

HOECHST AKTIENGESELLSCHAFT

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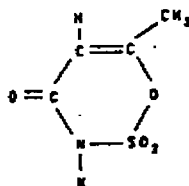
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Description

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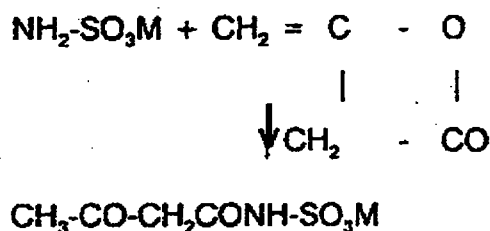
Process for the preparation of the non-toxic salts of
6-methyl-3,4-dihydro-1,2,3-oxathiazin-4-one 2,2-dioxide
5 and arrangement for carrying out this process

6-Methyl-3,4-dihydro-1,2,3-oxathiazin-4-one 2,2-dioxide
(hereafter acesulfam-H) has been used for some time in
the form of its potassium salt (hereafter acesulfam-K)



because of its intensive sweet taste as a sweetener in
10 the food sector.

A series of different processes are known for the pre-
paration of acesulfam-K (see inter alia Angewandte
Chemie, 22(1973), pages 965 to 973). Of particular
current interest is a process in which, in a solvent,
15 preferably methylene chloride, amidosulfonic acid, but
preferably a soluble salt of amidosulfonic acid, is first
reacted with diketene to form the acetoacetamido com-
pound. The salts of amidosulfonic acid used are generally
alkaline metal salts or ammonium salts, preferably
trialkylammonium salts. In the reaction with diketene,
20 the salt of acetoacetamididosulfonic acid (I) is formed in
accordance with the reaction equation



(I)

(M = base cation, in particular HN(alkyl C₁-C₆)₃)

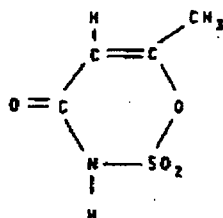
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10 $\text{CH}_3\text{-CO-CH}_2\text{-CONH-SO}_3\text{M} + \text{SO}_3 \text{ (excess)}$

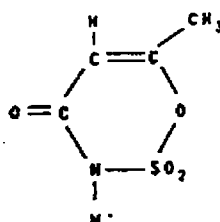
cyclization

15 | + water (excess)
hydrolysis



(II)

20. acesulfam-H
 ("sweetener acid")
 ↓ + M'OH



(III)

25 acesulfam salt (acesulfam-K)
 (M' = base cation, preferably K)

Both the cyclization and the hydrolysis are rapid, highly exothermic reactions.

5 The reaction mixture separates into an organic phase, preferably a methylene chloride phase, and an aqueous sulfuric acid phase. The majority (about 4/5) of the sweetener acid (II) is situated in the organic phase, about 1/5 is dissolved in the sulfuric acid phase. The sulfuric acid phase contains almost the entire amount of the alkali metal or ammonium, preferably trialkyl-
10 ammonium, as sulfate, preferably trialkylammonium hydrogen sulfate. The sulfuric acid phase, after stripping off any solvent still present, preferably methylene chloride, can be further utilized in another place.

15 The isolation of the sweetener acid in the form of the desired salt (III) is expediently carried out by the method described in the abovementioned EP-A 218,076. In this, the sweetener acid contained in the sulfuric acid phase is first extracted therefrom in an extraction vessel using a solvent, preferably methylene chloride.
20 The solvent used here is combined with the normal solvent phase which contains the majority of the sweetener acid. The small amount of sulfate dissolved in the solvent phase is then extracted from the combined solvent phases using only a little water in an extraction vessel and
25 returned with the extraction water to react the cyclization product with water in the hydrolysis vessel to form sweetener acid. The sweetener acid is extracted from the remaining, substantially sulfate-free solution of the sweetener acid in the solvent, in the form of the desired
30 salt, preferably the potassium salt, using aqueous alkali metal hydroxide solution, preferably potassium hydroxide solution. The aqueous solution of the sweetener (III) can be worked up by conventional methods, for example by evaporation or precipitation using external solvents, to
35 give the pure crystalline sweetener (III).





