	L1	L1	L1	L1	L3	L1	1
Mg	2	L1	2	5	1	3	5	1	3
Mn	L1	L0.3	L1	6.3	L1	3	5	L1	2
Mo	16	1.3	11	33	6	15	275	27	100
Na	L30	L10	L10	L30	L30	L30	L30	L10	L30
Ni	L1	L1	L1	L1	L1	L1	L3	L1	L1
Pb	5	L0.3	L1	L1	L1	L1	L3	L1	L3
Sb	L10	L3	L3	L3	L3	L3	L10	L1	L10
Si	200	2	25	130	18	40	500	17	100
Sn	3	L1	L1	L1	L1	L1	L1	0.1	L1
Ti	10	L1	L3	5	L1	L1	3.1	0.3	1
V	L1	L1	L1	L1	L1	L1	1.3	0.2	0.3
W	L10	L10	L10	L10	L10	L10	3	0.2	0.5
Zn	L30	L10	L10	L30	L10	15	L30	L1	L10
Zr	170	19	70	77	17	35	540	70	400

L = less than.

Silica contamination, which causes the formation of crud, is limited by the use of demineralised water for ammonium hydroxide make-up and as a filter wash.

Cobalt contamination results from the carry-over of anionic cyanide complexes into the pregnant solution. No attempt was made to eliminate cobalt in the Purlex section, but pilot plant work had indicated that an anion exchange resin scavenger may be used beneficially.

Investigations showed that iron removal in the scrub section was reasonably efficient, with the sulphuric acid playing the major part. It was found advantageous to by-pass the scrub aqueous solution from the fourth to the second stage to avoid dilution and neutralisation of the acid in the third stage. The iron in the scrubbed solvent may be reduced to less than 1 ppm.

Iron contamination on the main plant was severe owing to corrosion of overhead steel structures and from damaged rubber-lined piping in the precipitation section. Tanks, such as the O.K. liquor storage, the uranium thickeners, the ammonium sulphate storage, the final precipitator and the uranium slurry agitators have been equipped with suitable fibre-glass covers to combat contamination from outside sources.

The total sulphate content of the final product is controlled at approximately 2.5 per cent after calcining. The total sulphate content consists of fixed sulphate and soluble sulphate. The total sulphate content may be reduced by increasing the precipitation temperature and the soluble sulphate by more intense washing of the cake.

It has been found that the suitability of UO_3 as a feed material for the production of uranium fluoride is dependent on the chip strength, the granule size, a low sulphate content, a low surface area and the general purity of the UO_3.

The above mentioned physical properties of the UO_3 are influenced by precipitation conditions and the extent of washing to remove sulphate. Increased washing reduces the sulphate content, but increases the surface area and reduces the chip strength.

Higher precipitation temperatures reduce the fixed sulphate concentration, but result in higher surface areas. Short residence times have resulted in lower surface areas.

A pH of 7.3 was found to give the lowest fixed sulphate concentration, but no extensive tests have yet been carried out on the plant into the effect of various pH levels.

As a compromise, a temperature of 35°C, a residence time of nine minutes and a pH of 7.3 are being used at present. The slurry is being washed to give a sulphate content of about 7.0 gpl in the final filtrate.

FUTURE WORK

Three main courses, with the purpose of improving the process seem to be available, viz:—

Solvent loss

Due to the large volume of feed solution treated, relative to the uranium production, the South African plants are cost sensitive to solvent losses. It would seem that a solvent loss of less than 4 ppm amine in the raffinate will be relatively difficult to attain as this appears to be the solubility.

Improving the mixer design and operation could improve solvent losses.[3] Slowing down the impeller speed and generally avoiding conditions which promote shear can only lead to reduced solvent losses.

The installation of coalescers, followed by an after-settler, as inherent parts of the process, appear to be very attractive. Preliminary work has indicated that coalescers are probably a most useful ancilliary in the recovery of solvent.

Purity

Although it is not important to produce ultra-pure uranium at present, it is a factor which may become more important. A certain degree of purity may be achieved by loading the amine to the maximum and by scrubbing efficiently. However, an improved scrubbing technique is indicated and amines more specific for uranium would be advantageous.

Fire Hazard

The present solvent mixture is inflammable and this feature, coupled with the large inventory of solvent, constitutes a hazard. Diluents, other than paraffin, having a higher flashpoint, should be examined for suitability.

General

Since the solvent is not so sensitive to poisoning and competition from anions, other than uranium, it should be possible to improve the uranium leaching efficiency which in the past had been hampered by the limitations of the resin ion exchange process.

More use can be made of effluents, such as raffinate, and internal uranium plant solution recycles may be introduced.

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