

THE SULFURIC ACID PLANT
RECENT ADVANCES - REVIEW & ANALYSIS

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Introduction

In the last ten to fifteen years the sulfuric acid plant has undergone significant changes, some for the better, and some not so good. For sulfur burners and the contact section of metallurgical type (wet gas) plants, the majority of developments have involved the use of stainless steels for heat exchangers, converters, pump tanks, towers, piping, and acid cooling. Catalysts have evolved from the traditional pellet type to the lower ignition - low pressure drop ring type catalysts; and as steam pressures have increased the fire tube boiler is being moved aside in favor of the safer - low pressure drop water tube boiler. In addition, an unsteady state catalysis system was developed in the USSR that may change the face of the acid plant of the future.

Many in the fertilizer industry are looking at the Florida Institute of Phosphate Research (FIPR) process to produce aggregate and sulfuric acid from phos-gypsum. Recovery of sulfuric acid from gypsum decomposition systems requires a sulfuric acid plant with a gas cleaning system that is foreign to many in the fertilizer industry; but similar to the plant used to treat metallurgical gas or gas from the regeneration of spent sulfuric acid. The pretreatment - gas cleaning section of the plant has had a number of developments offered to the industry, some good and some not so good.

This paper will review each of the unit operations in the gas cleaning and contact sections of the sulfuric acid plant, discussing the history and recent developments in equipment designs, materials, and the handling of gas stream variables and impurities.

Gas Cleaning

The gas from a spent acid decomposition furnace, pyro metallurgical furnace used in the production of metals (Lead, Copper, Zinc, Nickel, Molybdenum, etc.), or gypsum decomposition contains in addition to sulfur dioxide, sulfur trioxide, excess water, halogens (fluorides, chlorides), nitrogen oxides, metallic vapors and fumes containing lead, arsenic, selenium, mercury, iron, etc., as well as solids and ash.

The gas cleaning area must treat the gas to remove impurities, SO_3 and sufficient water for the production of sulfuric acid in the contact section. Treatment is required to allow production of concentrated sulfuric acid while preventing catalyst plugging or degradation, and corrosion of the equipment in the contact section of the plant. The effects of some of the impurities are summarized below:

1. SO_3 - Forms sulfuric acid mist causing corrosion of contact section equipment, and may cause high acid mist emissions to atmosphere.
2. Excess Water - Dilutes the concentrated acid - causes corrosion of strong acid piping and equipment.
3. Halogens (F1,Cl1) - Causes catalyst degradation and corrosion of stainless steel components in the strong acid system.
4. NO_x - Contaminates product acid and causes corrosion of the strong acid carbon steel components - piping, product storage tanks, shipping cars.
5. Solids - Causes plugging of mist eliminators and catalyst. Contaminates product acid and may make it unsalable or unusable in fertilizer.
6. Arsenic - Causes catalyst degradation and contaminates product acid and fertilizer.
7. ~~Mercury~~ - Builds up in contact section equipment (heat exchangers) causing failures. Contaminates product acid and fertilizer.

To remove these impurities and condition the gas for the contact section, the gas cleaning area is composed of three basic unit operations.

Gas Saturation - saturates the gas with water, cooling the gas to its adiabatic saturation temperature, normally 50-70°C. This unit also provides the first stage of impurity scrubbing and converts SO₂ in the gas to sulfuric acid mist for later removal.

Gas Cooling - cools the gas to condense sufficient water for the production of concentrated sulfuric acid, and is the second impurity scrubbing stage.

Mist removal - a tube type electrostatic precipitator removes sulfuric acid mist and residual solids.

Mercury - For plants with high levels of mercury, an additional mercury removal system is added.

The first three unit operations are common to all wet gas - metallurgical type sulfuric acid plants (Figure 1). With proper equipment selection and design, the gas cleaning system will provide a properly cleaned gas for the contact section of the plant. Our experience has shown the majority of corrosion, equipment failure and down time problems in a metallurgical type acid plant usually occur in the contact section of the plant, and are traceable to problems in the gas cleaning system. For example, in one plant we were called in to investigate severe strong acid system corrosion problems causing pump, tower distributor and piping failures. The problem was determined to be poor gas cooling allowing too much water to enter the contact section of the plant, diluting the strong acid just enough to be highly corrosive. Modification of the gas cooling system corrected the problems. In another plant, severe corrosion was experienced in the contact section main blower and gas-gas heat exchangers. We determined the problem to be acid mist resulting from poor electrostatic precipitator performance due to high inlet loadings. Rearranging the precipitators for parallel sets of two units in series and addition of monitoring and control instrumentation insured precipitator performance solving the problem.

From an operators point of view, proper gas cleaning system design and operation are critical to the operation and maintenance of a metallurgical type sulfuric acid plant. The gas cleaning system is the place to put extra thought, use a conservative design, and the place to put extra money to avoid expensive downtime, maintenance, equipment failures, and early equipment replacement in the contact section of the plant.

Gas Saturation - Scrubbing

The hot, 250-350°C, gas from the decomposition system or smelter is cooled to its adiabatic saturation temperature by evaporation of water from a circulating stream of weak (5-30%) sulfuric acid. The saturation temperature, in the range of 50-80°C, is a function of the inlet gas temperature and water content. By definition, adiabatic saturation is saturation of the gas with no change in the total heat content of the gas. In addition to gas saturation, the gas saturation system removes some of the larger solid particles and is the first stage of impurity scrubbing.

To meet gas saturation and scrubbing requirements the equipment chosen must be an effective gas-liquid contacting device. Two basic types of contacting devices have been used; venturi type contactors, and open spray towers.

Open spray towers, have been used for over forty years either as separate towers or as the lower section of a combined gas saturation and cooling tower. Combined towers, sometimes referred to as Peabody Scrubbers or Swemco Scrubbers (named after the vendor who supplied them) included a lower liquid circulating spray tower and an upper independently liquid circulated and cooled tray tower in one vessel. The spray tower is an acceptable contacting device for gas saturation and scrubbing. With proper spray arrangement and design (liquid particle size and spray angle) it will perform its required function at a relatively low pressure drop - in the order of 75-100mm H₂O. Its major disadvantages are its relatively high capital cost, and compared to other available gas-liquid heat and mass transfer contacting devices, its poor efficiency.

A much better contacting device, in service for over twenty years, is the venturi contactor. In fact, essentially all gas saturators installed in metallurgical type acid plants around the world in the last fifteen years have been venturis. The venturi is an efficient liquid gas contactor and effective solids and gas scrubbing device. It is normally designed for a gas pressure drop of 75mm H₂O but can be designed, using higher pressure liquor sprays,² for zero pressure drop or even as an eductor providing a pressure rise rather than a pressure drop. Liquid circulation is significantly less in a venturi compared to a spray tower, reducing piping and pumping costs. Overall the venturi contactor is an efficient, low cost scrubbing device.

Recently Dupont has developed a variation on the traditional venturi scrubber incorporating a liquid spray into the venturi throat from below or counter to the gas flow. The device is designed using a low pressure spray and a throat sized for a gas pressure drop of 75 - 100 mm H₂O. Calculations indicate a properly designed Dupont venturi will have the same efficiency as

a conventional venturi system operating at the same gas pressure drop. Recent publications and an installation at Dupont's Burnside, Louisiana plant show the material of construction to be FRP, using a wetted wall to protect the FRP from damage from the hot 250 - 350°C incoming gas, with claims of 90% - 99% efficiency for sub-micron particles and acid mist. The use of a wetted wall protected FRP is extremely risky, risking severe damage (melting) of the venturi, resulting in an extended shut-down of the acid plant to replace the unit. In addition, particle/mist removal efficiency claims well in excess of that possible from the laws of particle dynamics, physics, or the laws of nature are highly unlikely.

In the last few years a spray tower design has been proposed using downward flow of gas with cooling of the circulating liquor. Theoretical calculations and operating experience show cooling the circulating liquor to have little effect on gas saturation - cooling in this system and I would be surprised if a unit of this design were actually installed with cooling of the saturator liquor. Analysis indicates the downflow spray tower to be less efficient than the venturi as a contacting device and more costly.

Materials of construction for gas saturation equipment are similar for all equipment types (except the Dupont venturi). The device is normally constructed of a steel shell, an impervious liner resistant to weak sulfuric acid, usually teflon or fiber reinforced polyester (FRP), covered by an acid brick lining for temperature and erosion protection. For gases that contain fluorides, an additional carbon brick lining is used to prevent fluoride attack on the silica in the acid brick. Many years ago lead was the common impervious liner material, but is not used very often today. The inlet gas duct is carbon steel and care must be taken in the design to insure the wet-dry interface is in the lined portion of the vessel to avoid corrosion of the steel inlet duct. The outlet duct is FRP. Weak acid circulating pumps are vertical plastic or plastic lined pumps, with graphite and Wilfley type plastic lined horizontal pumps also acceptable. Metal pumps, usually of high nickel alloys have been used, but are higher in cost and combined with the unusual corrosion properties of impurity contaminated weak sulfuric acid makes their choice questionable. Weak acid circulating piping is either FRP or polypropylene lined pipe, with spray nozzles of silicon carbide (Refrax) for temperature and erosion corrosion protection.