The remaining solvent is water-saturated and contains a large part of the by-product formed in the reaction, such as acetone, trialkylamine and trialkylammonium salt of the sweetener acid when the trialkylammonium salt of
5 amidosulfonic acid was used, undefined high boilers and dissolved resins.

The solvent thus contaminated cannot be reused in this form. Thus in the mixing of the SO_2 in the cyclization stage with the contaminants, dark-colored solid insoluble
10 compounds are formed which lead to blockages in the piping. In addition, with acetone, colored secondary products occur which pass through to the sweetener solution. A colorless sweetener can no longer be isolated from such solutions. Neither can this contaminated
15 solvent be used in the preparation of the acetoacetamido compound (I) from the salt of amidosulfonic acid and diketene, since the yield of acetoacetamido compound decreases and highly colored by-products are formed. These are carried over into the sweetener solution
20 obtained later and prevent the isolation of colorless solid acesulfam salt (III) from the solution.

The solvent used for the mixing with SO_2 and for the preparation of the acetoacetamido compound must therefore contain only small amounts of contaminants. The recovery
25 of pure solvent from the crude solvent which is produced downstream of the neutralization of the sweetener acid in the form of its salt requires complex apparatus with a considerable energy consumption, since, for reasons of yield and the limited solubility of the sweetener acid in
30 the solvent, only highly dilute solutions can be employed in the synthesis. The concentration of the sweetener acid, for example, in the methylene chloride solution which then passes on to the neutralization with the alkali metal hydroxide solution is generally only about
35 2 to 3% by weight. Therefore, in the purification by distillation of the crude solvent to recover pure



reusable solvent, with consideration of the reflux ratios required, up to 100 kilograms of solvent must be evaporated per kilogram of prepared sweetener.

5 In the methods conventional hitherto, the purification of the solvent is carried out in more than two distillation columns. The bottom product of the last column, which contains trialkylamine, higher boiling components such as acetone, undefined high boilers, solid residues such as
10 acid and still about 70% solvent as solubilizer for the solid residues, must be incinerated. A further evaporation, ie. a continuing recovery of solvent, is not possible since otherwise encrustations of solids occur which lead to mechanical faults.

15 In view of the disadvantages of this known process, the object was therefore in particular to provide a method for the preparation of acesulfam salts which permits a simpler recovery of the solvent used from the crude solvent with the use of at most two columns and at least
20 substantially avoids the incineration of the resulting by-product and residues.

A process has now been found by which the discarding of solvent-containing bottom streams is avoided and which additionally permits carrying out the solvent recovery to
25 give pure solvent, suitable for the preparation of the acetoacetamido compound (I) and for the preparation of the SO_3 mixture, in, optionally, one or at most two distillation columns, dispensing with a special distillation column for the removal of high boilers and residues.
30

The present invention therefore relates to a process for the preparation of acesulfam salts by reaction of salts of amidosulfonic acid with diketene to give a salt of acetoacetamidodisulfonic acid (I), ring closure by the
35 action of at least about an equimolar amount of SO_3 , at



least this ring closure reaction being carried out in the presence of a halogenated, aliphatic hydrocarbon as an inert solvent, treatment of the cyclization product (of the organic phase) with water and conversion of the resulting acesulfam-H (II) into the form of a non-toxic salt (III), which comprises, in the work-up by distillation of the resulting crude solvent, after removal of water and low-boilers and recovery of solvent sufficiently pure for reuse in the preparation of compounds (I) and/or (II), directly returning the remaining, solvent-containing distillation residue, without further purification, into the system downstream of the reaction vessel for carrying out the ring closure reaction.

This solvent-containing distillation residue is preferably returned to the vessel for treating the organic phase with water (hydrolysis vessel (5)), into the phase separation vessel (11) and/or into the extraction vessel (13); see Figure 1 below for this.

The preparation of the acesulfam salt, preferably acesulfam-K (III), is carried out in a manner known per se as described in the abovementioned EP-A 155,634, EP-A 159,516, EP-A 217,024 and EP-A 218,076, which are referred to here. Accordingly, the acetoacetamido compound (I) is first prepared by reaction of, preferably, a salt of amidosulfonic acid, in particular the trialkylammonium salt, at temperatures of generally 0 to 100°C, preferably 20 to 45°C, with diketene in an inert solvent, preferably likewise a halogenated, aliphatic hydrocarbon, in particular methylene chloride. The ring closure reaction is then carried out using SO₃, which is preferably used in a more than equimolar amount, to give an SO₃ adduct as an unisolated intermediate, the solvent used being preferably the same as that in the first stage, in particular methylene chloride. The temperatures here are generally between -20 and 40°C, preferably -10 and 15°C. This SO₃ adduct is then hydrolyzed to give the

sweetener acid (II), the temperatures here being generally 20 to 50°C, preferably 35 to 45°C. This sweetener acid is then converted by a suitable alkali metal hydroxide solution, preferably potassium hydroxide solution, in a neutralization stage into a non-toxic salt (III), preferably the potassium salt.

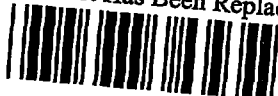
The halogenated, aliphatic hydrocarbons used as solvent are preferably those having up to 4 carbon atoms which form an azeotrope with water, such as methylene chloride, chloroform, 1,2-dichloroethane, trichloroethylene, tetrachloroethylene, trichlorofluoroethylene etc., methylene chloride being particularly preferred.

In a preferred embodiment of the invention, the same solvent, preferably methylene chloride, is used both for the salt of acetoamido-N-sulfonic acid (I) and for SO₃.

Although the molar ratio of acetoacetamido-N-sulfonate (I) to SO₃ can be about 1:1, a one to about 20-fold SO₃ excess is preferred, preferably about a 3- to 10-fold, in particular about a 4- to 7-fold molar excess.

In the preferred case of the use of salts of acetoacetamido-N-sulfonic acid (I) and SO₃ in a molar ratio of 1 to more than 1, an "SO₃ adduct" is formed in the ring closure reaction, from which the acesulfam-H (II) must be set free by hydrolysis. This hydrolysis is carried out by addition of water or ice and/or addition of dilute aqueous sulfuric acid, expediently in about a 2- to 6-fold molar amount, in relation to the SO₃ excess used. For this hydrolysis, both water and dilute aqueous sulfuric acid are preferably used, which dilute aqueous sulfuric acid originates from the downstream extractor (18) and generally has a concentration of 2 to 20% by weight.





After the hydrolysis, a 2- or (if acesulfam-H has already precipitated) 3-phase mixture is then present. The acesulfam-H (II) is present essentially dissolved in the organic phase and in the sulfuric acid phase. The organic
5 phase is then removed. The organic phase removed from the aqueous sulfuric acid phase, or the corresponding combined organic phases, are, preferably, then purified by extraction with water in the extractor (18) mentioned above, the sulfuric acid contained therein being first
10 extracted from the organic phase.

The volume ratio of the organic phase to the aqueous extraction phase is generally about (20-5):1. However, even with substantially smaller amounts of water, an effective purification can still be achieved.

15 The extraction is carried out in the simplest case by stirring the two phases in a stirred flask or in a stirred tank; special apparatuses which are useful are, in principle, all industrial extraction apparatuses such as mixer-settlers, sieve-plate columns, packed columns,
20 Karr columns etc. Mixing elements such as static mixers can also be used to intensify the contact of the extraction phases.

The proportion of acesulfam-H extracted simultaneously with the sulfuric acid is generally, depending on the
25 water amount used, between about 2 and 30% of the total acesulfam-H contained in the organic phase. For the economic efficiency of the overall process, it is important to return the water phase to the hydrolysis of the "SO₃ adduct".

30 The non-toxic salts (III) of the acesulfam-H are recovered from the purified organic phase or the purified combined organic phases by neutralization with bases. Bases which are useful here are those having non-toxic cations. Preference is given to potassium bases

(solutions of KOH, KHCO_3 , K_2CO_3), in particular KOH.