Gas Cooling

The saturated gas flows to a gas cooling system where the gas is cooled to condense sufficient water for production of concentrated sulfuric acid in the contact section of the plant. Efficient liquid-gas contacting in the gas cooler also provides a second stage of impurity scrubbing. Figure 2 shows the temperature the gas must be cooled to for production of 93 or 98% sulfuric acid, as a function of SO₂ gas strength, at sea level with a gas cooler exit pressure of -350 mm water, and an allowance for 10% of the water to be added in the contact section.. A high SO₂ gas strength minimizes the cost of the gas cooling system. Conversely, low SO₂ strength increases the difficulty of water removal and system cost, and at some point alternate and expensive cooling systems must be used to obtain proper water removal.

Equipment devices used for gas cooling include: a graphite shell and tube heat exchanger, commonly used in systems with SO₂ gas strength over 6-7%; tray towers using sieve or impingement type trays with indirectly cooled circulating liquor (Swemco or Peabody scrubbers); and packed towers also with indirectly cooled circulating liquor.

Tray towers were used as the upper stage of a combined gas saturation-cooling tower system. They are rarely used today because of high tray pressure drop for efficient heat and mass transfer, corrosion problems with the metal tray material in some plants, and the relatively high cost compared to alternate devices.

A packed gas cooling tower is the most widely used system in metallurgical gas cleaning systems with gas strength below 6-7%. The system consists of an FRP tower packed with 3" polypropylene saddle packing. Gas flows countercurrent to the liquor and is cooled by direct contact with a circulating, indirectly cooled weak acid stream. The circulating weak acid is cooled in water cooled Hastelloy plate type heat exchangers. Alternate materials are required for acid strengths above 20%, especially when in contact with gas streams with low (less than 4%) oxygen. Packed towers provide efficient heat and mass transfer at low gas pressure drop (about 100mm H₂O). Tower diameter is a function of gas flow and liquor circulation. Liquor cooler heat exchanger size is based on the SO₂ gas strength and cooling water temperature. It is common practice to provide a spare plate liquor cooler to maintain gas cooling during exchanger cleaning.

Shell and tube graphite gas coolers are used in plants with gas strength above 6-7% unless the more efficient gas impurity scrubbing of a packed tower is required, until recently primarily in plants with gas from fluid bed roasters (pyrite and zinc) and spent acid regeneration plants. The unit consists of parallel shell and tube exchangers with graphite tubes and tubesheets in a carbon steel shell. Gas flows downward through the tubes and is cooled by cooling water on the shell side of the unit. Impurities in the gas are effectively scrubbed by the condensed water augmented by a small recirculating stream of gas cooler condensate. The unit is economical for high SO₂ gas strength systems and is the preferred choice for its low cost, ease of operation, and low maintenance. As SO₂ gas strength decreases, heat exchanger surface area required for the final increments of gas cooling increases dramatically, increasing shell and tube gas cooler cost above the alternate packed tower system; with the break even point about 6-7% SO₂.

Recently a spray tower with gas flowing upward in direct contact with an indirectly cooled recirculating weak acid liquor stream has been proposed. With proper design the spray tower will perform the required gas cooling task at relatively low pressure drop. Based on heat and mass transfer principles, a spray tower is a significantly less efficient device for gas cooling and impurity scrubbing than the other alternates described above and is higher in cost.

Another recent development proposes a baffle tray tower in place of the packed tower. Baffle tray towers are one of the "least" efficient contacting devices available, and it is surprising anyone would propose its use in this decade. Their low efficiency for both gas cooling and impurity scrubbing as well as high cost would rule out the baffle tray tower for use in this system.

Mist Removal

The final treatment stage of a metallurgical type acid plant gas cleaning system is the removal of residual solids and sulfuric acid mist (formed by reaction of SO₃ in the gas with water). Mist/solids removal efficiencies of greater than 99% for particles less than 1 micron are required (producing an optically clear gas) to protect contact section equipment from acid corrosion, solids contamination of product acid, and catalyst plugging.

The device used for mist/solids removal for over forty years is the tube type electrostatic precipitator, usually arranged in parallel sets of two units in series. The units are constructed of up to 240, 250 mm diameter tubes, 5-6 meters long supported

between tube sheets. The tubes are grounded and have a discharge electrode running down the center. The gas flows upward through the tubes where mist and solids particles are charged by a Corona at the tube entrance (ionization of molecules by electron collision in a region of high electric field strength) generated by the high voltage potential between the electrode and the grounded tube wall (normally 50-90 KV). As the gas and particles move up through the tube, charged particles are repelled by the similar charge of the electrode and attracted to the grounded tube wall. The tube must be sufficiently long to allow time for the charged particles to migrate to the wall and be collected.

The efficiency of an electrostatic precipitator can be represented, in simplified form, by the Deutsch-Anderson equation:

$$\text{Eff} = 1 - \exp (W \cdot A / Vg)$$

A = Area of collecting surface

Vg = Gas flow rate

W = Precipitator rate parameter

The precipitation rate parameter, W, is a function of particle size and resistivity, distribution of gas, back corona and reentrainment, electrical field strength, particle charge, gas viscosity, and particle density. Increasing collecting area relative to gas velocity increases collection efficiency. Precipitators are operated at the maximum voltage potential, up to the point of arcing (sparking) for maximum efficiency. Allowable voltage potential (spark point) decreases with increasing inlet loading; therefore, for maximum mist removal, precipitators are arranged in parallel sets of two 90% efficiency units in series, so the second unit will operate at maximum voltage potential and efficiency at all times.

Acid mist precipitators are constructed of lead tubes supported between lead lined tube-sheets and heads. Electrode wires are usually lead star wires. The system includes a power rectifier to convert AC power to constant DC at 60-90 KV, and control instrumentation to drive the voltage potential to a maximum level - up to the point of sparking.

Two problems with lead precipitators; the lack of available lead burners and fatigue failure of the tubes over time - especially in systems with variable gas flow, led to the introduction of FRP - plastic precipitators about eighteen years ago. The first plastic tube precipitator installation in North America started up at the Anaconda smelter in 1973. Plastic precipitators using FRP tubes supported in FRP tubesheets and heads, are an acceptable alternate to lead with the choice dependent on economics. Recent cost analysis indicates plastic precipitators

to be 20% - 40% higher in cost than lead units.

An alternate mist removal device, of a type around for about twenty years, but little used due to its poor efficiency for fine particles, is recently being offered by LAB of France. The LAB unit, called an Electro Filtering Venturi, uses a high tension electrode to enhance the small particle collection efficiency of a low pressure drop venturi. The operating theory of this unit is described below:

The efficiency of wet particle scrubbers is limited by the ability to capture particles by the water droplets. High efficiency scrubbers are operated with large pressure drops to achieve a high relative velocity between the particle and the water droplet. Fine particles require a very large pressure drop because the mechanism that transports the particle through the water droplet boundary layer is primarily inertial. The concept of electrical augmentation of a wet venturi scrubber is to charge the water and particle with an opposite polarity in an attempt to utilize the attractive electrical forces to bring the particles and water droplets together. Literature reports (both theoretical calculations and test results) show a small increase in collection efficiency for particles less than 1 micron. From the available data, we do not believe electrical augmentation of a low pressure drop venturi scrubber will provide the dust and acid mist removal efficiency required to protect the contact section of the plant. This has been confirmed during operation of the first Electro Filtering Venturi in an acid plant gas cleaning system of a European Zinc smelter, the unit was replaced/augmented by tube type mist precipitators.

Monsanto is proposing a gas cleaning system without a mist precipitator. The system would use a Dupont scrubber, baffle tray tower, followed by a second Dupont scrubber. Claims of greater than 99% removal of sub-micron solids and acid mist particles required to protect the contact section of the plant from contamination and severe corrosion have been made. The risk of using this system is enormous and is considered unacceptable.

Impurity Scrubbing - Purge Treatment

Impurities in the gas are removed by the gas saturation, gas cooler, and mist precipitator. For efficient scrubbing, liquor is routed from the least contaminated (mist precipitator) to the gas cooler and finally to the gas saturator. For efficient fluoride removal silica is usually added to the scrubber solution producing a low vapor pressure silicon fluoride. Mercury, for the most part, will travel through the traditional gas cleaning system untouched and requires a separate treatment step. Mercury can be converted to a sulfide in the gas phase by addition of

hydrogen sulfide and absorbed on activated carbon beds; or it can be converted to a chloride by scrubbing with a circulating hydrochloric acid stream in a packed tower, with the insoluble mercurous chloride recovered by filtration. Another technique converts the mercury to a sulfate in a pre-drying tower using 70-85% sulfuric acid. The sulfate is then converted to insoluble sulfide by the addition of thiosulfate and removed by filtration. Each of the systems will effectively remove mercury, with the chloride route preferred by most.

Saturator liquor is usually maintained at 3-5% undissolved solids, and depending on the SO_2 level in the gas, corresponding to a sulfuric acid content of 5-30%. To maintain the solids level in the saturator liquor some plants with high inlet solids loadings require the addition of make-up water to augment the water condensed from the gas. Excess saturator liquor (from water condensed from the gas and make-up water) is purged from the saturator and treated prior to reuse or disposal.