The neutralization of acesulfam-H and the recovery of non-toxic salts thereof from this acesulfam-H-containing purified organic phase is advantageously carried out by
5 intensive contact of the purified organic phase, or of the corresponding combined organic phases, with aqueous alkali metal hydroxide solution. The intensive contact is generally carried out in the manner of an extraction by the processes conventional therefor in the conventional
10 apparatuses as have already been described above. Mixing elements, such as static mixers, can also be used here.

In the neutralization, sufficient base is generally used so that the aqueous phase achieves a pH of about 5 to 12, preferably about 8 to 11. The acesulfam salt is then
15 recovered from the aqueous phase in a conventional manner (by crystallization).

According to a preferred embodiment, the work-up of the solvent, preferably methylene chloride, by distillation is carried out in one distillation column only. According
20 to a further preferred embodiment, the solvent-containing distillation residue is returned to the system at more than one point, in particular to the phase separation vessel (11) and/or the extraction vessel (13).

A further preferred embodiment provides that
25 the salt of acetoacetamidossulfonic acid (I) used is trialkylammonium acetoacetamidossulfonate, dissolved in methylene chloride,
the ring closure is carried out by the action of a more than equimolar amount of SO_3 , if desired, likewise in
30 methylene chloride as solvent,
the undefined cyclization product produced as an SO_3 adduct after the ring closure reaction is hydrolyzed in a hydrolysis vessel to form acesulfam-H,



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a separation is carried out in downstream phase separation vessel into an organic phase and aqueous sulfuric acid-containing phase, this aqueous phase is then extracted in an extraction vessel with methylene chloride and this methylene chloride, together with the organic phase from the phase separation vessel is fed to a further extractor for washing out sulfuric acid residues and sulfate residues using water, the non-toxic salts of acesulfam-H are recovered, by neutralization using bases, from the organic phase thus purified and a separation into aqueous salt solution and a methylene chloride phase is carried out in a downstream phase separation vessel, and this methylene chloride phase is fed, if desired via at least one further extraction stage, to the distillation column.

The work-up and recovery of the solvent (methylene chloride) is preferably carried out in the context of the process according to the invention in such a manner that the crude solvent is introduced into a distillation column in the bottom thereof or above the bottom thereof, the distillation residue, which in addition to parts of the solvent, contains solid by-products and high-boilers, is withdrawn from this distillation column and returned to the system, preferably to the hydrolysis vessel (5), the phase separation vessel (11) and/or the extraction vessel (13) for extraction of the sulfuric acid phase, water, small amounts of the solvent and any low-boilers which may be present with this are withdrawn at the head of this distillation column, while the majority of the solvent is returned as reflux to the distillation column, and sufficiently purified solvent is withdrawn from the central part of the column from an intermediate plate which is located beneath the reflux and above the feed of crude solvent into the column and this purified solvent is returned into the system, preferably into the reaction

vessel for the preparation of compounds (I) and/or (II) and for the preparation of the SO_2 solution. Obviously, part of this purified solvent, which generally contains at most 2% by weight, preferably at most 1% by weight, and, in particular, only 0.9 to 0.1% by weight of contaminants, can also be used in the following points of the system.

The process according to the invention offers surprising advantages. Thus, the complexity of the apparatus is significantly less, since only one column or at most two columns are required instead of three columns. In addition, no solvent-containing residues are produced, the disposal (incineration) of which would require additional special devices. Furthermore, the energy consumption for purification by distillation of the crude solvent is lower.

The practicability of the process according to the invention was surprising, since on return of the by-products, such as acetone and polymeric solids, into the system, a concentration in the total circulation was to be expected which would finally pass through to the end product (acesulfam salt (K) solution). Contaminants in the acesulfam salt (K) solution, which is further worked up to give pure solid acesulfam salt (K), make the recovery thereof in pure form more difficult or impossible if highly colored polymers are present in the solution.

Furthermore, it was to be expected that the solvent withdrawn from the intermediate plate in the distillation column described would contain significantly more water than was actually found; however, the water content in the solvent discharged from the intermediate plate was surprisingly significantly lower than in the bottom and the head of the column.