The purge liquor is first air stripped of dissolved SO_2 in a packed stripping tower, then neutralized with lime to neutralize the acid and precipitate the impurities as calcium salts or hydroxides, and filtered. In most cases the treated purge water is reused in other areas of the complex, but can be discharged.

Effects of Inlet Gas Variables

Each of the gas stream variables will have an effect on the gas cleaning system. A thorough understanding of the effects of gas stream variables is required to properly select, design, and operate gas cleaning system equipment.

Gas Flow & SO_2 Strength - For the most part the sizing and cost of the gas cleaning system is a direct function of the gas volume. For a particular smelting/decomposition rate a specific quantity of sulfur as SO_2 is routed to the sulfuric acid plant. As SO_2 gas strength increases, the gas volume and cost of the gas cleaning system decrease (See Figure 3). Variation in gas flow, typical in a plant receiving gas from copper smelter Pierce Smith converters, is handled effectively in the plant, but will cause accelerated fatigue failure of the lead tubes in the mist precipitator.

Solid & Halogen Impurities - Metals and metal salts will generally be converted to sulfates by reaction with the weak acid circulating liquor and be removed in the saturator, gas cooler, and mist precipitator. Vapor components, such as arsenic oxides, will condense and also be efficiently removed. Fluorides, as mentioned above, requires silica addition to produce the low vapor pressure silicon fluoride for removal.

The quantity and size distribution of the solids in the gas will affect the quantity of weak acid purge and the sizing of the purge treatment system. At high levels and small size, mist precipitator sizing will also increase.

Sulfur Trioxide - Sulfur trioxide in the gas will be converted to sulfuric acid on cooling and contact with water in the gas saturator. The majority will form submicron mist particles, by a phenomenon called "Fog Formation", and pass through the saturator and gas cooling systems to be removed in the mist precipitators. Care must be taken to design the mist precipitators conservatively, for the highest levels of acid mist and solids anticipated. Higher than design mist/solids loadings will back-off precipitator voltage, reducing efficiency.

Nitrogen Oxides - Nitrogen oxides, formed in increasing levels by high temperature burners and oxygen smelting, will pass through the gas cleaning system and must be removed in the contact section of the plant. Most will go through the plant to the stack and the rest, fortunately, will collect as nitrosyl-sulfuric acid in the high efficiency mist eliminators of the contact section absorption towers, where they can be easily removed and treated. NO_x contaminated strong sulfuric acid has an enormous effect on the corrosion properties of the acid. Stainless steels are passivated, with corrosion rates dropping to near zero, while corrosion rates of carbon steel (used for product piping, storage tanks, and shipping cars) is increased by orders of magnitude - in the order of 1/2" per year or greater. The dramatic increase in corrosion of carbon steel has been experienced in a number of acid plants over the past fifteen years, and about seven years ago Kennecott materials group published the results of test work on NO_x contaminated acid, with the laboratory tests confirming and quantifying plant operating experience.

Contact Section

We are all-familiar with the process description and equipment arrangement of the modern sulfur burning plant. The following describes the typical metallurgical type wet gas plant that will be common place in the fertilizer industry when gypsum decomposition systems are installed.

In the contact section of the plant, cleaned SO_2 gas is converted to SO_3 then absorbed in water contained in 98% sulfuric acid to form sulfuric acid. Flow through the system is shown on figure 1 and described below:

Gas Flow - The wet, cleaned, gas from the purification system enters the drying tower where it is dried to a dew point of -40°C to prevent acid condensation corrosion of downstream equipment. In the drying tower the gas is contacted with a circulating stream of 93% sulfuric acid that absorbs the water from the gas. Dry SO_2 gas leaves the tower and passes through an entrainment separator before entering the main blower. The blower provides the suction to draw the gas through the gas cleaning system and drying tower, and the pressure to force it through the remainder of the plant.

The blower discharges to a series of gas to gas heat exchangers to heat the gas to catalyst ignition temperature of about 420°C , using heat from the converter SO_2 to SO_3 oxidation reaction.

The heated gas flows to the converter where the SO_2 is oxidized to SO_3 over a vanadium pentoxide catalyst. The converter contains four catalyst layers with cooling between each bed. Cooling is provided to increase overall conversion as shown in Figure 5. The figure shows a typical equilibrium line for 11.5 % SO_2 on a plot of percent conversion versus temperature. The first catalyst bed operating line is shown (line 1), with gas temperature increasing (heat of reaction) with conversion, and approaching the equilibrium line. The gas is indirectly cooled (line 2) and enters the second bed where conversion increases (line 3). Two basic types of converter arrangements are used; single absorption where the gas passes through four converter catalyst beds followed by SO_3 absorption (98% conversion of SO_2); and double absorption, where the gas is withdrawn after two or three passes over the catalyst, SO_3 is removed by absorption in sulfuric acid and the gas is returned to the converter for final conversion in the remaining one or two beds. Double absorption plants operate at a conversion efficiency of greater than 99.5%-99.7% depending on gas flow variation and SO_2 strength.

Gas from the converter is cooled in gas to gas heat exchangers before entering the absorption tower where the SO_3 is absorbed in water contained in a circulating stream of 98% sulfuric acid. Gas from the absorption tower, with 98%, or 99.5%-99.7% (double absorption) of the incoming SO_2 removed as sulfuric acid is discharged to atmosphere via a stack.

Acid System - In the drying tower 93% sulfuric acid entering the tower at 43°C absorbs water from the gas. A dual (93% & 98%) acid system arrangement is required to prevent SO_2 in the gas entering the drying tower that is absorbed by the acid from being stripped in the absorption tower and increasing SO_2 emissions. As the acid passes through the tower it is heated by the heat of condensation of water and dilution of the acid. Acid leaves the tower and flows to a pump tank, where it is pumped through a

shell and tube acid cooler before being returned to the tower. 98% acid from the absorber acid system is added to the pump tank to maintain drying acid concentration at 93%. A portion of the drying acid is transferred to the absorber acid system, via an SO₂ stripper, to control absorber acid strength. Product 93% acid is withdrawn from the stripper, cooled in an acid cooler and sent to product storage.

Sulfur trioxide is removed from the gas in the absorber system by contacting the gas with a circulating stream of 98% sulfuric acid. The absorber acid enters the tower at 80°C and is heated by sensible heat in cooling the gas, and the heat of condensation of SO₃. The acid leaves the tower and enters a pump tank where it is pumped through an acid cooler and returned to the tower. Water in the SO₂ gas, that enters the contact section of the plant from the gas cleaning system and removed in the drying tower, is added as 93% acid, to the absorber acid system to combine with the SO₃ removed from the gas forming sulfuric acid. About 10% more water than required for water balance is usually removed in the gas cleaning system to allow for concentration control by the addition of liquid water to the strong acid system.

Effect of Variables

There are three variables that effect the sizing, cost and operation of the contact section of the plant. These are gas volume, SO₂ gas concentration, and the amount of water in the entering gas.

Gas Volume - The majority of equipment in the contact section of the plant is sized based on the volume of gas to be handled. This includes; acid towers, gas to gas heat exchangers, converter and gas ducts; representing the bulk of the cost of the contact section of the plant. The remaining equipment, gas to gas heat exchanger, acid circulation and cooling, are sized based on acid production and SO₂ gas strength.

SO₂ Gas Strength - The SO₂ gas strength is critical to the design, equipment sizing and operation of the contact section of the plant. The upper operating limit for SO₂ concentration in a conventional type plant to prevent overheating the catalyst in the first bed (loss of vanadium/activity), is 11.5% SO₂ at an O₂/SO₂ ratio greater than 0.8. A high strength design (Patented by L.J. Friedman) developed to handle gases from oxygen flash smelting and regenerative scrubbing systems operates at 20%-22% equivalent SO₂ to the converter.

The lower limit of SO₂ concentration, without the addition of external heat, is 3%-3.5% for a single absorption plant and

3.5%-4% for a double absorption plant (3.5% is based on a co-current venturi type interstage absorber). This lower limit is based on the available heat recoverable from the converter oxidation reaction needed to heat the cold incoming SO₂ gas to converter catalyst ignition temperature (defined as the autothermal SO₂ level). Plants can be operated at SO₂ strengths below the autothermal level with the addition of heat, usually provided by fossil fuel in indirect preheaters.

Gas to gas heat exchanger sizing and cost vary with the SO₂ gas strength and flow. As SO₂ strength decreases the size and cost of the exchangers increase, especially at levels below 5%-6%. For a fixed decomposition/smelting rate, sending a fixed amount of SO₂ to the sulfuric acid plant, the SO₂ strength and gas volume are obviously related. As the SO₂ strength increases the gas volume decreases along with the size and cost of the plant. Figure 3 shows the relationship of gas cleaning, contact section, and total plant capital cost as a function of gas strength for a plant processing a fixed quantity of SO₂.

Water Balance - All of the water that enters the contact section of the plant leaves as product acid. Therefore, the amount of water permitted in the gas entering the drying tower is limited by the product acid strengths produced by the plant (Figure 2). A properly designed plant would reduce this value further by requiring a minimum of 10% of the water added as liquid to the acid system for acid strength control. As SO₂ gas strength decreases, water balance considerations become critical, and at some point acid strength can not be maintained with conventional gas cooling, and expensive gas cleaning system gas cooling using refrigerated gas coolers or pre-drier - concentrator systems are required. If too much water enters the contact section of the plant, acid strength in the drying and absorber systems can not be maintained. The diluted acid will cause extensive corrosion damage to the acid system equipment (pumps, coolers, piping, tower internals, instruments).

Some of the problems associated with gas volume, SO₂ strength, and water balance are illustrated in the following examples. In one plant handling gas from a copper smelter Pierce Smith copper converters, the acid plant design was based on 200% air leakage at the converter hoods (considered a conservative number) resulting in an SO₂ strength at the acid plant of 4%-8% averaging 6%, with gas cooling comfortably able to produce 93% product acid (Figure 5). After about six months operation converter leakage increased to 400%-500%, reducing SO₂ gas strength to 2%-4% averaging 3%. The acid plant designed to accept a fixed volume of gas could not handle all of the additional gas and a portion was bypassed direct to the stack. SO₂ gas strength was below the autothermal level requiring continuous operation of the fossil fuel fired preheater to maintain the plant in heat balance,

increasing operating costs. Water balance could not be maintained, and the smelter purchased 98% sulfuric acid for addition to the plant to remove the excess water as 93% acid. Another plant had similar problems, but added an expensive pre-drier - concentration system to solve the water balance problem as well as extensive modifications to the plant to increase gas handling capacity.

Equipment/System Design

Gas Heat Exchangers

Gas to gas heat exchangers use the heat of the converter SO_2 to SO_3 oxidation reaction to preheat incoming cold SO_2 gas to catalyst ignition temperature; and in double absorption plants, to cool and reheat the gas entering and leaving the interstage absorption system. The exchangers also provide the cooling between catalyst beds for high conversion efficiency. In plants with high SO_2 strengths, excess heat is available from the converter system and is used to preheat air, heat boiler feed water and/or superheat steam.

Twenty years ago gas - gas exchangers were designed as shell and tube, segmented baffle full bundle units with heat transfer coefficient of $10-20 \text{ Kg-Cal/M}^2\text{-}^\circ\text{C-Hr}$. Low heat transfer coefficients and high gas pressure drops of this design led to the development of alternate designs commonly used today. These include shell and tube single pass cross flow exchangers, multi-pass cross flow exchangers using baffles with no tubes in the baffle window, and disc and doughnut exchangers with no tubes in the baffle window (disc and doughnut cross flow). Each of these designs operate at comparable pressure drops with heat transfer coefficients three times the old full bundle design.

Cold heat exchangers are normally constructed of carbon steel. Hot units (heat exchanger between the first and second catalyst bed, and the exchanger between the second and third bed) are constructed of $1 \frac{1}{4} \text{ Cr} - \frac{1}{2} \text{ Mo}$ alloyed tubes for high temperature strength and scale resistance or 300 series stainless steel, with the choice dependent on economics. Essentially all new acid plants are using stainless steel for hot exchangers. This may change when the cost and delivery effects of the recent five fold increase in Nickel prices fully reach the stainless steel market. These large units require careful design, including considerable oversurface for fouling, and incorporating tube support and detuning baffles to prevent vibration.

Gas flow through cold exchangers, where acid condensation is possible, is arranged with the SO_3 gas flowing down through the

tubes to force condensed acid out of the tube bundle and into the bottom head of the exchanger. Hot exchangers are also arranged with the SO_3 gas in the tubes to avoid the need for temperature and scale protection of the shell.

Boilers - Steam System

For sulfur burning plants, the main changes in the steam system have been aimed at increasing energy recovery. These have involved higher steam pressures, higher steam temperatures, and lower economizer gas exit temperatures. Steam pressures have increased to the point that 900 psig is commonplace, with corresponding temperature increases to 480-500°C. The converter boiler has been replaced with a steam superheater, and cast iron gill ring economizers have given way to carbon steel fin tube units that are more compact and lower in cost.

The most recent change is in the furnace boiler. As plants have gotten larger the size and cost of the furnace boiler has increased, forcing designers to increase boiler pressure drop to as high as 500-650 mm H_2O in an effort to reduce cost. These high pressure drop units have reduced available energy recovery by increasing main blower pressure rise. Recently a number of plants have installed a water tube boiler, reducing boiler pressure drop to 125-250mm H_2O . A water tube boiler is safer, since the high pressure water and steam are in the tubes and steam drum instead of the high volume 75-100mm thick boiler shell. This also permits steam pressures to be increased safely to 1250-1450 psig, increasing energy recovery. All new energy efficient plants will use water tube boilers.

In the economizer area, the old wives tale acid dew point gas exit temperature minimum of 200°C is a thing of the past. Over ten years ago, in a paper I presented to the AIChE, theory was presented to demonstrate the fallacy of the then thinking on economizer acid dew point. A review of the various heat transfer coefficients showed the gas side tube wall temperature to be essentially the water temperature; therefore, if no condensation occurred in current units at 200°C, then acid condensation would not occur at gas temperatures of 120-150°C. One contractor developed an elaborate water flow scheme to prevent acid condensation, and sold a number of units capitalizing on acid dew point fears. All modern energy efficient plants currently operate low temperature economizers constructed of carbon steel finned tubes with no special or complex water flow arrangement or acid dew point problems.

Converter System

Catalyst - Sulfuric acid oxidation catalyst of essentially the same basic formulation has been in use for over fifty years. The catalyst contains 6-9 wt% vanadium pentoxide as the active component with potassium and sodium sulfate promoters, supported on a diatomaceous earth (silica) carrier. Catalyst ignition temperature is a function of the vanadium and sodium/potassium sulfate content and is about 410°C. The upper temperature limit for continuous operation is 620°C to prevent loss of vanadium and reduced activity. For many years catalyst was in the form of extruded pellets 6 mm in diameter by 10 mm long. Clean pellet catalyst beds operate at relatively high pressure drop, about 800-1200 mm water total for the four beds. The catalyst is also a good filter of dust particles, especially at operating temperature when the material is sticky. Dust is held in a shallow layer at the top of the bed increasing bed pressure drop about 800 mm over a one year operating period, reducing plant gas handling capacity and requiring screening to remove the dust.

About twelve years ago Monsanto marketed a larger pellet, called 516, that is 5/16" or 10 mm in diameter to reduce clean pressure drop and increase dust holding capacity (reduce pressure drop build-up). 516 catalyst is little used today. At about the same time Chemetics developed and marketed a "radial flow converter" design for the first catalyst bed where the catalyst was in candles with increased surface area to reduce pressure drop and increase dust holding capacity. One unit was built at the Copper Cliff acid plant, but extremely poor economics have prevented other installations.

In the same time period, Haldor Topsoe developed a 10 mm diameter ring shape catalyst. Topsoe rings have a clean pressure drop about one-third that of pellets and a dust holding capacity to reduce dust pressure drop by a factor of six (from 800 mm to 130 mm water). All new sulfuric acid plants and most existing plants use ring catalyst. All catalyst manufacturers now produce ring catalyst, with operating experience showing about equal performance.

Converter Design - Sulfuric acid converters are a vertical cylindrical vessel with internals to support four separate layers of catalyst. Until recently the vessels were constructed of a carbon steel shell (metalized with aluminum in hot areas for scale resistance) with either cast iron posts or steel beams supporting cast iron grids for the catalyst. Units of this design have been in operation for over thirty years and have operated successfully with minimal maintenance.

Currently all acid plant designers are using a stainless steel converter design (300 series). Stainless steel converters use a stainless steel shell, hemispherical domes, posts and beams to support stainless steel catalyst trays. The design evolved from an old European design where stainless steel catalyst trays were supported by brick hemispherical domes and piers in a brick lined steel shell. There are a number of installations of stainless steel converters, and the design is considered satisfactory. One should be cautious to insure the converter shell design pressure is above main blower shut off pressure, and the catalyst trays (first bed) are designed conservatively above expected dirty gas pressure drop. Many of these converters are designed for the normal operating pressure and pressure drop to reduce metal thickness and cost. In one case a damper was inadvertently closed down stream of the converter, putting the blower in a surge condition, increasing system pressure to blower shut off pressure, causing extensive damage to the converter. Stainless converters offer a number of advantages at a small premium cost over steel converters. They eliminate the need to metalize hot areas of exposed steel to prevent scale, and the increased metal strength at converter operating temperatures permits the location of the first catalyst bed to be at the bottom (rather than the top), simplifying the first mass exit duct expansion problem that has caused many leaks with some designs.

Unsteady Oxidation Process

This new method of SO_2 oxidation is based on a periodic reversal of the direction of the reaction mixture flow over the catalyst bed (Figure 6). The process was developed at The Institute of Catalysis of the USSR Academy of Sciences in Novosibirsk, Siberia, USSR, by Prof. Yuri Matros, et.al., and covered by patent around the world (U.S. patent no. 447 88 08).

Basically a large bed of catalyst is used as both a reversing regenerating heat exchanger and as a catalytic reactor for the SO_2 oxidation reaction. Cold SO_2 gas is fed into the catalyst bed and is heated by the heat stored in the bed to catalyst ignition temperature. At this point the conversion reaction proceeds, producing heat. The heat is absorbed by catalyst in the bed, increasing its temperature. When the temperature front comes close to the exit side of the bed, the flow through the reactor is reversed (Figure 6). Flow reversals are made every 30-120 minutes. The main advantage of the unsteady process can best be seen on Figure 5, where the operating line for bed one would be almost vertical, giving a one bed conversion of about 80%-90% at a low bed exit temperature, compared to about 57% conversion at an exit temperature of over 600°C for the conventional process.