The present invention further relates to an arrangement for carrying out the process according to the invention, essentially comprising the reaction vessel (3) for carrying out the cyclization reaction from the compounds (I) and SO_3 , a downstream hydrolysis vessel (5) for the recovery of the sweetener acid (II), a phase separation vessel (11) connected thereto for the separation of the organic phase and the sulfuric acid phase, preferably an extractor (13) for the extraction of the isolated sulfuric acid phase using the solvent, preferably a further extractor (18) for the extraction of the organic phases combined from (11) and (13) using water, a neutralization vessel (22) for the conversion of the sweetener acid (II) into the corresponding salt (III), a phase separation vessel (25) for the separation of the aqueous solution of the salt (III) from the solvent and a distillation column (32).

A phase separation vessel (36) is preferably connected to the distillation column (32), into which phase separation vessel the vapor condensate passes over from the column head and is here separated into an aqueous phase and a solvent phase. The latter is returned into the column. Pure solvent is withdrawn from an intermediate plate of the column, which generally has 15 to 50, preferably 20 to 30, theoretical plates. The solvent-containing distillation residue is preferably returned from the column bottom into the hydrolysis vessel (5), into the phase separation vessel (11) and/or into the extraction vessel (13). This distillation residue generally still contains at least 10% by weight, preferably 50 to 98% by weight and, in particular, 75 to 95% by weight of solvent.

The process according to the invention is now described by way of example with reference to Figure 1:

Via line (1), a solution of SO_3 in methylene chloride is fed to the reaction vessel (mixer) (3); the feed of



acetoacetamido trialkylammonium salt solution in methylene chloride is carried out via line (2) into the reaction vessel (3). Additional methylene chloride is added to this solution via line (2') for dilution. The heat of reaction is removed via evaporation of part of the methylene chloride, evaporating methylene chloride and liquid product passing together via line (4) to the vessel (5) furnished with the stirrer. Water is fed to the vessel (5) via the line (6); the heat of reaction of the hydrolysis (hydration) step is also removed via methylene chloride evaporation. The methylene chloride vapors pass via line (7) to the condenser (8); the condensate returns via line (9) to the vessel (5). The liquid mixture in vessel (5) passes via line (10) to the phase separation vessel (11). The lower sulfuric acid phase of the vessel (11) is passed via line (12) to the extractor (mixer-settler vessel) (13), in which this sulfuric acid phase is stirred with methylene chloride from line (14) for the extraction of the sweetener acid. The sulfuric acid phase is withdrawn from the extractor (13) via line (15). The methylene chloride loaded with sweetener acid leaves vessel (13) via line (16). The methylene chloride phase from the phase separation vessel (11) is withdrawn via line (17) and is transported with the methylene chloride from line (16) to the extractor (18). Water is fed to the extractor (18) for the extraction of dissolved sulfate below via line (19), which water, after leaving the extractor (18), is conducted via line (20) to the hydrolysis vessel (5). The methylene chloride phase of the extractor (18) is conducted via line (21) to the stirred neutralization vessel (22), to which aqueous potassium hydroxide solution is passed via line (23), the rate of which is controlled via the pH in the stirred vessel (22). The mixture is conducted via line (24) to the phase separation vessel (25). The upper phase is aqueous acesulfam-K solution, which is withdrawn via line (26). The lower methylene chloride phase is conducted via line (27) to the mixer-settler vessel (28),

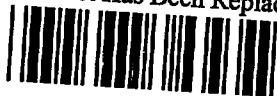


in which it is washed with a little water which is added to the mixer-settler (28) via line (29). The washing water is conducted via line (30) together with the aqueous potassium hydroxide solution to the neutralization vessel (22). The methylene chloride phase withdrawn from the mixer-settler (28) is crude methylene chloride which is passed via line (31) into the bottom of the distillation column (32). The vapors from the column (32) pass via line (33) to the condenser (34) and the two-phase condensate passes via the line (35) to the phase separation vessel (36). The aqueous phase separating in the phase separation vessel (36) is withdrawn via line (37); the methylene chloride phase is returned via line (38) as reflux into column (32). The by-product-containing methylene chloride is withdrawn at the bottom of column (32) via line (14) for the extraction of the sweetener acid from sulfuric acid in extractor (13). Pure methylene chloride is taken off from an intermediate plate situated in the column via line (39), which pure methylene chloride is reused for the SO_2 mixture and/or to prepare the trialkylammonium acetoacetamidodisulfonate solution.