The claims for the process include lower capital cost by saving heat exchanger equipment - surface area, lower pressure drop, and the process is autothermal at low (0.5-3%) SO₂ gas strengths. Analysis indicates the major advantage of the unsteady process to be for weak metallurgical gas plants where very large gas heat exchangers can be eliminated or significantly reduced in size, and the plant can operate autothermally - without fossil fuel external heating. For rich gas plants the degree of savings is yet to be proven. The process is currently in operation in four weak gas (2%-4%) metallurgical plants in the USSR, and was licensed by Lurgi for a weak gas metallurgical plant in Eastern Europe. I expect to meet with Prof. Matros in the USSR in the near future, and will have further details on the process after seeing the operating plants and completing a more detailed operational and economic analysis for high SO₂ gas strength plants.

Acid System

Each acid tower system (drying, interstage absorber, final absorber) is composed of a gas-liquid contacting tower, acid collection/surge volume either in a pump tank or a reservoir in the base of the tower, circulating pump, acid cooler, and acid entrainment/mist removal device. The acid system has undergone significant change over the last fifteen years with current work aimed at recovering heat as low pressure steam from 200°C circulating acid.

Acid Towers - An efficient liquid-gas contacting device is required to remove water from the gas in the drying tower and absorb SO₂ from the gas in the absorption towers. The 1980's conventional acid tower is a steel tower lined with Picora mastic, 10 mil teflon sheet and acid brick. Two layers of acid brick are used in the bottom section of the tower up to the packing support, and one layer of acid brick for the remainder of the straight side. The tower bottom is a dished head with a bottom acid outlet. The packing is supported on a ceramic Steuler dome to transfer the load to the tower shell and avoid brick arches or piers. Drying towers are packed with 2500 mm to 3050 mm of 3" ceramic intalox saddles, topped with 600 mm of 2" saddles to aid in acid distribution and reduce acid entrainment. Absorption towers have 3050 mm to 3650 mm of 3" packing. The preferred acid distributor is cast iron trough and spouts with a minimum of ten acid feed points per square meter of tower cross-section and the acid discharging from the spouts 30-100 mm below the top of the packing. This design is well proven, highly efficient and flexible.

Recent difficulty in obtaining quality castings for trough and spout distributors as well as cost, has led most designers to offer pipe distributors, usually buried in the 2" packing. This is an old design that lost favor about twenty years ago due to problems of poor distribution and tower performance resulting from distributor holes plugging and/or corroding, and leaks in flanges, all below 300-600 mm of packing. Recent pipe distributors are offered with teflon outlets, Lewmet nozzles, and can be constructed of a silicon stainless alloy (approximate composition - 17Cr, 17Ni, 4Si, 2-5%Cu) marketed by Chemetics as Saramet[™] and Edmeston of Sweden as Sandvik SX[™]. However, many of the problems of pipe distributors still remain. In addition the high silicon stainless has a much narrower operating range to minimize corrosion. Corrosion data indicate both Saramet and Sandvik SX are acceptable up to 150-160°C at acid concentrations above 96%. Severe corrosion occurs at higher temperatures (gas inlet to absorption towers) or lower acid concentrations (vapor space of pump tanks, around gas inlets of drying towers). These corrosion effects have occurred in operating units. Cast iron has a demonstrated operating range of 80% to 99+%, whereas the silicon stainless is limited to operation at acid concentrations above 96%. Again high Nickel prices and the dollar exchange rate for these foreign alloys will most likely eliminate them from consideration.

An alternate tower packing support uses acid brick arches spaced about 2-2.5 meters apart, supporting Aludur ceramic beams and ceramic grid blocks. Although acceptable and used for many years, this design has the disadvantages of having brick arches cluttering the bottom of the tower with the arches supported from the curved brick lined tower bottom, and transferring the tower load to the tower bottom putting more stress on the brick lined bottom head.

A few years ago a 98% acid drying tower and pump tank constructed of unlined Saramet went into operation at the Simplot sulfur burning acid plant in Idaho. Saramet and the Sandvik SX alloy have also been used for acid pipe distributors, piping and acid coolers. Recently an absorption tower was started-up at the Stauffer Baton Rouge plant. Drying tower problems include dilute acid corrosion in the area of the tower near the moist gas inlet; pump tank ~~problems~~ appear as corrosion in the vapor space again due to low ~~acid~~ concentration; and absorption tower problems are expected ~~in~~ the hot gas inlet area of the tower where metal temperatures exceed allowable operating temperature for the silicon stainless alloy. High silicon stainless alloy tower cost was running about 20% above a conventional tower when the dollar was strong and nickel prices were \$1.60-\$3.70 per pound. The current exchange rate and nickel price (\$7-\$10 per pound) should increase the cost differential to 30%-50% above a conventional tower.

Alternate tower designs include the Lurgi venturi tower and the Monsanto HE tower. The Lurgi venturi tower uses a two stage, horizontal and vertical, venturi for acid-gas contacting followed by a packed and irrigated separator tower. In weak gas metallurgical plants the Lurgi venturi has been used as an interstage absorber without acid irrigation on the separator, where the co-current gas and acid flow leaves more heat in the exit gas, reducing interstage heat exchanger cost and allowing the plant to be autothermal at a lower SO₂ concentration. The design is well proven in acid plants around the world for over seventeen years. Marginal economics and reduced operating flexibility have limited installations in North America.

The Monsanto HE tower has been on the market for about five years with a troubled history in many plants. The design is based on work performed by the Fractionation Research Institute indicating increased packing mass transfer efficiency can be obtained with increased liquid distribution. The HE tower originally used a Mondri cast iron pipe distributor incorporating Lewmet nozzles with 4-5 times the number of acid feed points per square meter of cross section, and 3" packing depth reduced to about 2150 mm in the drying tower and 2450 mm in the absorption towers. No 2" packing is used and the acid pipe distributor is located above the packing. The resulting acid spray/mist from the distribution system puts a high loading on the mist removal device (about ten times trough and spout or buried distributors). Tower design velocity (tower diameter) is the same as the 1980's conventional tower, but is less flexible and more sensitive. The reduced packing depth requires optimum acid distribution to maintain efficiency and the high entrainment/mist loadings require maximum removal efficiency be maintained by the mist removal system. Problems experienced in the approximately thirty some-odd towers built to date include poor performance due to plugging of the distributor holes with brick chips, packing chips and/or sulfate (Lewmet screens are used to help prevent this problem, after failure of Hastelloy screen material); corrosion failure (casting defects) in the Mondri cast iron pipe distributor - Meehanite cast iron or Sandvik SX is now used; and quality, efficiency and acid reentrainment problems with the Monsanto CS series entrainment - mist eliminator candles that must be augmented with reentrainment separators.

Acid Pump Tanks - Pumps - Acid pump tanks provide acid surge volume for the vertical acid pumps and sufficient volume to hold the acid contained in the system (tower, piping, cooler) on pump shut-down. Pump tank design has developed over the years from a horizontal cylindrical brick lined tank to a vertical cylindrical, flat bottom, brick lined tank, usually 2290 mm high; to alternate designs using a dish bottom to avoid the brick expansion problems of a flat bottom; back to horizontal

cylindrical brick lined tanks to avoid corrosion problems of the unlined flat steel tank roof; or eliminating the tank altogether by locating the tower at grade and using the base of the tower for the required acid surge reservoir, with external vertical pump with or without a pump boot. Each of the alternate designs is acceptable, with the horizontal lined tank and reservoir in the base of a tower at grade the preferred - most widely accepted today. Recently Chemetics and Monsanto have installed an unlined horizontal cylindrical tank constructed of high silicon stainless steel which is marginal due to high corrosion in the vapor space, and the lack of acid concentration flexibility. Once again cost (dollar exchange rate and nickel prices) as well as limited operating flexibility should eliminate high silicon stainless steel from consideration.

Acid Pumps are traditionally vertical pumps to avoid suction valves, seals and packing. Essentially all acid plants around the world use Chas. Lewis vertical pumps constructed of their proprietary process iron and Lewmet high nickel alloy. Pump operating life between required maintenance is a function of acid concentration and temperature, and is about one year with 98% acid at 115°C.

Acid Coolers - Acid coolers have progressed from parallel banks of stacked "S" shaped cast iron sections to stainless steel shell and tube or plate exchangers, with and without anodic protection. Chemetics has marketed anodically protected stainless steel shell and tube acid coolers for about eighteen years, with numerous installations around the world. About eight years ago Monsanto introduced its version of the Chemetics acid cooler and is considered acceptable. A few years ago Edmeston installed a non-anodically protected shell and tube acid cooler using the Sandvik SX alloy at the Boliden Rohnskar smelter in Sweden. Boliden is also using the Sandvik SX and Saramet alloys for an acid distributor and acid piping.

Alternate acid cooler designs, used mainly in Europe with mixed success, include Hastelloy plate exchangers with and without anodic protection, teflon shell and tube exchangers, and cast iron tubed air coolers.