Comparison Example 1

Only pure fresh methylene chloride is employed without reprocessing. The rates given are per hour.

A solution of 11.0 kg of SO_2 in 46.0 kg of methylene chloride at a temperature of $+5^\circ\text{C}$ is fed via line (1) to the reaction vessel (3). 15.0 kg of methylene chloride solution at 0°C , which contains 6.6 kg of triethylammonium acetoacetamidodisulfonate and 1.0 kg of polymeric components of different composition and unreacted starting material in a small amount are introduced into the reaction vessel (3) via line (2). This solution is diluted with 12.0 kg of methylene chloride via line (2'). The preparation of the solution is carried out in a



stainless steel stirred tank by the introduction of 10.0 kg of methylene chloride, addition of 2.5 kg of amidosulfonic acid, 2.8 kg of triethylamine, 0.1 kg of acetic acid and then careful addition of 2.2 kg of diketene after the above components have completely dissolved. The mixture formed in the reaction vessel, which contains a not precisely definable cyclization product as a solution in methylene chloride, by-products formed and methylene chloride evaporated by the heat of reaction, is fed via line (4) to the so-called hydrolysis vessel (5) (stirred tank).

8.0 kg of water are added via line (6) to the hydrolysis tank (5). 20.0 kg of the methylene chloride evaporated by the heat of reaction pass via line (7) to the condenser (8), while the condensate returns via line (9) to the stirred tank (5). The liquid mixture in hydrolysis tank (5), which is held in a constant state, is conducted via line (10) to the phase separator (11). The lower sulfuric acid phase flows via line (12) to the mixer-settler (13). 30.0 kg of pure fresh methylene chloride are conveyed via line (14) to the mixer-settler (13). 26.0 kg of sulfuric acid phase are withdrawn via line (15). This sulfuric acid phase contains 41% sulfuric acid, 31% water, 21% triethylammonium hydrogen sulfate, 0.5% acetic acid, 1% acetone, 0.1% sweetener acid, 3% undefined organic compounds and 1% dissolved methylene chloride. Via line (16), the methylene chloride phase is passed from the mixer-settler (13) into line (17). 100.5 kg of methylene chloride phase are withdrawn from the phase separator (11) via line (17) which contain, in addition to other components, 0.2% water, 2.4% sweetener acid and 0.4% sulfate. The combined methylene chloride streams from the lines (16) and (17) are fed to the extractor (18). 2.7 kg of water are introduced at the bottom of the extractor via line (14); the extraction water is returned via line (20) to the hydrolysis tank (5). The methylene chloride phase of the extractor (18) is fed via line (21) to the



neutralization vessel (22) (stirred tank, water cooling in outer jacket); approximately 10% strength potassium hydroxide solution is transported via line (23) into the tank (22) with pH control to pH 9 in the tank (22). The
5 resulting two-phase mixture is conducted via line (24) to the phase separator (25). The lower methylene chloride phase is fed from the phase separator (25) via line (27) to the mixer-settler (28) in which this methylene chloride phase is washed using 6.4 l of water via line
10 (29). The washing water passes, together with the potassium hydroxide solution, via line (30) to the neutralization vessel (22).

99.0 kg of methylene chloride are withdrawn via line (31). The methylene chloride contains 0.15% water, 0.1%
15 triethylamine, 0.4% acetone and 0.5% polymers.

17.6 kg of acesulfam-K solution are withdrawn as the upper phase of the phase separator (25) via line (26). The light-yellow solution contains 1% methylene chloride, 17% acesulfam-K, 0.3% potassium sulfate and 0.5% acetone,
20 0.1% triethylamine and <0.5% unknown compounds.

On evaporation of this sweetener solution to approximately 30% of the original volume, cooling and filtration, colorless acesulfam-K is obtained as a solid.

Comparison Example 2

25 Feed of crude methylene chloride to the reaction by dilution of the triethylammonium acetoacetamido sulfonate solution with crude instead of pure methylene chloride.