Entrainment Separators / Mist Eliminators - Acid entrainment separators are used in the top of the drying tower to remove acid droplets from the gas to prevent corrosion of the main blower and gas heat exchangers. Typical entrainment separators collect by particle impingement and use a 12.5 mm thick woven teflon pad or candle with compressed fiberglass in a stainless steel wire frame. The majority of installations in recent years have used the candle type, but problems with fiberglass size distribution,

quality, and manufacturing defects have caused difficulty in many plants.

Mist eliminators are required in the absorption towers to remove acid entrainment and sub-micron acid mist. Two types of units are used; dual pad or candle impingement devices and candle type Brownian movement devices. The dual pad incorporates a high density woven pad to agglomerate mist particles to large droplets, and a low density woven teflon pad as an entrainment separator for the large droplets. Candle mist eliminators of compressed fiberglass using impingement have about the same efficiency as the dual pad device, candles using Brownian movement as the collecting method can be designed for efficiencies of greater than 99% for particles less than one micron.

Impingement type mist eliminator efficiency is gas velocity sensitive with collection efficiency decreasing with decreasing velocity or gas flow. The units have been used successfully in all types of acid plants (including metallurgical acid plants with varying gas flow) for about twenty years, and are low in cost. For plants producing oleum (generates a large quantity of sub-micron particles) and plants wanting a more efficient mist removal device, high efficiency candle units are used. The candles are constructed of 50 mm to 75 mm thick compressed fiberglass supported in a stainless steel wire frame. Collection is by Brownian movement of particles at laminar gas velocity (low velocity). One advantage of these type candle units is collection efficiency increases as gas flow - velocity is reduced. High efficiency candles cost about five times that of impingement type devices.

Acid Heat Recovery - Currently the major emphasis in sulfuric acid plant development is directed at recovering heat at usable levels from the acid system. Anodically protected shell and tube acid coolers have been used (about sixteen years) for preheating boiler feed water to 95°C and producing hot water for process heating. Boliden is using acid heat as the heat source in their sea water absorption-stripping SO₂ concentration system. Recent work is aimed at recovering acid heat as low pressure steam (50 psig - 100 psig).

To recover acid heat directly as steam acid temperatures must be increased from 110°C - 120°C to about 200°C, requiring new materials of construction for acid system components. All major acid plant designers are actively working in this area. Chemetics efforts are reported to involve using anodically protected Saramet; while Monsanto is marketing a system using 300 series stainless steel (310) in a narrow acid concentration range of 99.5% to 100% strength - operation outside this range will

result in severe corrosion. We anxiously await operating data from the first unit sold for installation in a Falconbridge plant in Europe (expected to come on line in the end of 1987) and the unit at Namhae in Korea to determine whether a plant can be operated over the short or long term within the required concentration constraints and avoid severe corrosion damage. The Namhae unit is reported to have had a minor concentration excursion during start-up resulting in severe damage to the acid pump (corrosion rates - metal loss in other areas of the unit was not reported). Another major risk of the Monsanto system is the possibility of a leak in the acid cooler - steam boiler. Acid coolers have always been designed to leak acid to water to allow easy detection and minimize damage to the heat exchanger; the Monsanto system operates with 100-150 psig steam on the water side, so a leak will force 100-150 psig steam into 175-200°C 99+% acid resulting in hot dilute acid and severe damage to the entire system.

Final Comments

The 1980's is an unusual period for accelerated development in sulfuric acid plant technology. The depressed state of the engineering business has forced engineering contractors to reduce staff and let go many experienced people. In many cases, design changes and process developments are being implemented by relatively inexperienced people lacking the years of design and operating experience needed to minimize errors and reduce the learning curve to acceptable levels. Life was much simpler in the 1970's, when a fairly reliable and good operating plant could be obtained by just calling one or more acid plant contractors and leaving the design to them. This is not true today, where operating personnel must be intimately involved in all areas of the acid plant design, equipment and materials selection to avoid one of a kind and/or high risk designs that may be detrimental to acid/energy production or put the entire fertilizer operation in jeopardy.

For any new or major modification acid plant project, especially in this current environment, Acid Engineering & Consulting, Inc. strongly recommends project development, analysis and execution be by a team composed of operating, engineering, and materials personnel, augmented with consultants with extensive sulfuric acid plant design and operating experience. The combined knowledge and experience of the team and team approach to project development and execution assures a highly reliable, good operating and low maintenance acid plant at the lowest capital cost.

The aim of this presentation was to provide an understanding of recent acid plant developments, and the design and operating considerations of a metallurgical as well as sulfur burning sulfuric acid plant. A better understanding of the factors that affect equipment design, materials, operation, and the effect of variables should lead to a successful sulfuric acid plant project.



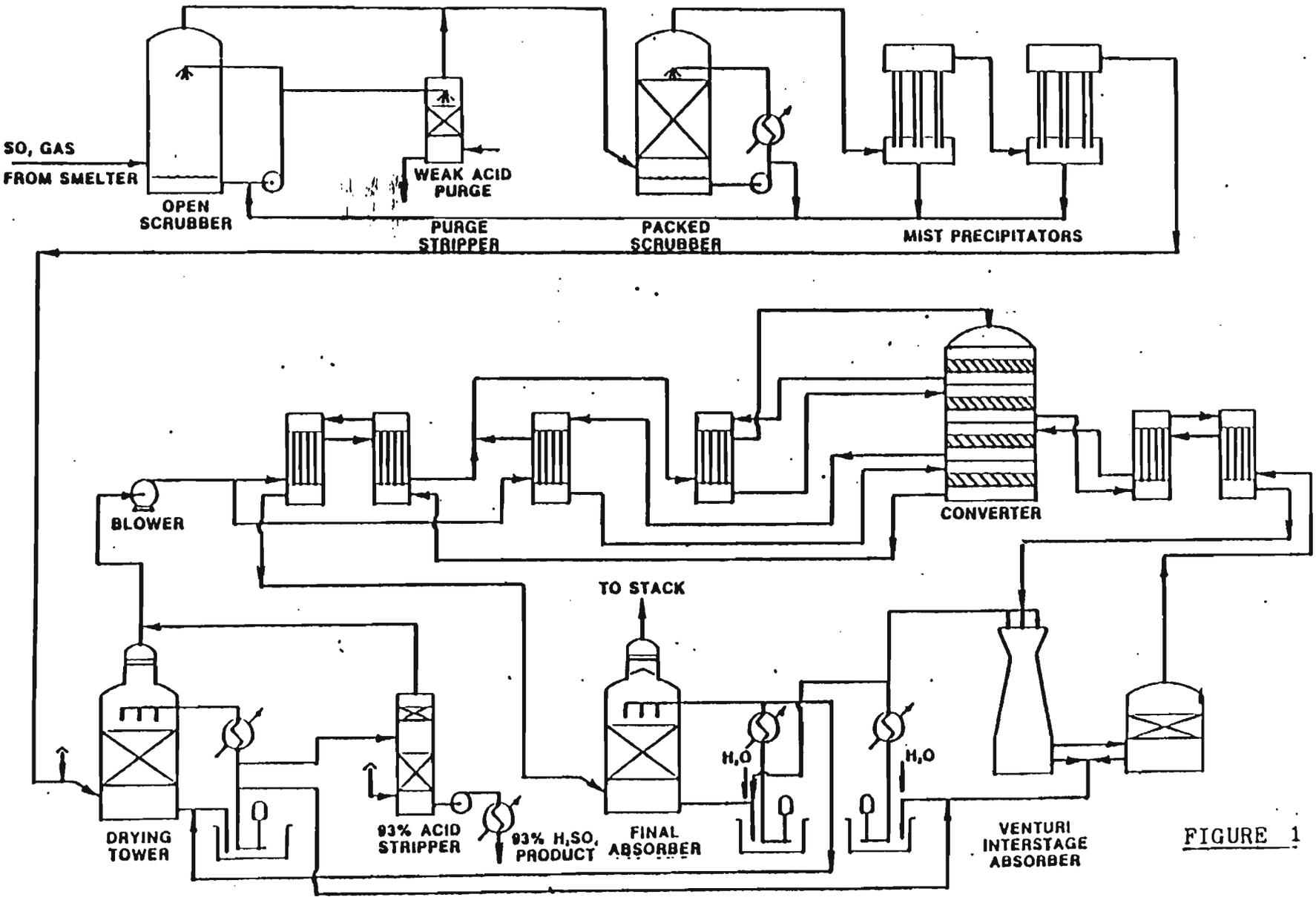


FIGURE 1

GAS COOLER TEMP VS. % SO2

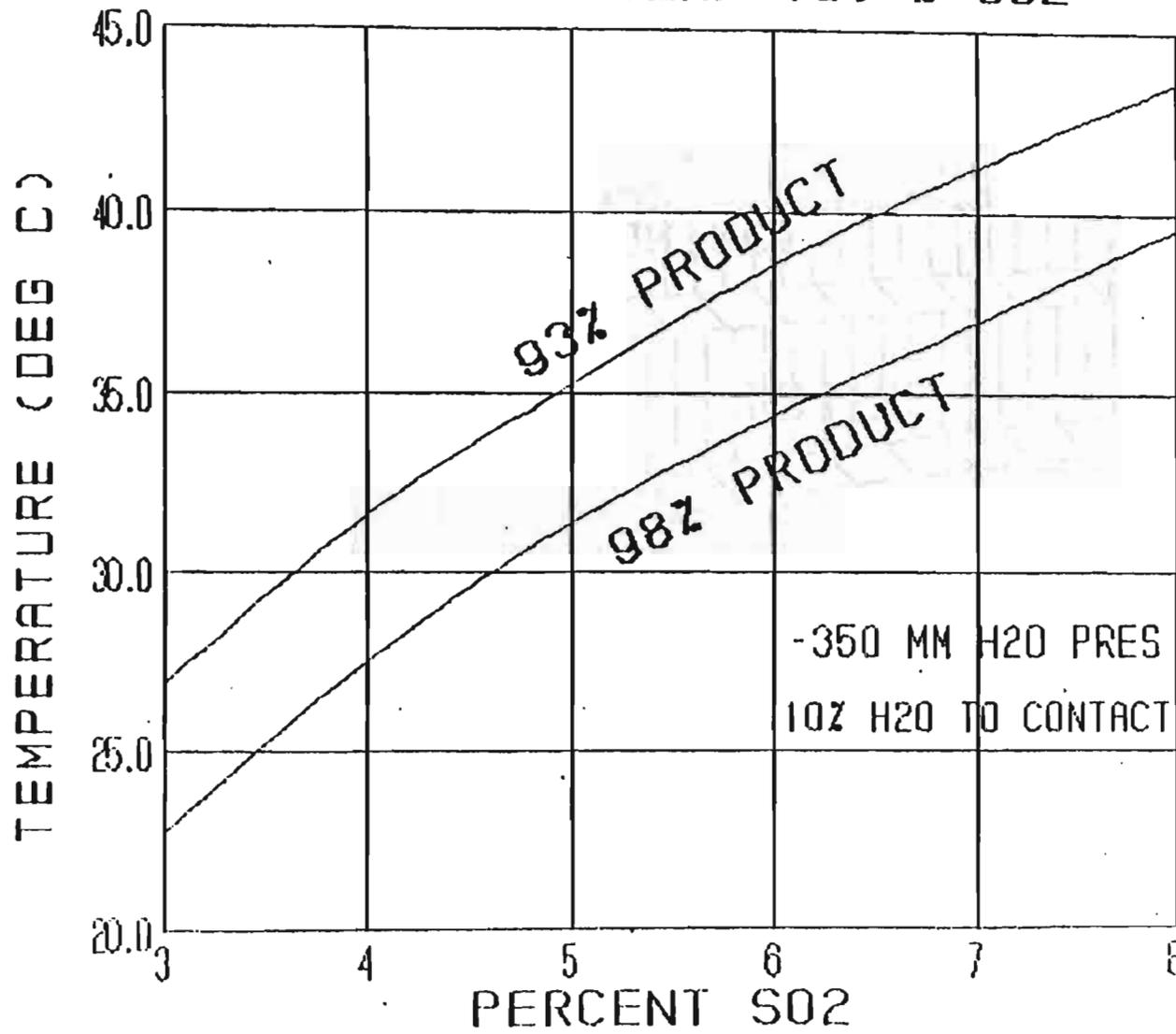


FIGURE 2

CAPITAL COST VS. SO2

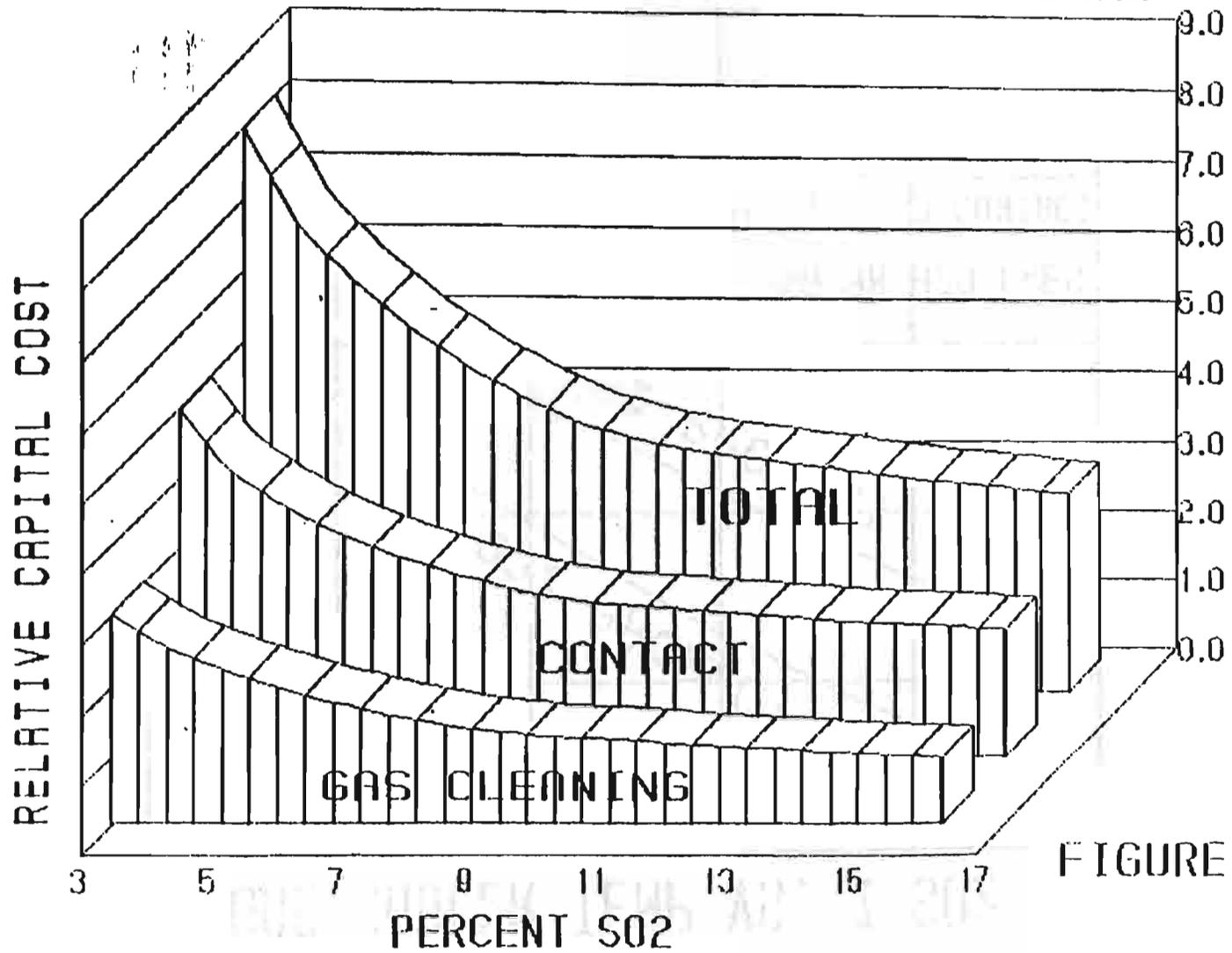


FIGURE 3

TYPICAL COPPER CONVERTER GAS

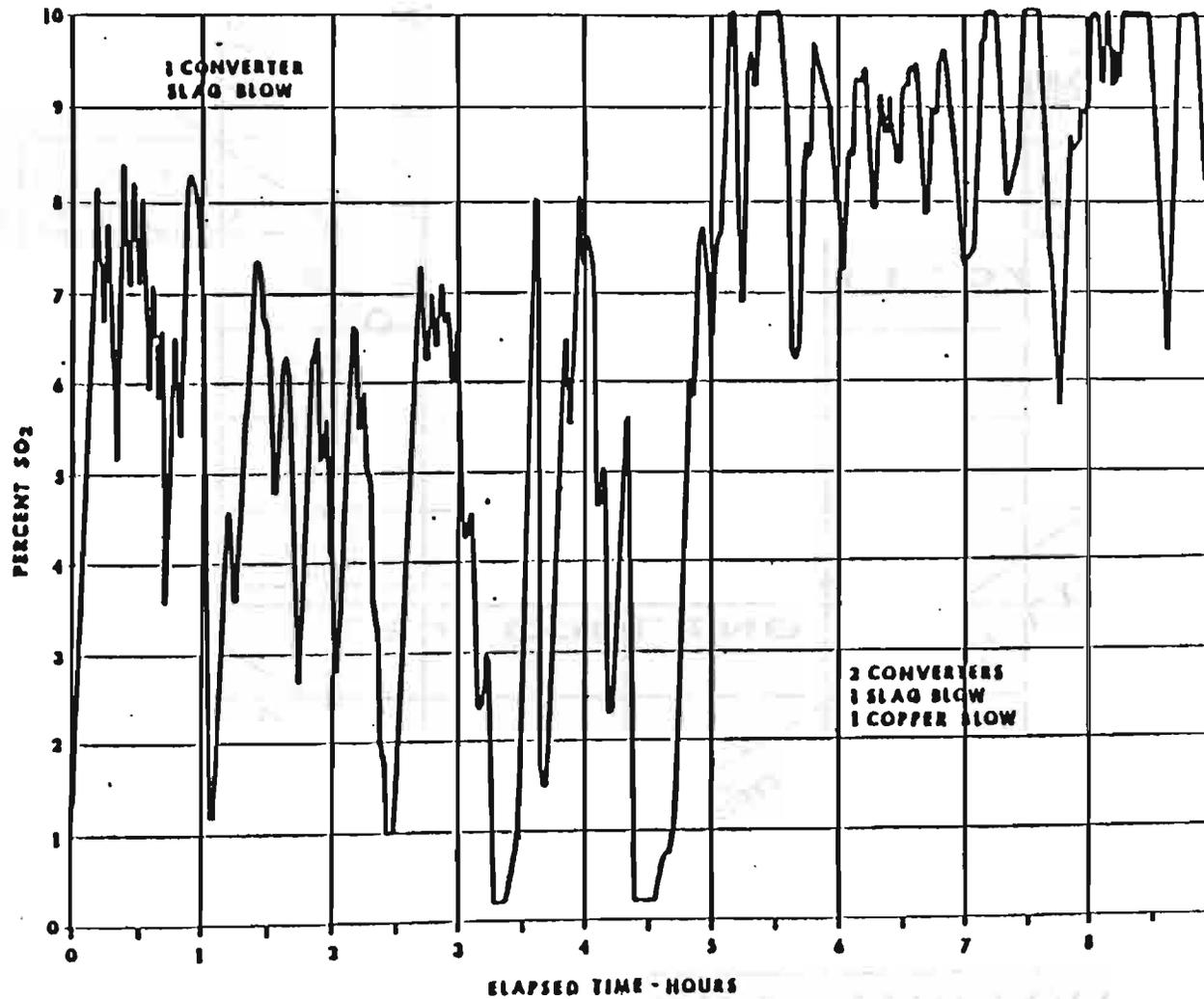
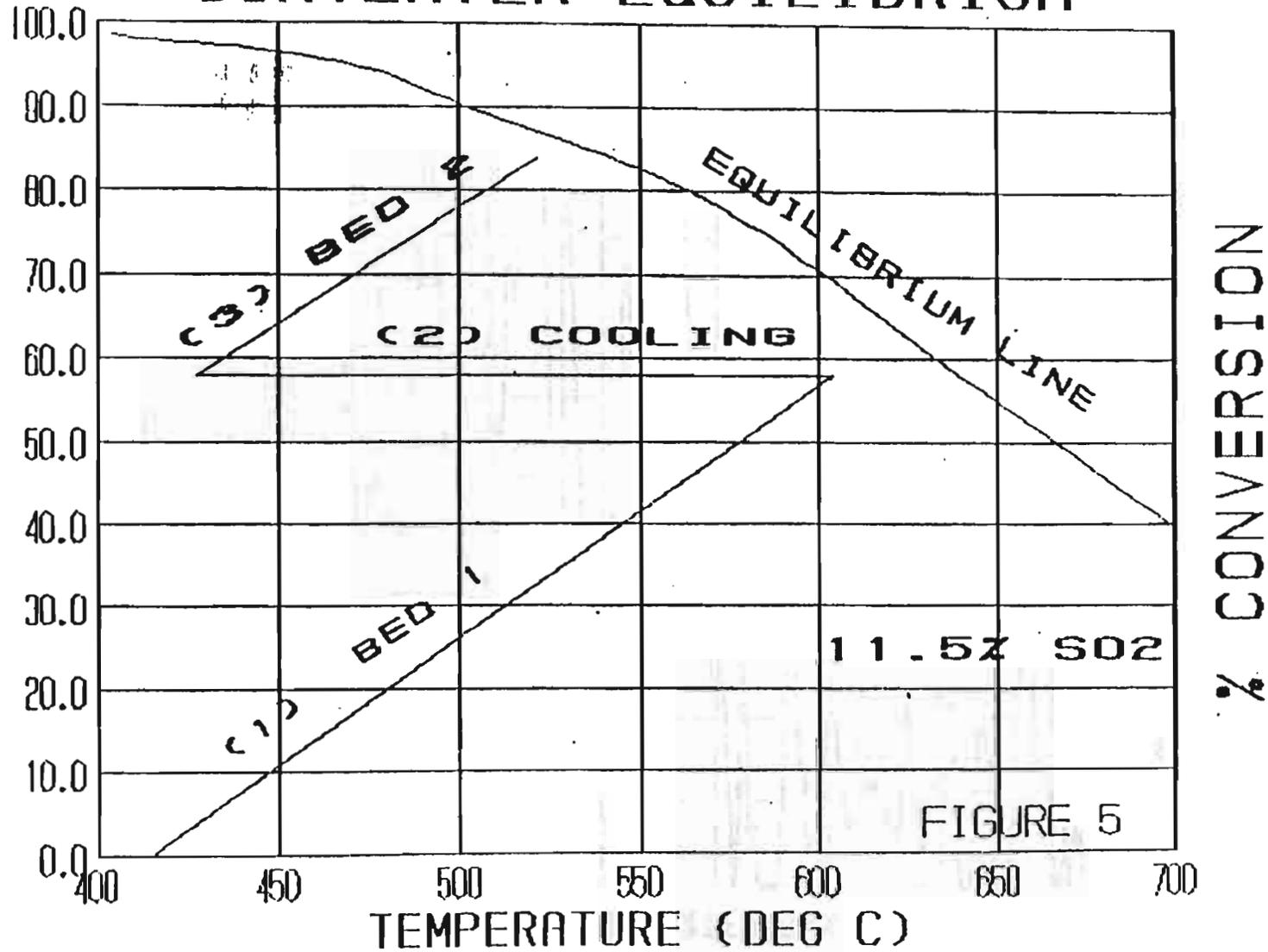
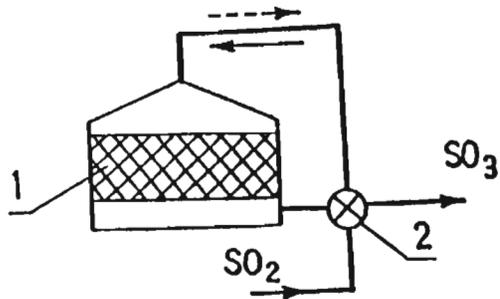


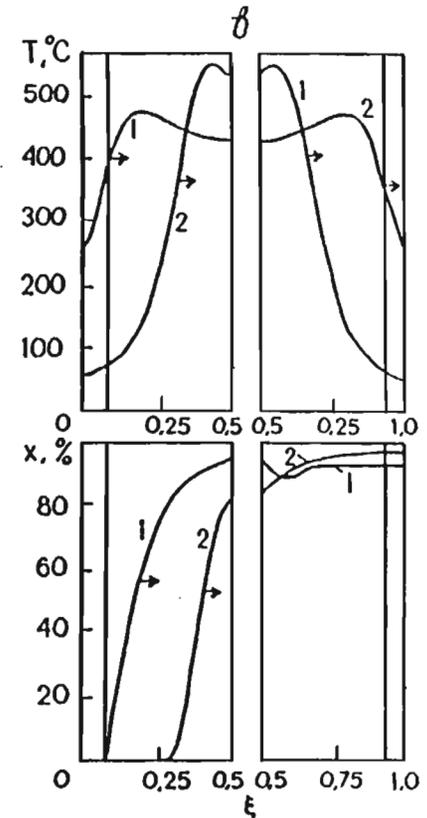
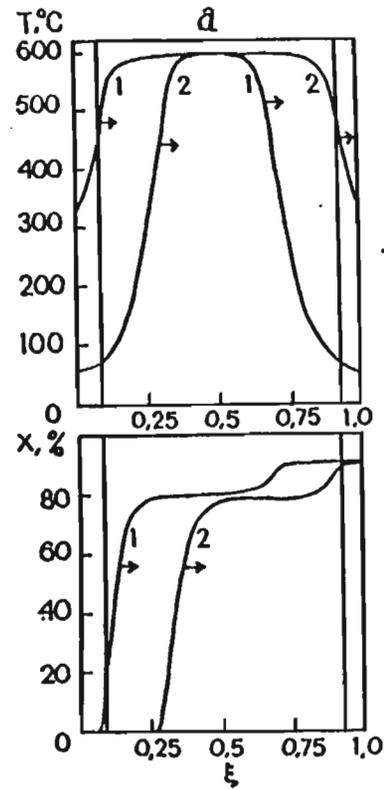
FIGURE 4

CONVERTER EQUILIBRIUM





Principal flowsheet of the reactor operated in conditions of a periodic reaction mixture flow reversal within the catalyst bed (1) with the aid of valve (2)



Temperature (T) and conversion (X) profiles along (dimensionless) bed length (ξ) in the beginning of a cycle (1) and immediately before reaction mixture reversal (2). Arrows indicate the direction of mixture filtration and heat wave-front motion. Vertical lines indicate the inert material-catalyst and catalyst-heat removal device interfaces. ($C_{SO_2} = 4.2\%$; $C_{O_2} = 14\%$; total length of the catalyst and inert material 4.8 m; linear velocity reduced to normal conditions 0.3 m/s; cycle duration 40 m). a) Conversion averaged over a cycle is 85%, $T_{max} = 600^\circ\text{C}$; b) system with removal of the chemical reaction heat; $X = 96.0\%$; $T_{max} = 551^\circ\text{C}$