The experimental arrangement and rates are as in Comparison Example 1, but to dilute the acetoacetamido
30 compound in line (2), instead of pure methylene chloride, dilution is carried out via line (2') using 12.0 kg of crude methylene chloride from line (31) in Comparison



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Example 1.

On evaporation of the acesulfam-K solution obtained via line (26), which solution is now dark-colored, in correspondence with Comparison Example 1, yellow acesulfam-K is
5 obtained as a solid which can only be obtained in a colorless form via redissolution and further treatment of the solution using adsorbents.

Example 3

The experimental arrangement is as in the comparison
10 examples. However, the crude methylene chloride from the mixer-settler (28) is now conducted via line (31) into the bottom of the distillation column (32) for reprocessing.

99.0 kg of crude methylene chloride are introduced via
15 line (31) into the bottom of the column (32) (45 sieve plates, intermediate plate for liquid withdrawal at plate 30).

The crude methylene chloride corresponds in the composition to the crude methylene chloride in Comparison
20 Example 1; the other individual streams also initially correspond to the individual streams of Comparison Example 1.

The bottom heating of column (32) is regulated so that
25 110.0 kg of distillate are produced via vapor line (32) in the condenser (34) (cooling: cold water), which distillate is conducted via line (35) to the phase separator (36). 0.13 kg of water phase are withdrawn via line (37). The lower organic methylene chloride phase is returned via line (38) as reflux into the column. 66.0 kg of pure
30 methylene chloride are withdrawn at the intermediate plate via line (39). This methylene chloride contains, as contaminant, 140 ppm of water, 40 ppm of acetone,



<30 ppm of triethylamine; it is reused as solvent for the preparation of the acetoacetamido compound and for the dilution of the solution prior to the reaction and for the preparation of the SO₂ solution.

- 5 29.7 kg of material are withdrawn from the bottom of column (32) via line (14) which are used for the extraction of the sweetener acid from the sulfuric acid phase in the mixer-settler (13) and are returned via line (16) to the methylene chloride phase via line (17).
- 10 The bottom material from the column (32) in line (14) is essentially composed of methylene chloride which contains 300 ppm of water, 1.4% of acetone, 0.3% of triethylamine and 1.7% of polymers.

- 15 After an experimental duration of several days, when all individual streams are in concentration equilibrium, the bottom stream (14) shows a slight increase in acetone concentration to 2%, but a concentration of the polymers or other components does not occur. The sulfuric acid withdrawn contains 2% acetone, 0.5% condensation products
- 20 of the acetone, 5% polymers and undefined organic compounds.

- 25 17.6 kg of acesulfam-K solution are withdrawn via line (26); the light-yellow solution contains 17% acesulfam-K, 1% methylene chloride, 0.3% potassium sulfate, 0.6% acetone, 0.1% triethylamine, <0.5% unknown compounds. On evaporation of this solution to 30% of the original volume, cooling and filtration, colorless acesulfam-K is obtained as a solid.

- 30 In principle, an identical result is obtained if the bottom product of column (32) is fed to the hydrolysis vessel (5). In this case, for example, head product of column (32) can also be used for the extraction of the sweetener acid from the sulfuric acid phase in vessel

(13). The phase separator (32) is not necessarily required in this variant.

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PATENT CLAIMS

1. A process for the preparation of acesulfam salts by reaction of salts of amidosulfonic acid with di-
ketene to give a salt of acetoacetamidossulfonic
acid, ring closure by the action of at least about
an equivalent amount of SO_3 , at least this ring
closure reaction being carried out in the presence
of a halogenated, aliphatic hydrocarbon as an inert
solvent, treatment of the cyclization product with
water and conversion of the resulting acesulfam-H
into the form of a non-toxic salt, which comprises,
in the work-up by distillation of the resulting
crude solvent, after removal of water and low-
boilers and recovery of solvent sufficiently pure
for reuse in the preparation of compounds (I) and/or
(II), directly returning the remaining, solvent-
containing distillation residue, without further
purification, into the system downstream of the
reaction vessel for carrying out the ring closure
reaction.
2. The process as claimed in claim 1, wherein the
solvent used is methylene chloride.
3. The process as claimed in claim 1 or 2, wherein the
work-up by distillation is carried out in only one
distillation column.
4. The process as claimed in at least one of claims 1
to 3, wherein the solvent-containing distillation
residue is returned at least at one point of the
system, preferably into the vessel for treating the
cyclization product with water, the phase separation
vessel downstream of this and/or the extraction
vessel following this.



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5. The process as claimed in at least one of claims 1 to 4, wherein the salt of acetoacetamidossulfonic acid (I) used is trialkylammonium acetoacetamidossulfonate,
- 5 the ring closure is carried out by the action of a more than equimolar amount of SO_3 , if desired, preferably in methylene chloride as solvent, the SO_3 adduct produced after the ring closure reaction is hydrolyzed in a hydrolysis vessel to
- 10 give acesulfam-H, a separation is carried out in downstream phase separation vessel into an organic phase and aqueous sulfuric acid-containing phase, this aqueous phase is then extracted in an extraction vessel with
- 15 methylene chloride and this methylene chloride, together with the organic phase from the phase separation vessel is fed to a further extractor for washing with water, the non-toxic salts of acesulfam-H are recovered, by
- 20 neutralization using bases, from the organic phase thus purified and a separation into aqueous salt solution and a methylene chloride phase is carried out in a downstream phase separation vessel, and this methylene chloride phase is fed, if desired via
- 25 at least one further extraction stage, to the distillation column.
6. The process as claimed in at least one of claims 1 to 5, wherein the crude solvent is introduced into a distillation column in the bottom thereof or above
- 30 the bottom thereof, the distillation residue, which in addition to parts of the solvent, contains solid by-products and high-boilers, is withdrawn from this distillation column and returned into the vessel for the treatment of
- 35 the cyclization product, the phase separation vessel downstream of this and/or the extraction vessel following this for extraction of the sulfuric acid





phase,

water, small amounts of the solvent and any low-boilers which may be present with this are withdrawn at the head of this distillation column, while the majority of the solvent is returned as reflux to the distillation column, and

sufficiently purified solvent is withdrawn from the central part of the column from an intermediate plate which is located beneath the reflux and above the feed of crude solvent into the column and this purified solvent is returned into the system.

7. The process as claimed in claim 6, wherein the purified solvent is returned into the reaction vessel.

8. The process as claimed in at least one of claims 1 to 7, wherein the solvent-containing distillation residue is returned into the hydrolysis vessel.

9. The process as claimed in at least one of claims 1 to 8, wherein the solvent-containing distillation residue is returned into the extractor, provided downstream of the hydrolysis vessel, for the extraction of the acesulfam-H compound still present in the aqueous phase.

10. The process as claimed in at least one of claims 1 to 9, wherein the distillation residue from the distillation column still contains at least 10% by weight, preferably up to 98% by weight of solvent.

11. An arrangement for carrying out the process according to the invention, essentially comprising the reaction vessel (3) for carrying out the cyclization reaction from a salt of acetoacetamididosulfonic acid (I) and SO_3 , a downstream hydrolysis vessel (5) for the recovery of the sweetener acid (II), a phase



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5 separation vessel (11) connected thereto for the separation of the organic phase and the sulfuric acid phase, a neutralization vessel (22) for the conversion of the sweetener acid (II) into the corresponding salt (III), a phase separation vessel (25) for the separation of the aqueous solution of the salt (III) from the solvent and a distillation column (32).

10 12. The arrangement as claimed in claim 11, which is additionally provided with an extractor (13) for the extraction of the separated sulfuric acid phase using the solvent and a further extractor (18) for the extraction of the organic phases combined from (11) and (13) using water.

15 13. The arrangement as claimed in claim 11 or 12, wherein a phase separation vessel (36) is connected to the distillation column (32), into which phase separation vessel the condensed vapors pass from the column head and in which a separation is carried out
20 into an aqueous phase and a solvent phase.

14. The arrangement as claimed in at least one of claims 11 to 13, wherein the distillation column (32) has 15 to 50 theoretical plates.



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